NATIONAL ENERGY TECHNOLOGY LABORATORY

Carbon Capture Approaches for Natural Gas Combined Cycle Systems

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CARBON CAPTURE AND SEQUESTRATION APPROACHES FOR NATURAL GAS COMBINED CYCLE SYSTEMS

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FINAL REPORT

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List of Acronyms and Abbreviations

$/MMkJ	Dollars per million kilojoules (also shown as 10^6kJ)
$/MMBtu	Dollars per million British thermal units (also shown as 10^6 Btu)
$MM/yr	Million dollars per year
$/ton	Dollars per ton
$/MWhnet	Dollars per megawatt hour net
°C	Degrees Centigrade
°F	Degrees Fahrenheit
AACE	Association for the Advancement of Cost Engineering
abs	Absolute
AEO	Annual Energy Outlook
AFUDC	Allowance for funds used during construction
AGR	Acid gas removal
Ar	Argon
ASME	American Society of Mechanical Engineers
ASU	Air separation unit
ATR	Auto Thermal Reformer/Auto Thermal Reforming
barg	bar gauge
BB	Bituminous Baseline
BEC	Bare erected cost
BFD	Block flow diagram
BFW	Boiler feed water
Btu	British thermal unit
Btu/hr	British thermal unit per hour
Btu/kWh	British thermal unit per kilowatt hour
Btu/lb	British thermal unit per pound
Btu/scf	British thermal unit per standard cubic foot
CCF	Capital charge factor
CCS	Carbon capture and sequestration
CDR	Carbon Dioxide Recovery
CES	Clean Energy Systems
CF	Capacity Factor
C_2H_6	Ethane
C_3H_8	Propane
C_4H_{10}	n-Butane
CH_4	Methane
CMB	Combustor
CMU	Carnegie Mellon University
CO	Carbon monoxide
CO_2	Carbon dioxide
CoP	ConocoPhillips
CT	Combustion turbine
CTG	Combustion Turbine-Generator
DEA	Diethanolamine
DLN	Dry low NOx

DOE	Department of Energy
EGR	Exhaust gas recycle
EIA	Energy Information Administration
EPC	Engineering Procurement and Construction
EPCM	Engineering/Procurement/Construction Management
EPRI	Electric Power Research Institute
ESPA	Energy Sector Planning and Analysis Services
FEED	Front-end engineering design
FGD	Flue gas desulfurization
FOAK	First-of-a-kind
ft	Foot, Feet
GE	General Electric
GJ/hr	Gigajoules per hour
gpm	Gallons per minute
gpm/MWnet	Gallons per minute per megawatt net
GT	Gas turbine
H_2O	Water
HGCU	Hot gas cleanup
HHV	Higher heating value
HP	High pressure
HRSG	Heat recovery steam generator
HTHP	High temperature, high pressure
HVAC	Heating, ventilating, and air conditioning
HVAP	Heat of vaporization
HX	Heat exchanger
IEA	International Energy Agency
IGCC	Integrated Gasification Combined Cycle
IOU	Investor owned utilities
IP	Intermediate pressure
IRROE	Internal rate of return on equity
ISO	International Standards Organization
kg/hr	Kilogram per hour
kg/MWh	Kilograms per megawatt hour
kJ	Kilojoules
kJ/hr	Kilojoules per hour
kJ/kg	Kilojoules per kilogram
kWe	Kilowatts electric
kWh	Kilowatt-hour
kW_{th}	Kilowatt-thermal
lb	Pound
lb/hr	Pounds per hour
lb/MWh	Pound per megawatt hour
LCOE	Levelized cost of electricity
LF	Levelization factor
LHV	Lower heating value
LNB	Low NOx burner

LP	Low pressure
m^3/min	Cubic meter per minute
µm	Micrometer
MAC	Multi-stage air Compressor
md	millidarcy
MFC	Multi-stage fuel compressor
MOC	Multi-stage oxygen compressor
MDEA	Methyldiethanolamine
MEA	Monoethanolamine
MMBtu	Million British thermal units (also shown as 10^6 Btu)
MMBtu/hr	Million British thermal units (also shown as 10^6 Btu) per hour
MMkJ	Million kilojoules (also shown as 10^6 kJ)
MJ/scm	Megajoules per standard cubic meter
MPa	Megapascals
MW	Megawatts
MWe	Megawatts Electric
N/A	Not applicable
N_2	Nitrogen
NETL	National Energy Technology Laboratory
NGCC	Natural gas combined cycle
Ni	Nickel
NO_x	Oxides of nitrogen
O_2	Oxygen
O&M	Operating and maintenance costs
O&GJ	Oil and gas journal
OGT	Optimized Gas Treating
P	Pressure
PC	Pulverized coal
PM	Particulate matter
PSFM	Power Systems Financial Model
ppmv	Parts per million volume
ppmvd	Parts per million volume dry
POX	Partial Oxidation
psia	Pounds per square inch absolute
psig	Pounds per square inch gage
SC	Supercritical
SCR	Selective catalytic reduction
SNG	Substitute natural gas
SO_2	Sulfur dioxide
STG	Steam turbine generator
T	Temperature
TAG	Technical Assessment Guide
TASC	Total As-Spent Cost
Ti	Titanium
TIT	Turbine inlet temperature
TOC	Total overnight cost

Ton/MWhnet Ton per megawatt hour net
Tonne Metric ton (1000 kg)
TPC Total plant cost
TS&M Transport, storage, and monitoring
U.S. United States
V-L Vapor Liquid portion of stream (excluding solids)
WGS Water gas shift
yr Year
ZnO Zinc Oxide

Executive Summary

The purpose of this study is to present the cost and performance of advanced natural gas combined cycle (NGCC) plants with CO_2 capture, using a consistent technical and economic approach that accurately reflects current market conditions for future developmental technologies.

For this study, there were three types of carbon capture approaches examined: pre-combustion, post-combustion and oxy-combustion. In pre-combustion carbon capture, the carbon in the fuel is converted to CO_2 and removed before the combustion process, whereas in post-combustion, the more dilute CO_2 is separated from the flue gas at a lower pressure. Oxy-combustion technologies use nearly pure oxygen as the oxidant so that the flue gas consists primarily of CO_2 and water vapor.

Seven different plant design configurations were analyzed as listed in Exhibit ES-1. Two cases are NGCC reference cases without carbon capture (Ref1) and with carbon capture (Ref2). These cases are taken from the Bituminous Baseline (BB) study [1]. The Ref2 case includes a modification to the amine auxiliary load calculation from the BB study NGCC with capture case. Since the comparable NGCC cases in this study also use the modified approach, case Ref2 is used as the baseline with capture reference case throughout the study.

The methodology included performing steady-state simulations of the various technologies using the Aspen Plus (Aspen) modeling program. The resulting mass and energy balance data from the Aspen models were used to size major pieces of equipment. The cost estimating methodology for Cases 1 through 5 uses reference costs established in the BB study. Performance and process limits were based upon published reports and/or best engineering judgment. Capital and operating costs for the reference cases were estimated by WorleyParsons based on simulation results and using vendor quotes/discussions, costing software, or a combination of the two. All costs are in June 2007 dollars, matching the cost basis of the BB study, to facilitate comparison. Note that, according to the *Chemical Engineering* Plant Cost Index, June-2007 dollars are nearly equivalent to January-2010 dollars. Owner's costs are included in the present estimates as they were in the BB study estimates. Baseline fuel cost for the reference cases was determined using data from the Energy Information Administration's (EIA) Annual Energy Outlook (AEO) 2008. The first year of capital expenditure (2007) cost used for natural gas is \$6.21/MMkJ (\$6.55/MMBtu), both on a HHV basis and in 2007 United States (U.S.) dollars.

All plant configurations are evaluated based on installation at a Greenfield site. The capacity factor was chosen to be consistent with the reference cases, or 85 percent for all cases. Since this study includes some one-of-a-kind equipment, the sensitivity of the cost of electricity (COE) to equipment cost was evaluated and is presented in the body of the report.

The net outputs of the cases in this study range from 406 to 650 MW. Exhibit ES-2 shows the cost, performance, and environmental summary for all cases. A brief description of each study case follows:

- Case 1 includes a series of three post-combustion capture cases that examine the effect of exhaust gas recycle (EGR) on the combustion turbine performance and on the monoethanolamine (MEA)-based carbon capture process. Case 1a recycles 35 percent of the exhaust gas and Case 1b 50 percent. Case 1c combines 35 percent EGR with reduced reboiler duty in the MEA process.

- Case 2 is a pre-combustion capture case that uses an autothermal reactor to convert the feed methane to primarily hydrogen (H_2), carbon monoxide (CO), carbon dioxide (CO_2) and water (H_2O). The reformed gas is then shifted to primarily H_2 and CO_2, and the CO_2 is removed in a methyldiethanolamine (MDEA) acid gas removal process. The high hydrogen content fuel is utilized in the combustion turbine.
- Case 3 is another pre-combustion CO_2 control process and is similar to Case 2. In Case 3 a high pressure partial oxidation reactor is used to convert the natural gas feed to primarily H_2, CO, CO_2 and H_2O. The gas is shifted to primarily H_2 and CO_2 and the CO_2 is removed using the MDEA process. The high hydrogen content gas stream is fed to the combustion turbine.
- Case 4 is an oxy-combustion based process. Natural gas is combusted in the combustion turbine using high purity oxygen as the oxidant. The flue gas exiting the heat recovery steam generator (HRSG) is recycled to the combustion turbine to act as a diluent. Most of the water is condensed in the HRSG prior to recycle, and the portion of the stream not recycled is sent to CO_2 compression for sequestration.
- Case 5 is another oxy-combustion process based on the technology being developed by Clean Energy Systems (CES). Using high purity oxygen, natural gas is combusted in a high pressure gas generator with recycled water/steam acting as diluents. Power is recovered in a high pressure expander. The exhaust enters a reheat combustor where additional natural gas and oxygen are combusted. Additional power is recovered in subsequent turbine expanders before the working fluid enters a partial condenser that recovers water for recycling. The remaining gas is sent to CO_2 compression where additional water is recovered and recycled.

The results are discussed below in the following order:

- Performance (efficiency and water usage)
- Cost (total overnight cost (TOC), cost of electricity (COE), and First Year Cost of CO_2 Avoided)

Exhibit ES-1 Case Descriptions

Case	Unit Cycle	Description	Steam Cycle, psig/°F/°F	Combustion Turbine	Gasifier/ Boiler Technology	Oxidant	NOx Control	Exhaust Gas Recycle	CO_2 Separation	CO_2 Capture Target
Ref1	NGCC	Without capture	2400/1050/1050	Advanced F Class	HRSG	Air	LNB and SCR	N/A	N/A	0%
Ref2	NGCC	Post-Combustion with capture	2400/1050/1050	Advanced F Class	HRSG	Air	LNB and SCR	N/A	MEA	90%
1a	NGCC	Post-Combustion with exhaust gas recycle	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	35%	MEA	90%
1b	NGCC	Post-Combustion with exhaust gas recycle	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	50%	MEA	90%
1c	NGCC	Post-Combustion with exhaust gas recycle	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	35%	Enhanced MEA	90%
2	NGCC	Pre-Combustion with autothermal reactor	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	N/A	MDEA	90%
3	NGCC	Pre-Combustion with high pressure Partial oxidizer	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	N/A	MDEA	90%
4	NGCC	Oxy-combustion with CO_2 recycle	2400/1200/1200	High Pressure Ratio	HRSG	O_2	N/A	N/A	Oxy-fuel	>99%
5	NGCC	Oxy-combustion with water/steam recycle	CES - Based Design	CES - Based Design	N/A	O_2	N/A	N/A	Oxy-fuel	>99%

Exhibit ES-2 Cost and Performance Summary and Environmental Profile for All Cases

Case	Ref1	Ref2	1a	1b	1c	2	3	4	5
CO_2 Capture	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Gross Power Output (kW)	564,700	511,00	513,000	515,500	521,800	645,200	727,148	585,900	528,700
Auxiliary Power Requirement (kW)	9,620	44,214	38,197	35,942	38,577	59,200	77,450	136,480	122,272
Net Power Output (kW)	555,080	466,786	474,803	479,558	483,223	586,000	649,698	449,420	405,928
Natural Gas Flowrate, kg/hr (lb/hr)	75,901 (167,333)	75,901 (167,333)	75,374 (166,172)	75,648 (166,774)	75,374 (166,172)	94,971 (209,375)	108,022 (238,148)	74,083 (163,325)	62,272 (137,285)
HHV Thermal Input (kW_{th})	1,105,812	1,105,812	1,098,140	1,102,121	1,098,140	1,383,644	1,573,791	1,079,327	907,255
Net Plant HHV Efficiency (%)	50.2%	42.2%	43.2%	43.5%	44.0%	42.4%	41.3%	41.6%	44.7%
Net Plant HHV Heat Rate, kJ/kWh (Btu/kWh)	7,172 (6,798)	8,528 (8,083)	8,326 (7,892)	8,274 (7,842)	8,181 (7,754)	8,500 (8,057)	8,720 (8,265)	8,646 (8,195)	8,046 (7,626)
Raw Water Withdrawal, m^3/min (gpm)	8.9 (2,362)	15.1 (3,980)	14.2 (3,741)	14.1 (3,729)	14.9 (3,944)	16.8 (4,430)	14.2 (3,762)	12.7 (3,444)	10.6 (2,801)
Raw Water Consumption, m^3/min (gpm)	6.9 (1,831)	11.3 (2,985)	10.6 (2,802)	10.5 (2,781)	11.2 (2,959)	13.8 (3,638)	11.7 (3,091)	9.3 (2,454)	7.8 (2,056)
Total Overnight Cost ($ x 1,000)	398,290	709,039	618,008	649.113	622,441	904,522	998,934	891,165	1,184,515
COE, total including TS & M costs (mills/kWh)[1]	58.90	87.17	81.22	82.02	80.01	87.44	88.08	96.69	112.24
LCOE, total including TS & M costs (mills/kWh)[1]	74.66	110.50	102.96	103.97	101.42	110.84	111.66	122.57	142.28
CO_2 emissions, tonne/yr (tons/yr)[1]	1,507,427 (1,661,654)	141,875 (156,391)	149,285 (164,558)	149,654 (164,966)	149,285 (164,558)	204,492 (225,414)	247,961 (273,331)	Negligible	Negligible
CO_2 emissions, kg/MWh (lb/MWh)[2]	359 (790)	40 (87)	39 (86)	39 (86)	38 (85)	43 (94)	46 (101)	Negligible	Negligible
CO_2 emissions, kg/MWh (lb/MWh)[3]	365 (804)	43 (96)	42 (93)	42 (92)	41 (91)	47 (103)	51 (113)	Negligible	Negligible
NO_X emissions, tonne/yr (ton/yr)[1]	115 (127)	98 (109)	102 (112)	102	102 (112)	770 (848)	874 (964)	Negligible	Negligible
NO_X emissions, kg/MWh (lb/MWh)[2]	0.027 (0.060)	0.027 (0.061)	0.027 (0.059)	0.027 (0.059)	0.026 (0.058)	0.160 (0.35)	0.161 (0.356)	Negligible	Negligible

[1] Capacity Factor is 85%
[2] Value is based on gross output
[3] Value is based on net output

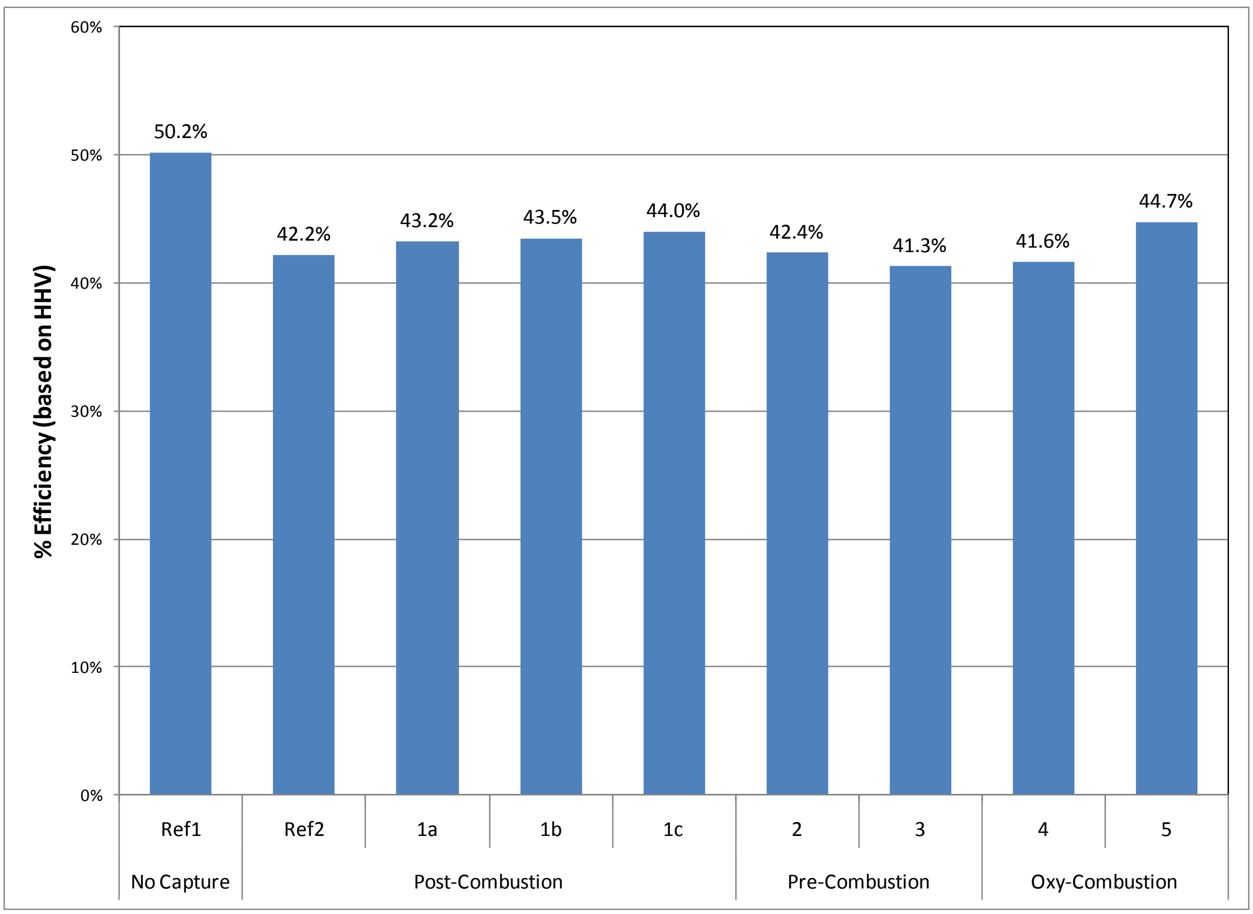
PERFORMANCE

Energy Efficiency

The net plant efficiency (HHV basis) for all cases is shown in Exhibit ES-3. The following conclusions can be drawn:

- As expected, the NGCC with no CO_2 capture (Ref1) has the highest net efficiency of the technologies modeled in this study with an efficiency of 50.2 percent.
- The NGCC case based on the Clean Energy Systems technology (case 5) results in the highest efficiency (44.7 percent) among all of the capture cases.
- The post-combustion-based NGCC cases have a slightly higher net efficiency than the pre-combustion or conventional oxy-combustion (CO_2 recycle – case 4) cases.
- The efficiency spread is only 3.5 percentage points between the highest and lowest efficiency capture technologies.

Exhibit ES-3 Net Plant Efficiency (HHV Basis)

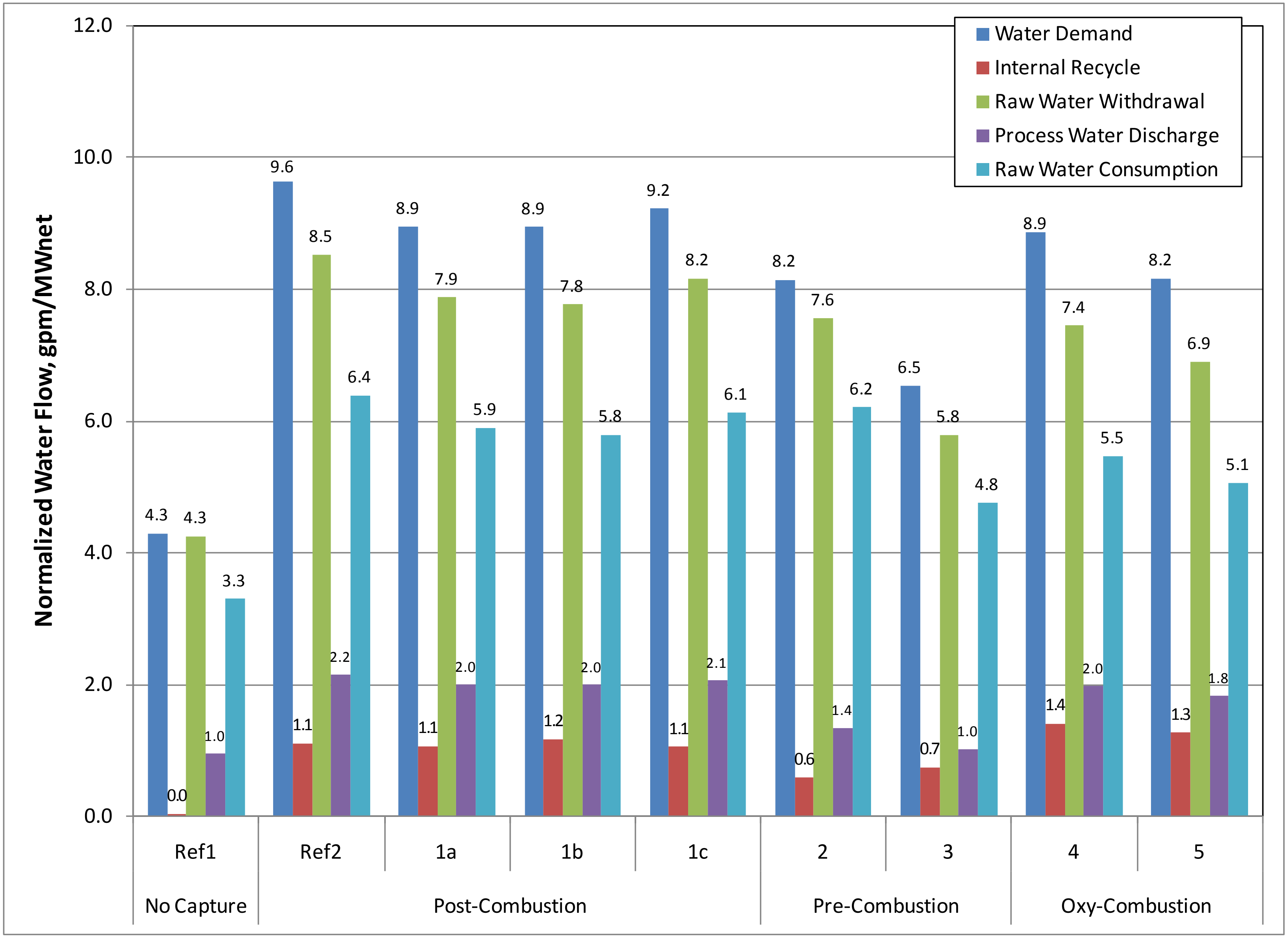


Water Use

Five water values are presented for each technology in Exhibit ES-4: water demand, internal recycle, raw water withdrawal, process water discharge, and raw water consumption. Each value is normalized by the net output. The total water demand for each subsystem was determined and internal recycle water available from various sources like boiler feed water (BFW) blowdown and condensate was applied to offset the water demand. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is the water removed from the ground or diverted from a surface-water source for use in the plant. Raw water consumption is also accounted for as the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source it was withdrawn from. Thus the difference between withdrawal and process water returned to the source is consumption. Consumption represents the net impact of the process on the water source. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). The following conclusions can be drawn:

- In all cases, the primary water consumer is the cooling tower makeup, which ranges from 70-100 percent of the total raw water withdrawal.
- The normalized water demand for the CO_2 capture cases is on average twice as great as the reference non-capture case. Additional water demand comes from the large cooling loads in the post-combustion cases using the Econamine process, from water gas shift and humidification requirements in the pre-combustion cases, and from lower net power output in the oxy-combustion cases.
- The normalized water demand varies from 6.5 to 9.6 gpm/MWnet for the capture cases. At the low end of the range is case 3, high pressure partial oxidation with amine CO_2 separation, primarily because the additional power recovered in the syngas expander results in the highest net output of all the cases as opposed to a significant reduction in water demand. At the high end of the range is the post-combustion capture reference case (Ref2), which has a relatively high water demand because of the Econamine process and relatively low net output.
- Raw water consumption for the capture cases varies over a narrower range than demand, 4.8 to 6.4 gpm/MWnet, primarily because the technologies with high demand also had high process water discharge which serves to reduce consumption.

Exhibit ES-4 Water Usage



COST RESULTS

Total Overnight Cost

The Total Overnight Cost (TOC) for each plant was calculated by adding owner's costs to the Total Plant Cost (TPC). The TPC for two reference cases was determined through a combination of vendor quotes, scaled estimates from previous design/build projects, or a combination of the two. The TPC for the new cases in this study were scaled from the reference estimates with certain unique equipment items estimated by WorleyParsons or scaled from other technology estimates. TPC includes all equipment (complete with initial chemical and catalyst loadings), materials, labor (direct and indirect), engineering and construction management, and contingencies (process and project). Escalation and interest on debt during the capital expenditure period were estimated and added to the TOC to provide the Total As-Spent Cost (TASC).

The normalized TOC and TASC components for each technology are shown in Exhibit ES-5.

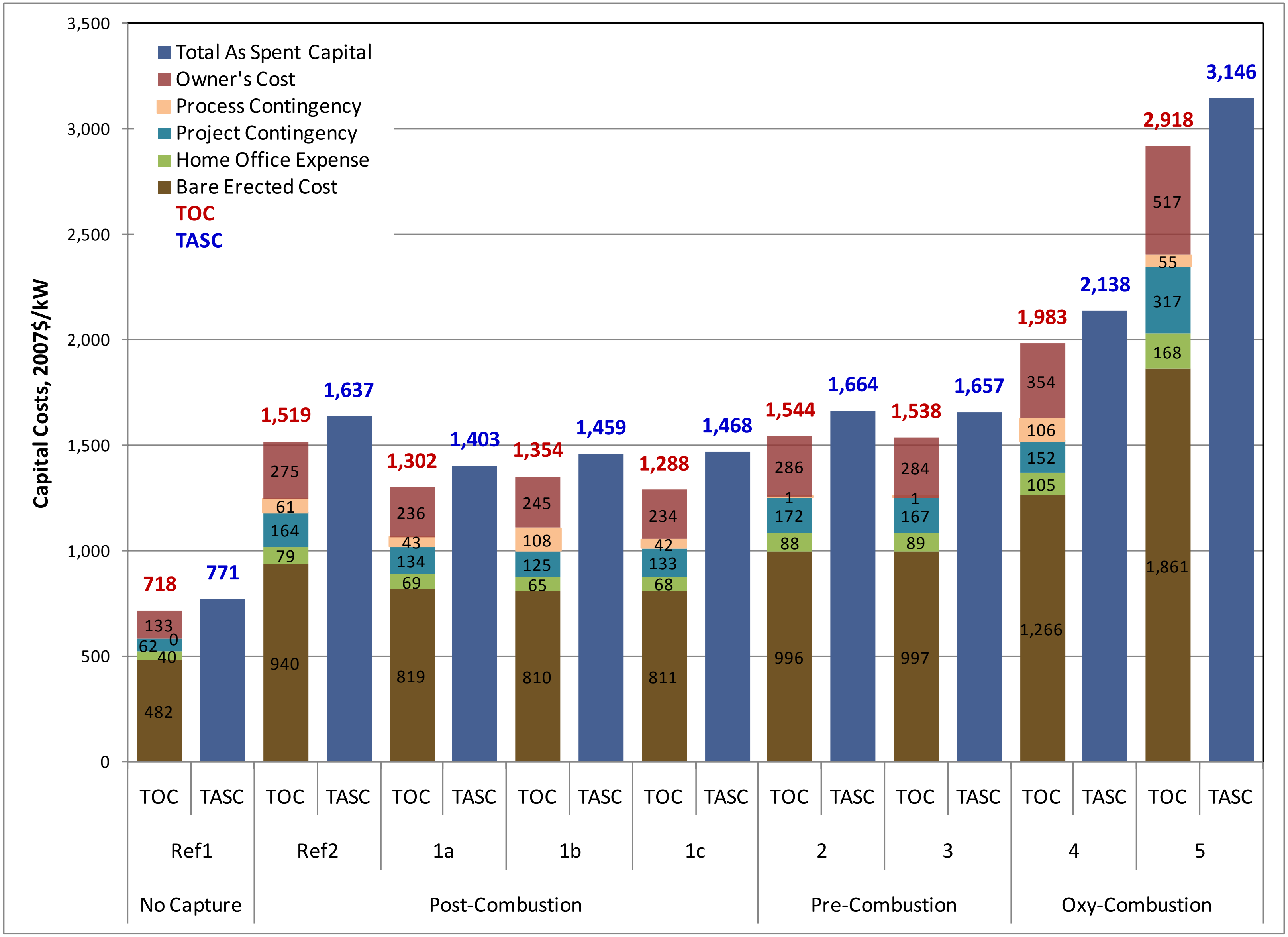
The cost estimates carry an accuracy of -15%/+30%, consistent with a "feasibility study" level of design engineering applied to the various cases in this study. The value of the study lies not in the absolute accuracy of the individual case results but in the fact that all cases were evaluated under the same set of technical and economic assumptions. This consistency of approach allows meaningful comparisons among the cases evaluated.

Project contingencies were added to the Engineering/Procurement/Construction Management (EPCM) capital accounts to cover project uncertainty and the cost of any additional equipment that would result from a detailed design. The contingencies represent costs that are expected to occur. Each bare erected cost (BEC) account was evaluated against the level of estimate detail and field experience to determine project contingency. Process contingency was added to cost account items that were deemed to be first-of-a-kind (FOAK) or posed significant risk due to lack of operating experience. The cost accounts that received a process contingency include:

- Combustion Turbine Generator - five percent contingency on Cases 1a and 1c for modifications required to the inlet air system (material, static mixer) and the compressor. A 40 percent contingency is used in Case 1b since major design modifications would be required to accommodate the low combustor oxygen concentration, and 40 percent was also used for Case 4 because of the oxy-combustion configuration and the high pressure ratio.
- MEA-based carbon capture processes - 20 percent contingency on Cases Ref2, 1a, 1b, and 1c because it is considered unproven at commercial scale for power plant applications.
- Gas Generator and Reheat Combustor – 15 percent contingency for high-pressure oxy-combustion reactors.

All cases except Cases 1a, 1b, and 1c, deviate substantially from the reference Case Ref2 in cycle configuration and in operating condition of key pieces of equipment. The severe operating temperatures and pressures required for some of this equipment pushes the envelope of existing material metallurgy and design practices. Because of the uncertainty associated with equipment cost in these instances, a sensitivity analysis was performed to show the impact of both higher and lower than estimated costs. The equipment items included in the sensitivity analysis include: the HRSG and Econamine process in the EGR cases, the high temperature, high pressure gas-gas heat exchanger and autothermal reactor in Case 2, and the partial oxidation reactor and high temperature, high pressure expander in Case 3, the high pressure ratio oxy-combustion turbine in Case 4, and the gas generators, air separation unit, and high temperature turboexpander in Case 5.

Exhibit ES-5 Plant Capital Costs



The following conclusions can be drawn:

- The post-combustion processes using EGR have the lowest normalized capital cost of any of the capture technologies, including the reference case.
- The pre-combustion capture cases have normalized TPC values that are 17 percent greater than the average of the post-combustion EGR cases, and the conventional oxy-combustion case is 51 percent more costly than the post-combustion cases.
- The oxy-combustion (CES-based) system is the most capital intensive of the systems studied with a normalized TPC more than double the post-combustion cases.

Cost of Electricity

The cost metric used in this study is the COE, which is the revenue received by the generator per net megawatt-hour during the power plant's first year of operation, *assuming that the COE escalates thereafter at a nominal annual rate equal to the general inflation rate, i.e., that it remains constant in real terms over the operational period of the power plant*. To calculate the COE, the Power Systems Financial Model (PSFM) [2] can be used to determine a "base-year" (2007) COE that, when escalated at an assumed nominal annual general inflation rate of 3 percent[1], provides the stipulated internal rate of return on equity over the entire economic analysis period (capital expenditure period plus thirty years of operation). The first year capital charge factors (CCF) shown in Exhibit ES-6 were derived using the PSFM and used to calculate COE using a simplified equation as detailed in Section 4.7.1.

Project financial structures vary depending on the type of project (high risk or low risk) and the length of the capital expenditure period. All cases in this study were assumed to be undertaken at investor owned utilities (IOUs). High risk projects are those in which commercial scale operating experience is limited. All NGCC cases with CO_2 capture were considered to be high risk. The non-capture NGCC reference case, Ref1, was considered to be low risk. All natural gas fueled cases were assumed to have a 3 year capital expenditure period. The current-dollar, 30-year levelized cost of electricity (LCOE) was also calculated and is shown in Exhibit ES-2 for reference, but the primary metric used in the balance of this study is COE. A more detailed discussion of the two metrics is provided in Section 4.7 of this report and Section 2.7.4 of the Bituminous Baseline study report [1].

Exhibit ES-6 Economic Parameters Used to Calculate COE

	High Risk (3 year capital expenditure period)	Low Risk (3 year capital expenditure period)
Capital Charge Factor	0.1111	0.1048

[1] This nominal escalation rate is equal to the average annual inflation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods. This index was used instead of the Producer Price Index for the Electric Power Generation Industry because the Electric Power Index only dates back to December 2003 and the Producer Price Index is considered the "headline" index for all of the various Producer Price Indices.

The COE results are listed in Exhibit ES-7 and shown graphically in Exhibit ES-8. The capital cost, fixed operating cost, variable operating cost, and fuel cost are shown separately. In the capture cases, the CO_2 transport, storage, and monitoring (TS&M) costs are also shown as a separate bar segment.

Exhibit ES-7 COE Component Details (mills/kWh or $/MWh) for All Cases

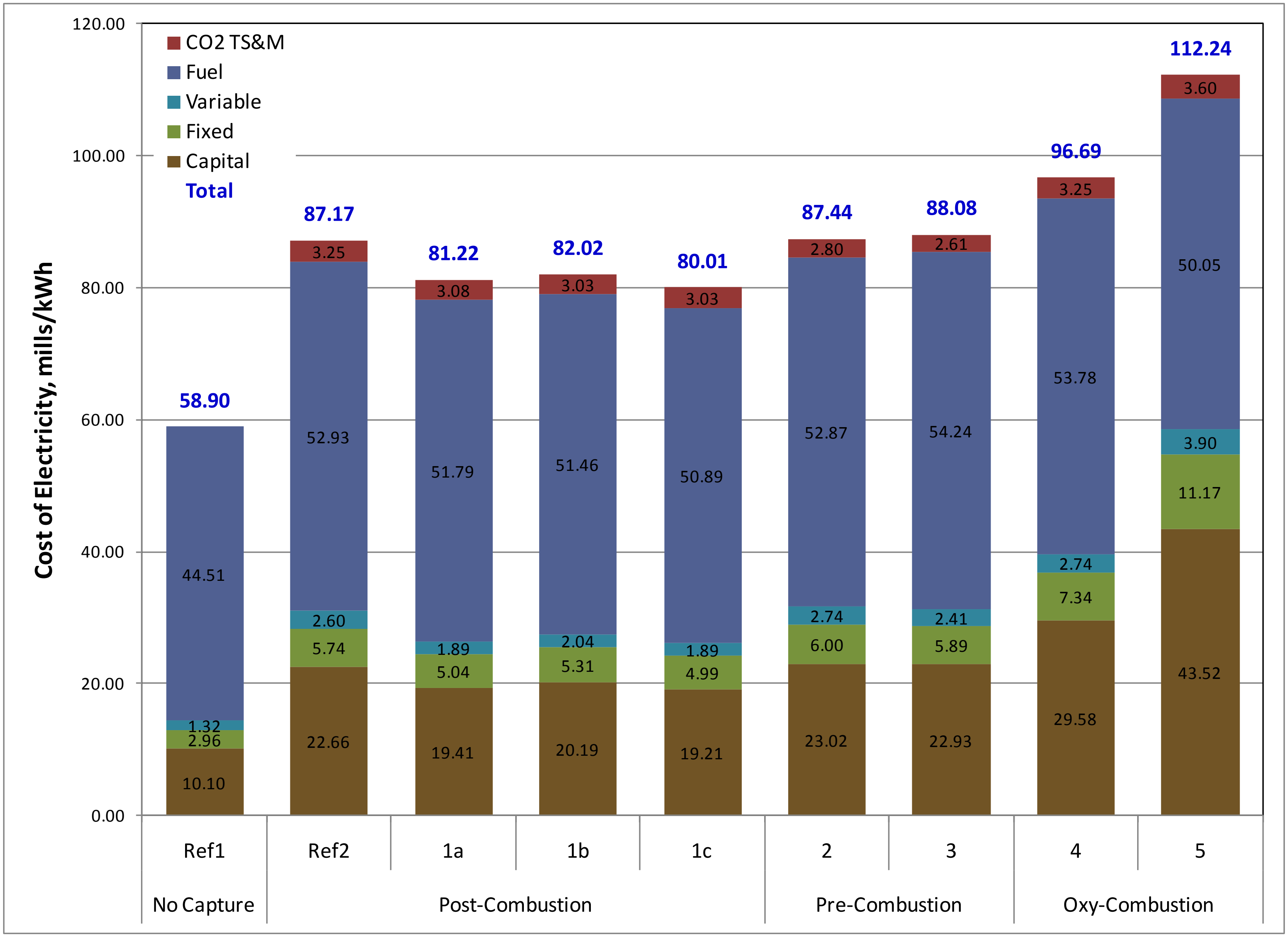
Case	Ref1	Ref2	1a	1b	1c	2	3	4	5
Capital	10.10	22.66	19.41	20.19	19.21	23.02	22.93	29.58	43.52
Fixed O&M	2.96	5.74	5.04	5.31	4.99	6.00	5.89	7.34	11.17
Variable O&M	1.32	2.60	1.89	2.04	1.89	2.74	2.41	2.74	3.90
Fuel	44.51	52.93	51.79	51.46	50.89	52.87	54.24	53.78	50.05
CO_2 TS&M total	0.0	3.25	3.08	3.03	3.03	2.80	2.61	3.25	3.60
Transport	*0.0*	*1.95*	*1.92*	*1.88*	*1.89*	*1.73*	*1.56*	*2.03*	*2.25*
Storage	*0.0*	*0.90*	*0.83*	*0.82*	*0.81*	*0.74*	*0.71*	*0.87*	*0.97*
Monitoring	*0.0*	*0.40*	*0.33*	*0.33*	*0.33*	*0.33*	*0.34*	*0.35*	*0.39*
COE Total	**58.90**	**87.17**	**81.22**	**82.02**	**80.01**	**87.44**	**88.08**	**96.69**	**112.24**
LCOE, total (including TS&M)	74.66	110.50	102.96	103.97	101.42	110.84	111.66	122.57	142.28

[1] CF is 85 percent for NGCC cases

The following conclusions can be drawn:

- Of the capture cases, the post-combustion cases utilizing EGR have the lowest COE by at least six percent. The next least expensive technologies are the reference case, which also uses post-combustion amine-based CO_2 capture but without EGR, and the pre-combustion cases, which are approximately equal.
- The COE of the conventional oxy-combustion case is approximately 11 percent higher than the reference case and 17 percent greater than the EGR cases.
- The CES-based process has the highest COE out of all the cases at 112 mills/kWh, primarily due to the very high capital cost.
- CO_2 transport, storage, and monitoring add approximately 3 mills/kWh to the COE, which is less than 3 percent of the total COE for all cases.

Exhibit ES-8 COE Components for All Cases



Cost of CO_2 Avoided

The cost of CO_2 avoided was calculated using Equation ES-1:

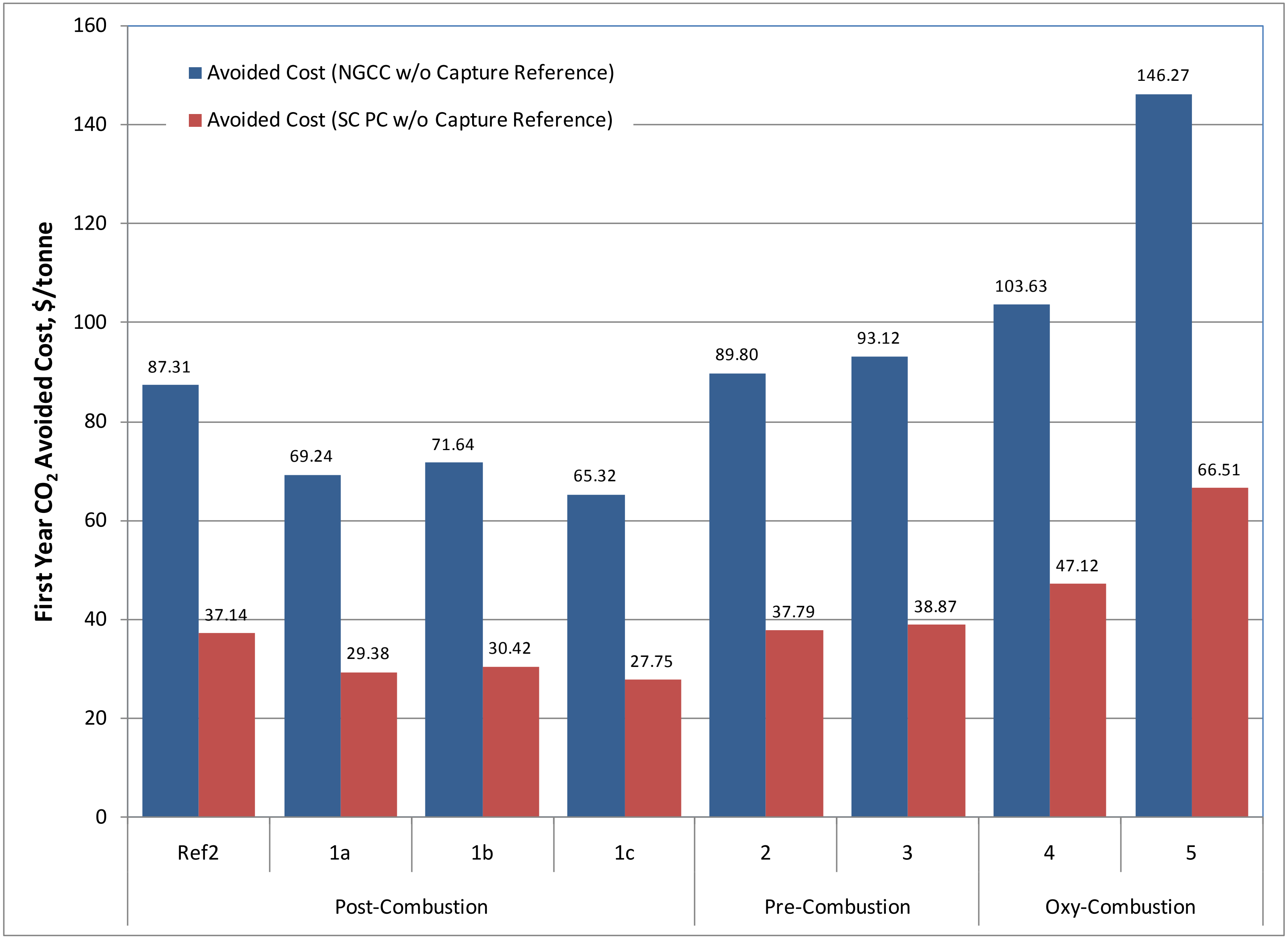
$$Avoided\ Cost = \frac{\{COE_{with\ removal} - COE_{reference}\}\ \$/MWh}{\{CO_2\ Emissions_{reference} - CO_2\ Emissions_{with\ removal}\}\ tonne/MWh} \qquad \text{(ES-1)}$$

The COE with CO_2 removal includes the costs of capture and compression as well as TS&M costs. The resulting avoided costs are shown in Exhibit ES-9 for each of the cases in this study. The avoided costs for each capture case are calculated using the NGCC non-capture case (Ref1) as the reference and again with Bituminous Baseline Supercritical Pulverized Coal without CO_2 capture as the reference [1].

The following conclusions can be drawn:

- The post-combustion capture cases using EGR have the smallest CO_2 avoided cost ($65.32/tonne).
- The CO_2 avoided cost of the conventional oxy-combustion case is approximately 1.5 times the EGR-based cases.
- The CO_2 avoided cost of the pre-combustion cases is approximately 1.3 times the EGR-based cases and approximately equal to the reference capture case.
- The CO_2 avoided cost of the CES based system is approximately twice as great as the EGR cases.

Exhibit ES-9 Summary and Comparison of CO_2 Avoided Costs for All Cases



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1. Introduction

Despite its recent price volatility, natural gas has historically been a relatively inexpensive fuel. As carbon regulations become more likely, natural gas is the favored fossil fuel because of its lower carbon intensity relative to coal. The use of carbon capture and sequestration (CCS) can further reduce the carbon emissions from a natural gas fueled power plant.

There are potential sources of renewable natural gas, including landfills and biomass-derived substitute natural gas (SNG). Landfill gas is currently being tapped at many locations, and gasification-based processes can be used to convert biomass into syngas, which can then be used to produce SNG. The quantity of available landfill gas is relatively small compared to the requirements of large scale electricity generation. Biomass-derived SNG is not currently cost competitive, and is still many years away from commercialization.

Until clean coal plants and other alternative energy sources become more feasible, conventional natural gas will continue to be used as a means of reducing emissions and generating electricity.

The objective of this report is to present a case study of the performance of advanced natural gas combined cycle (NGCC) plants with CCS, in a consistent technical manner that accurately reflects current market conditions for future developmental technologies.

Five different plant configurations were analyzed and compared to two reference cases from the Bituminous Baseline (BB) study [1]. The reference cases are NGCC plants without and with carbon capture. For consistency and writing purposes, these cases will be called Ref1 (Case 13 from the BB study) and Ref2 (Case 14 from the BB study with a modification to the amine auxiliary load calculation), respectively. The configurations are listed in Exhibit 1-1. Components of each plant configuration are described in more detail in the corresponding case description sections. All plants include CO_2 capture except Case Ref1.

Exhibit 1-1 Case Descriptions

Case	Unit Cycle	Description	Steam Cycle, psig/°F/°F	Combustion Turbine	Gasifier/ Boiler Technology	Oxidant	NOx Control	Exhaust Gas Recycle	CO_2 Separation	CO_2 Capture Target
Ref1	NGCC	Without capture	2400/1050/1050	Advanced F Class	HRSG	Air	LNB and SCR	N/A	N/A	0%
Ref2	NGCC	Post-Combustion with capture	2400/1050/1050	Advanced F Class	HRSG	Air	LNB and SCR	N/A	MEA	90%
1a	NGCC	Post-Combustion with exhaust gas recycle	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	35%	MEA	90%
1b	NGCC	Post-Combustion with exhaust gas recycle	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	50%	MEA	90%
1c	NGCC	Post-Combustion with exhaust gas recycle	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	35%	Enhanced MEA	90%
2	NGCC	Pre-Combustion with autothermal reactor	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	N/A	MDEA	90%
3	NGCC	Pre-Combustion with high pressure Partial oxidizer	2400/1050/1050	Advanced F Class	HRSG	Air	SCR	N/A	MDEA	90%
4	NGCC	Oxy-combustion with CO_2 recycle	2400/1200/1200	High Pressure Ratio	HRSG	O_2	N/A	N/A	Oxy-fuel	>99%
5	NGCC	Oxy-combustion with water/steam recycle	CES-Based Design	CES-Based Design	N/A	O_2	N/A	N/A	Oxy-fuel	>99%

2. General Evaluation Basis

For each of the plant configurations in this study an AspenPlus model was developed and used to generate material and energy balances. Performance and process limits were based upon published reports, information obtained from vendors and users of the technology, performance data from design/build utility projects, and/or best engineering judgment.

2.1 Site Characteristics

All plants in this study are assumed to be located at a generic plant site in Midwestern USA, with ambient conditions and site characteristics as presented in Exhibit 2-1 and Exhibit 2-2. The ambient conditions are the same as ISO conditions.

Exhibit 2-1 Site Ambient Conditions

Elevation, ft	0
Barometric Pressure, psia	14.696
Design Ambient Temperature, Dry Bulb, °F	59
Design Ambient Temperature, Wet Bulb, °F	51.5
Design Ambient Relative Humidity, %	60

Exhibit 2-2 Site Characteristics

Location	Greenfield, Midwestern USA
Topography	Level
Size, acres	100
Transportation	Rail
Ash/Slag Disposal	Off Site
Water	Municipal (50%) / Groundwater (50%)
Access	Land locked, having access by rail and highway
CO_2 Storage	Compressed to 15.3 MPa (2,215 psia), transported 80 kilometers (km) (50 miles), and sequestered in a saline formation at a depth of 1,239 m (4,055 ft)

The following design parameters are considered site-specific, and are not quantified for this study. Flood plain considerations

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise regulations – Impact on site and surrounding area

2.2 Natural Gas Characteristics

Natural gas is utilized as the main fuel, and its composition is presented in Exhibit 2-3 [3].

Exhibit 2-3 Natural Gas Composition

Component		Volume Percentage	
Methane	CH_4	93.1	
Ethane	C_2H_6	3.2	
Propane	C_3H_8	0.7	
n-Butane	C_4H_{10}	0.4	
Carbon Dioxide	CO_2	1.0	
Nitrogen	N_2	1.6	
	Total	100.0	
	LHV	**HHV**	
kJ/kg	47,454	52,581	
MJ/scm	34.71	38.46	
Btu/lb	20,410	22,600	
Btu/scf	932	1,032	

Note: Fuel composition is normalized and heating values are calculated

The first year cost of natural gas used in this study is \$6.21/MMkJ (\$6.55/MMBtu) (2007 cost of natural gas in 2007 dollars). The cost was determined using the following information from the EIA's 2008 AEO:

- The 2007 East North Central region delivered cost of natural gas to electric utilities in 2006 dollars, \$231.47/1000 m^3 (\$6.55/1000 ft^3), was obtained from the AEO 2008 reference case Table 108 and converted to an energy basis, \$6.02/MMkJ (\$6.35/MMBtu).
- The 2007 cost was escalated to 2007 dollars using the GDP chain-type price index from AEO 2008, resulting in a delivered 2007 price in 2007 dollars of \$6.21/MMkJ (\$6.55/MMBtu).[4] (Note: The natural gas cost of \$6.5478/MMBtu was used in calculations, but only two decimal places are shown in the report.)

3. Natural Gas Combined Cycle Case Performance Results

Five natural gas combined cycle (NGCC) power plant configurations were evaluated and are presented in this section. Each design is based on future developmental technologies, and all except Case 5 consist of two Advanced F-class combustion turbine generators (CTGs), two heat recovery system generators (HRSGs) and one steam turbine generator (STG) in a multi-shaft 2x2x1 configuration. Case 5 is based on Clean Energy System's (CES) oxy-combustion turbine. However, the case is a modified version because it recycles liquid H_2O and steam while the typical CES system recycles only liquid H_2O. Also included in this study are NGCC Reference Cases Ref1 and Ref2. Case Ref1 is Case 13 of the Bituminous Baseline (BB) study, while Case Ref2 is Case 14 from the same study [1] using an alternative method to estimate the auxiliary loads for the amine capture plant.

The study cases are evaluated with carbon capture, while Cases Ref1 and Ref2 are evaluated without and with carbon capture. Case Ref2 uses an Econamine FG Plus system to capture CO_2, while NGCC Cases 1a, 1b, and 1c also use the Econamine process, and Cases 2 and 3 use an MDEA system. Cases 4 and 5 are oxy-combustion cases. The NGCC designs that include Carbon Dioxide Recovery (CDR) have a smaller plant net output resulting from the additional CDR facility auxiliary loads. The sizes of the NGCC designs were determined by the output of the commercially available combustion turbine.

3.1 NGCC CASES

This section contains an evaluation of plant designs for NGCC Reference Cases (Ref1 and Ref2) and for new NGCC Cases 1 through 5. Cases Ref1 and Ref2 are similar in design and based on an NGCC plant with a constant thermal input. Both plants use an advanced 7F gas turbine and a single reheat steam cycle of 16.5 MPa/566°C/566°C (2400 psig/1050°F/1050°F). The only difference between the plants is that Case Ref2 includes CO_2 capture, while Case Ref1 does not.

Case s 1a, 1b, and 1c use exhaust gas recycle to increase the CO_2 concentration entering the post-combustion capture process. Case 2 and Case 3 are based on pre-combustion CO_2 removal. Case 2 uses an AutoThermal Reformer (ATR) and Case 3 uses a Partial Oxidation Reactor (POX) to generate syngas. Cases 1, 2 and 3 use a steam cycle based on the reference cases. Case 4 is an oxy-combustion process based on using CO_2 recycle. This design requires a unique turbine design (compressor pressure ratio = 45) to obtain a similar temperature profile as a 7F frame gas combustion turbine. Due to a higher gas turbine exhaust temperature, a single reheat steam cycle of 16.5 MPa/649°C/649°C (2400pisg/1200°F/1200°F) is used. Case 5 is an oxy-combustion process based on the Clean Energy Systems (CES) design. The working fluid is a mixture CO_2/Steam used in a series of turboexpanders for power generation. The working fluid in this design is a mixture of steam and carbon dioxide. For both oxy-combustion cases the CO_2 capture is > 99 percent. All new plants cases in the study capture and sequester CO_2.

3.1.1 Case Ref1-NGCC Reference Case without CO_2 Capture

In this section, the NGCC process without CO_2 capture is described and represents the non-capture reference case. The system description follows the block flow diagram (BFD) in Exhibit 3-1. A stream table, corresponding to the numbers listed on the BFD, is shown in

Exhibit 3-2. The BFD shows only one of the two combustion turbine/HRSG combinations, while the stream table shows totals for both process trains.

Ambient air (stream 1) and natural gas (stream 2) are combined in the dry Low NOx burner (LNB), which is operated to control the rotor inlet temperature at 1,371°C (2,500°F). The flue gas exits the turbine at 629°C (1,163°F) (stream 3) and passes into the HRSG. The HRSG generates both the main steam and reheat steam for the steam turbine. Flue gas exits the HRSG at 106°C (222°F) and passes to the plant stack

The plant produces a net output of 555 MW at a net plant efficiency of 50.2 percent (HHV basis).

Overall plant performance is summarized in Exhibit 3-3, which includes auxiliary power requirements.

Exhibit 3-1 Case Ref1 Block Flow Diagram, NGCC Reference Case without CO_2 Capture

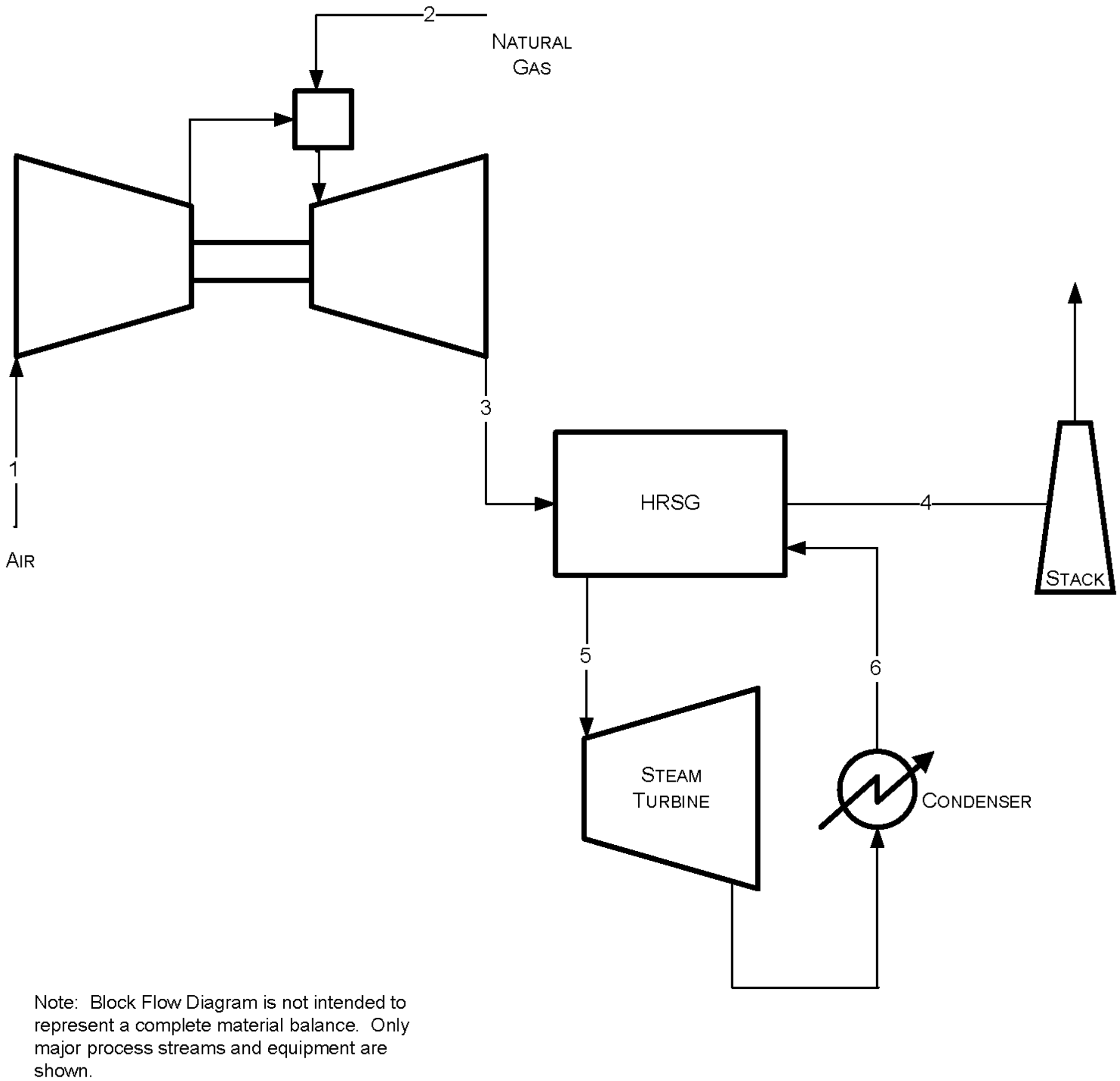


Exhibit 3-2 Case Ref1 Stream Table, NGCC Reference Case without CO_2 Capture

	1	2	3	4	5	6
V-L Mole Fraction						
Ar	0.0092	0.0000	0.0089	0.0089	0.0000	0.0000
CH_4	0.0000	0.9310	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0320	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0070	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0100	0.0404	0.0404	0.0000	0.0000
H_2O	0.0099	0.0000	0.0867	0.0867	1.0000	1.0000
N_2	0.7732	0.0160	0.7432	0.7432	0.0000	0.0000
O_2	0.2074	0.0000	0.1209	0.1209	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	109,323	4,380	113,831	113,831	21,589	28,545
V-L Flowrate (kg/hr)	3,154,735	75,901	3,230,636	3,230,636	388,927	514,240
Solids Flowrate (kg/hr)	0	0	0	0	0	0
Temperature (°C)	15	38	629	106	566	38
Pressure (MPa, abs)	0.10	3.10	0.11	0.10	16.65	0.01
Enthalpy (kJ/kg)[A]	30.23	46.30	835.81	248.81	3,472.36	160.61
Density (kg/m^3)	1.2	22.2	0.4	0.9	47.7	992.9
V-L Molecular Weight	28.857	17.328	28.381	28.381	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	241,016	9,657	250,954	250,954	47,595	62,930
V-L Flowrate (lb/hr)	6,955,000	167,333	7,122,333	7,122,333	857,437	1,133,706
Solids Flowrate (lb/hr)	0	0	0	0	0	0
Temperature (°F)	59	100	1,163	222	1,050	101
Pressure (psia)	14.7	450.0	15.2	14.7	2,414.7	1.0
Enthalpy (Btu/lb)[A]	13.0	19.9	359.3	107.0	1,492.8	69.1
Density (lb/ft^3)	0.076	1.384	0.025	0.057	2.977	61.982

Exhibit 3-3 Case Ref1 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	362,200
Steam Turbine Power	202,500
TOTAL POWER, kWe	**564,700**
AUXILIARY LOAD SUMMARY, kWe	
Condensate Pumps	170
Boiler Feedwater Pumps	2,720
Amine System Auxiliaries	0
CO_2 Compression	0
Circulating Water Pump	2,300
Ground Water Pumps	210
Cooling Tower Fans	1,190
SCR	10
Gas Turbine Auxiliaries	700
Steam Turbine Auxiliaries	100
Miscellaneous Balance of Plant[1]	500
Transformer Losses	1,720
TOTAL AUXILIARIES, kWe	**9,620**
NET POWER, kWe	**555,080**
Net Plant Efficiency (HHV)	50.2%
Net Plant Efficiency (LHV)	55.7%
Net Plant Heat Rate (HHV), kJ/kWh (Btu/kWh)	7,172 (6,798)
Net Plant Heat Rate (LHV), kJ/kWh (Btu/kWh)	6,466 (6,129)
CONDENSER COOLING DUTY, GJ/hr (10^6 Btu/hr)	**1,139 (1,080)**
CONSUMABLES	
Natural Gas Feed Flow, kg/hr (lb/hr)	75,901 (167,333)
Thermal Input (HHV), kW_{th}	1,105,812
Thermal Input (LHV) , kW_{th}	997,032
Raw Water Withdrawal, m^3/min (gpm)	8.9 (2,362)
Raw Water Consumption, m^3/min (gpm)	6.9 (1,831)
CO_2 Capture	N/A
CO_2 emissions tonne/yr (ton/yr)[2]	1,507,427 (1,661,654)
CO_2 emissions kg/MWh (lb/MWh)	359 (790)

1 Includes plant control systems, lighting, HVAC and miscellaneous low voltage loads
2 Based on 85% capacity factor

The estimated air emissions are shown in Exhibit 3-4. Operation of the modern, state-of-the-art gas turbine fueled by natural gas, coupled to a HRSG, results in very low NOx emissions and negligible amounts of particulate and SO_2. There are no mercury emissions in an NGCC plant.

The low level of NOx production (2.5 ppmvd at 15 percent O_2) is achieved by utilizing Selective Catalytic Reduction (SCR).

Exhibit 3-4 Case Ref1 Estimated Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (ton/year) 85% capacity factor	kg/MWh (lb/MWh)
SO_2	Negligible	Negligible	Negligible
NOx	0.004 (0.009)	115 (127)	0.027 (0.060)
Particulate	Negligible	Negligible	Negligible
Hg	Negligible	Negligible	Negligible
CO_2	50.9 (118.3)	1,507,427 (1,661,654)	359 (790)
CO_2 net[1]			365 (804)

[1] CO_2 emissions based on net power instead of gross power

The carbon balance is shown in Exhibit 3-5. The carbon input to the plant consists of carbon in the air and the carbon in the natural gas. Carbon leaves the plant as CO_2 through the stack. The percent of total carbon sequestered is defined as the amount of carbon product produced divided by the carbon in the natural gas feedstock, expressed as a percentage.

% Captured = Carbon in Product for Sequestration / Carbon in the Natural gas

or

0/54,822 *100 = 0%

Exhibit 3-5 Case Ref1 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	54,822 (120,863)	**Stack Gas**	55,251 (121,808)
Air (CO_2)	429 (946)		
Total	55,251 (121,808)	**Total**	55,251 (121,808)

An overall water balance for the plant is shown in Exhibit 3-6. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process and is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

Exhibit 3-6 Case Ref1 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Condenser Makeup *BFW Makeup*	**0.1 (23)** *0.1 (23)*	**0.0 (0)**	**0.1 (23)** *0.1 (23)*	**0.0 (0)**	**0.1 (23)**
Cooling Tower *BFW Blowdown*	**8.9 (2,362)** *0.0 (0)*	**0.1 (23)** *0.1 (23)*	**8.9 (2,339)** *-0.1 (-23)*	**2.0 (531)**	**6.8 (1,808)**
Total	**9.0 (2,385)**	**0.1 (23)**	**8.9 (2,362)**	**2.0 (531)**	**6.9 (1,831)**

An overall plant energy balance is provided in tabular form in Exhibit 3-7. The power out is the combined combustion turbine and steam turbine power after generator losses.

Exhibit 3-7 Case Ref1 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Natural Gas	3,981 (3,773)	2.7 (2.5)	0 (0)	3,984 (3,776)
GT Air	0 (0)	95.4 (90.4)	0 (0)	95 (90)
Raw Water Makeup	0 (0)	33.6 (31.9)	0 (0)	34 (32)
Auxiliary Power	0 (0)	0 (0)	35 (33)	35 (33)
TOTAL	**3,981 (3,773)**	**131.6 (124.8)**	**35 (33)**	**4,147 (3,931)**
Heat Out GJ/hr (MMBtu/hr)				
Cooling Tower Blowdown	0 (0)	14.9 (14.2)	0 (0)	15 (14)
Stack Gas	0 (0)	804 (762)	0 (0)	804 (762)
Condenser	0 (0)	1,141 (1,082)	0 (0)	1,141 (1,082)
Process Losses*	0 (0)	154 (146)	0 (0)	154 (146)
Power	0 (0)	0.0 (0.0)	2,033 (1,927)	2,033 (1,927)
TOTAL	**0 (0)**	**2,114 (2,004)**	**2,033 (1,927)**	**4,147 (3,931)**

*Process losses including steam turbine, combustion reactions, HRSG, gas turbine, and gas cooling are estimated to match the heat input to the plant.

3.1.2 Case Ref2–NGCC Reference Case with CO_2 Capture

As previously mentioned, Case Ref2 is a modified version of BB Case 14. The plant configuration for BB Case 14 is the same as Case Ref1 with the exception that a Carbon Dioxide Recovery (CDR) facility was added based on the Fluor Econamine FG Plus^{SM} technology [5, 6]. A typical flow diagram is shown in Exhibit 3-8. The BB Case 14 nominal net output decreased to 474 MW because the combustion turbine output was constant, and the CDR facility significantly increased the auxiliary power load. This auxiliary power load and the reboiler steam requirement, 1,598 Btu/lb$_{CO_2}$, were estimated by scaling from a Fluor quote provided for an earlier study [7].

The Case Ref2 modified version of BB Case 14 was developed to include a change to the amine system auxiliary load calculation. The modified amine system auxiliaries were determined by using Thermoflow's GT Pro software. The auxiliary load increased from 9,600 kW in the BB Case 14 to 16,364 kW for Case Ref2 based on the revised estimate. This modification only affected the performance. No changes were made in the modeling.

The process description for Case Ref2 is essentially the same as Case Ref1 with one notable exception, the addition of CO_2 capture. A BFD and stream tables are shown in Exhibit 3-9 and Exhibit 3-10 , respectively.

Case Ref2 produces a net output of 467 MW and a net plant efficiency of 42.2 percent (HHV basis). Overall performance is summarized in Exhibit 3-11, which includes auxiliary power requirements. For Case Ref2, the CDR facility, including CO_2 compression, accounts for over 71 percent of the auxiliary plant load. The circulating water system (circulating water pumps and cooling tower fan) accounts for nearly 15 percent of the auxiliary load, largely due to the high cooling water demand of the CDR facility.

High pressure (HP), intermediate pressure (IP), and low pressure (LP) steam streams are produced from their appropriate sections in the HRSG. The temperature and pressure for each steam level are the following:

- HP: 1050°F and 2,415 psia,
- IP: 1055°F and 360 psia, and
- LP: 537°F and 80 psia

Back pressure on the CT was maintained by an ID fan that is part of the amine unit, refer to Exhibit 3-8. This fan is a significant part of the amine auxiliaries accounting for more than 75 percent of the power.

Exhibit 3-8 Fluor Econamine FG Plus Typical Flow Diagram

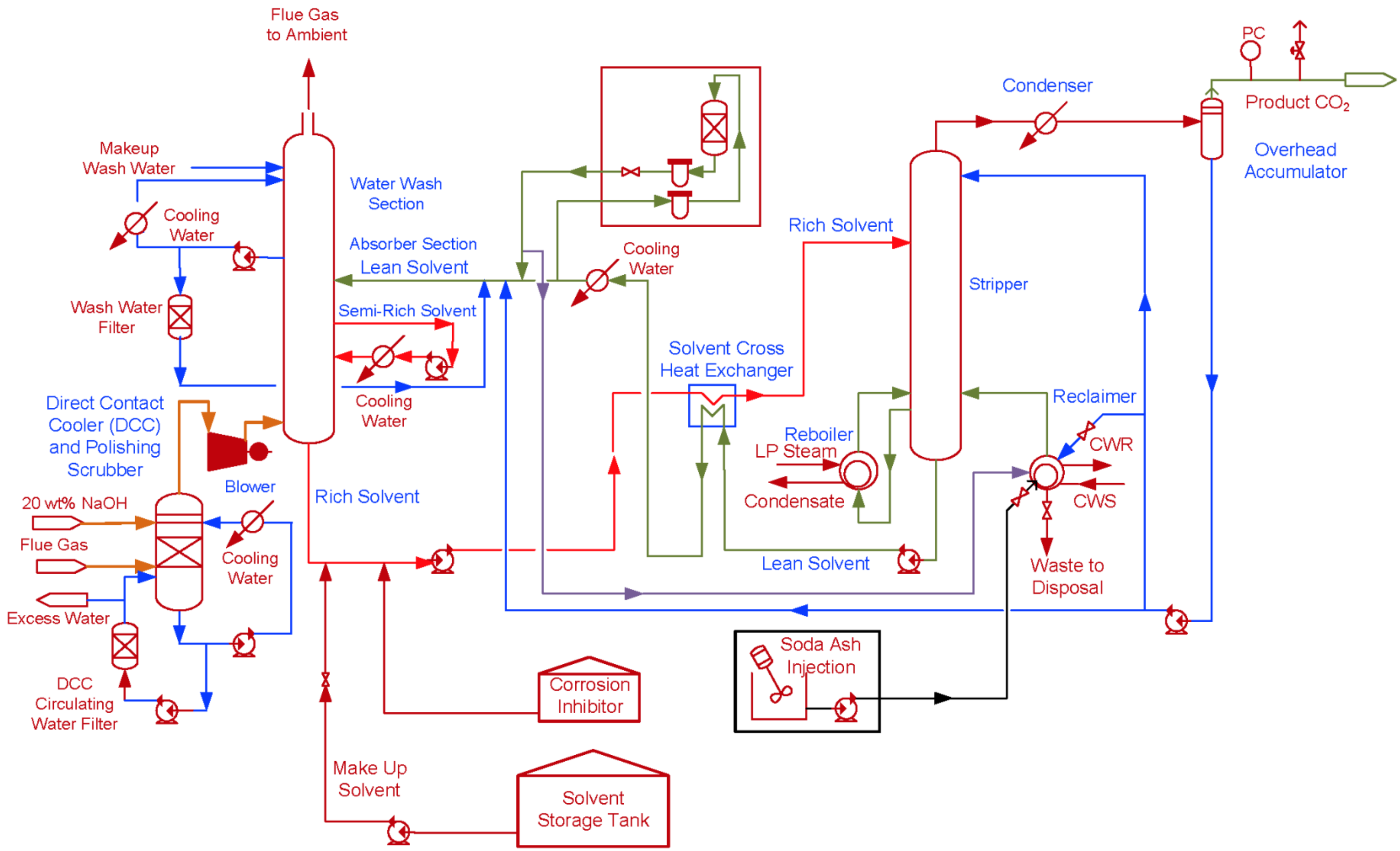
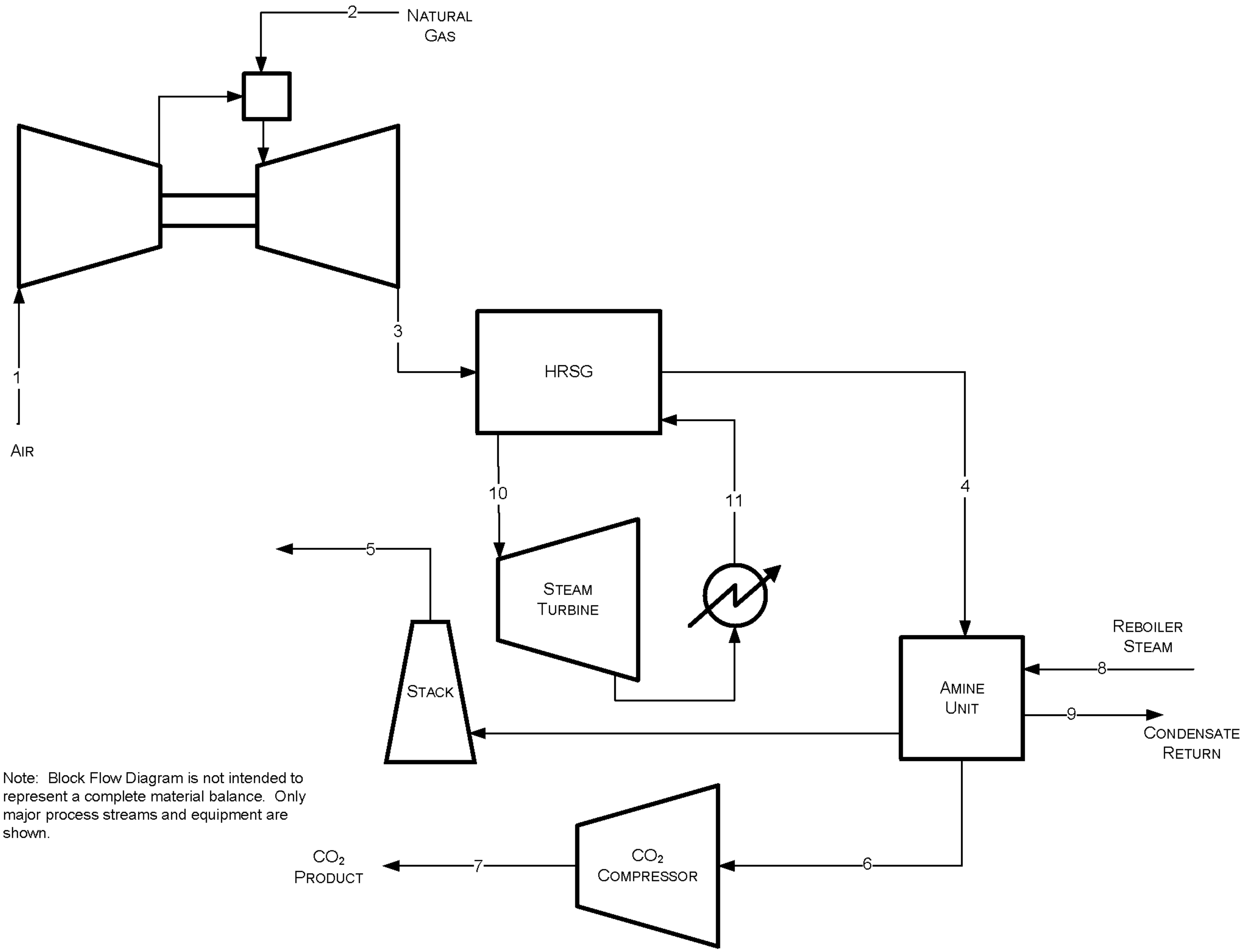


Exhibit 3-9 Case Ref2 Block Flow Diagram, NGCC Reference Case with CO_2 Capture



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

Exhibit 3-10 Case Ref2 Stream Table, NGCC Reference Case with CO_2 Capture

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0092	0.0000	0.0089	0.0089	0.0098	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.9310	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0100	0.0404	0.0404	0.0045	0.9893	1.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0099	0.0000	0.0867	0.0867	0.0339	0.0107	0.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.7732	0.0160	0.7432	0.7432	0.8187	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2074	0.0000	0.1209	0.1209	0.1332	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	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	109,323	4,380	113,831	113,831	103,333	4,185	4,140	17,887	17,887	21,589	13,482
V-L Flowrate (kg/hr)	3,154,735	75,901	3,230,636	3,230,636	2,933,892	183,013	182,203	322,243	322,243	388,927	242,889
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	38	629	143	30	21	51	152	151	566	38
Pressure (MPa, abs)	0.10	3.10	0.11	0.10	0.10	0.16	15.27	0.51	0.49	16.65	0.01
Enthalpy (kJ/kg)[A]	30.23	46.30	835.81	288.61	83.96	26.65	-164.90	2,746.79	635.72	3,472.36	160.61
Density (kg/m^3)	1.2	22.2	0.4	0.8	1.1	2.9	653.5	2.7	915.8	47.7	992.9
V-L Molecular Weight	28.857	17.328	28.381	28.381	28.393	43.731	44.010	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	241,016	9,657	250,954	250,954	227,809	9,226	9,127	39,435	39,435	47,595	29,724
V-L Flowrate (lb/hr)	6,955,000	167,333	7,122,333	7,122,333	6,468,125	403,474	401,689	710,425	710,425	857,437	535,480
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	100	1,163	290	85	69	124	306	304	1,050	101
Pressure (psia)	14.7	450.0	15.2	14.7	14.7	23.5	2,214.7	73.5	71.0	2,414.7	1.0
Enthalpy (Btu/lb)[A]	13.0	19.9	359.3	124.1	36.1	11.5	-70.9	1,180.9	273.3	1,492.8	69.1
Density (lb/ft^3)	0.076	1.384	0.025	0.052	0.071	0.183	40.800	0.169	57.172	2.977	61.982

Exhibit 3-11 Case Ref2 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	362,200
Steam Turbine Power	148,800
TOTAL POWER, kWe	**511,000**
AUXILIARY LOAD SUMMARY, kWe	
Condensate Pumps	80
Boiler Feedwater Pumps	2,710
Amine System Auxiliaries	16,364
CO_2 Compression	15,200
Circulating Water Pump	4,360
Ground Water Pumps	360
Cooling Tower Fans	2,250
SCR	10
Gas Turbine Auxiliaries	700
Steam Turbine Auxiliaries	100
Miscellaneous Balance of Plant[1]	500
Transformer Losses	1,580
TOTAL AUXILIARIES, kWe	**44,214**
NET POWER, kWe	**466,786**
Net Plant Efficiency (HHV)	42.2%
Net Plant Efficiency (LHV)	46.8%
Net Plant Heat Rate (HHV), kJ/kWh (Btu/kWh)	8,528 (8,083)
Net Plant Heat Rate (LHV), kJ/kWh (Btu/kWh)	7,689 (7,288)
CONDENSER COOLING DUTY, GJ/hr (10^6 Btu/hr)	**528 (500)**
CONSUMABLES	
Natural Gas Feed Flow, kg/hr (lb/hr)	75,901 (167,333)
Thermal Input (HHV), kW_{th}	1,105,812
Thermal Input (LHV) , kW_{th}	997,032
Raw Water Withdrawal, m^3/min (gpm)	15.1 (3,980)
Raw Water Consumption, m^3/min (gpm)	11.3 (2,985)
CO_2 Capture	90.7%
CO_2 emissions tonne/yr (ton/yr)[2]	141,875 (156,391)
CO_2 emissions kg/MWh (lb/MWh)	40 (87)

1 Includes plant control systems, lighting, HVAC and miscellaneous low voltage loads
2 Based on 85% capacity factor

The estimated air emissions for both plants are shown in Exhibit 3-12. Operation of the modern, state-of-the-art gas turbine fueled by natural gas, coupled to a HRSG, results in very low NOx emissions and negligible amounts of particulate and SO_2. There are no mercury emissions in an NGCC plant.

The low level of NOx production (2.5 ppmvd at 15 percent O_2) is achieved by utilizing a Selective Catalytic Reduction (SCR).

The Econamine FG+ system is assumed to remove 90 percent of the CO_2 in the flue gas.

Exhibit 3-12 Case Ref2 Estimated Air Emissions

	kg/GJ (lb/10^6 Btu)	Tonne/year (ton/year) 85% capacity factor	kg/MWh (lb/MWh)
SO_2	Negligible	Negligible	Negligible
NOx	0.004 (0.008)	115 (127)	0.027 (0.060)
Particulate	Negligible	Negligible	Negligible
Hg	Negligible	Negligible	Negligible
CO_2	5.1 (11.8)	141,875 (156,391)	40 (87)
CO_2 net[1]			43 (96)

[1] CO_2 emissions based on net power instead of gross power

For both plants, the carbon balance is shown in Exhibit 3-13. The carbon input to the plant consists of carbon in the air and the carbon in the natural gas. Carbon leaves the plant as CO_2 in the stack gas and CO_2 product. The percent of total carbon sequestered is defined as the amount of carbon product produced divided by the carbon in the natural gas feedstock, expressed as a percentage.

% Captured = Carbon in Product for Sequestration / Carbon in the Natural gas

or

$$49{,}726/54{,}822 * 100 = 90.7\%$$

Exhibit 3-13 Case Ref2 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	54,822 (120,863)	**Stack Gas**	5,525 (12,808)
Air (CO_2)	429 (946)	**CO_2 Product**	49,726 (109,628)
Total	55,251 (121,808)	**Total**	55,251 (121,808)

An overall water balance for both plants is shown in Exhibit 3-14. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process and is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

Exhibit 3-14 Case Ref2 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Econamine	**0.04 (12)**	**0.0 (0)**	**0.04 (12)**	**0.0 (0)**	**0.04 (12)**
Condenser Makeup	**0.1 (23)**	**0.0 (0)**	**0.1 (23)**	**0.0 (0)**	**0.1 (23)**
BFW Makeup	*0.1 (23)*		*0.1 (23)*		
Cooling Tower	**17.0 (4,477)**	**2.0 (520)**	**15.0 (3,958)**	**3.8 (1,007)**	**11.2 (2,951)**
BFW Blowdown		*0.1 (23)*	*-0.1 (-23)*		
Flue Gas Condensate		*1.9 (497)*	*-1.9 (-497)*		
CO_2 Product Condensate		**0.03 (8)**	**-0.03 (-8)**		
Total	**17.1 (4,512)**	**2.0 (528)**	**15.1 (3,992)**	**3.8 (1,007)**	**11.3 (2,985)**

An overall plant energy balance is provided in tabular form in Exhibit 3-15. The power out is the combined combustion turbine and steam turbine power after generator losses.

Exhibit 3-15 Case Ref2 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Natural Gas	3,981 (3,773)	2.7 (2.5)	0 (0)	3,984 (3,776)
GT Air	0 (0)	95.4 (90.4)	0 (0)	95 (90)
Raw Water Makeup	0 (0)	56.7 (53.7)	0 (0)	57 (54)
Auxiliary Power	0 (0)	0 (0)	159 (151)	159 (151)
TOTAL	**3,981 (3,773)**	**154.7 (146.6)**	**159 (151)**	**4,295 (4,071)**
Heat Out GJ/hr (MMBtu/hr)				
CO_2	0 (0)	-30.0 (-28.5)	0 (0)	-30.0 (-28.5)
Cooling Tower Blowdown	0 (0)	28.3 (26.8)	0 (0)	28.3 (26.8)
Econamine Losses	0 (0)	1,010.1 (957.4)	0 (0)	1,010.1 (957.4)
CO_2 Compression Intercooling	0 (0)	84.9 (80.5)	0 (0)	84.9 (80.5)
Stack Gas	0 (0)	246 (233)	0 (0)	246 (233)
Condenser	0 (0)	532 (504)	0 (0)	532 (504)
Process Losses*	0 (0)	584 (553)	0 (0)	584 (553)
Power	0 (0)	0.0 (0.0)	1,840 (1,744)	2,033 (1,927)
TOTAL	**0 (0)**	**2,455 (2,327)**	**1,840 (1,744)**	**4,295 (4,071)**

*Process losses including steam turbine, combustion reactions, HRSG, gas turbine, and gas cooling are estimated to match the heat input to the plant.

3.2 CASE 1-POST-COMBUSTION FLUE GAS RECYCLE AND AMINE ABSORPTION CASES

This case includes NGCC plants with 35 percent exhaust gas recycle (EGR) and amine absorption (Case 1a); 50 percent EGR and amine absorption (Case 1b); and 35 percent EGR, amine absorption, and a lower amine system reboiler steam requirement (Case 1c). The systems consist of two gas turbines, one steam turbine, and two HRSGs. The amine system used is monoethanolamine (MEA). Performance of the MEA system is based on in house performance estimates obtained from Fluor in 2005 [7]. A block flow diagram and stream tables for Cases 1a, 1b, and 1c are shown in Exhibit 3-17, Exhibit 3-18, Exhibit 3-19, and Exhibit 3-20, respectively. These cases were modeled after Case Ref2. The only addition was the EGR back to the combustion turbine (CT). The amine system auxiliaries were calculated using Thermoflow's GT Pro software in the same manner as for Case Ref2. This notion will be discussed further in the performance section.

The objective of this plant configuration is to decrease the volume and increase the CO_2 concentration of the flue gas treated by the amine CO_2 capture system. This task is accomplished using EGR from downstream of the HRSG to the CT compressor suction.

CO_2 concentrations of 6.7 and 8.8 volume percent are achieved by recycling 35 and 50 percent of the flue gas, respectively. The CO_2 concentration in the reference case without EGR was 4.0 volume percent. A constant CT compressor volumetric flow rate was maintained by reducing the ambient air intake as required.

Since 2005, Fluor has made process improvements that reduce the reboiler steam requirements and possibly the auxiliary electrical requirement. Case 1c uses an improved version of the process with the reboiler steam requirement reduced to 1,310 Btu/lb_{CO2} [8]. In all other respects Case 1c is identical to Case 1a.

For a conventional NGCC system, the EGR will reduce the oxygen concentration of the flue gas from about 12 percent to about 8 and 4 volume percent for EGRs of 35 and 50 percent, respectively.

The oxygen content of the gas stream exiting the CT combustor will be lower than the flue gas because additional air is added downstream of the combustor. Modeling for this study does not include detailed modeling of the CT combustor. However, the oxygen concentration of the gas exiting the combustor is calculated in Aspen and results in oxygen concentrations of approximately 5 and 2 percent for EGRs of 35 and 50 percent, respectively. A combustor study conducted by General Electric (GE) [9] indicates a minimum combustor exhaust oxygen concentration of about 4 percent without a negative impact on the combustor efficiency or CO emissions. Consequently, the 50 percent EGR configuration would not be acceptable, but was retained in this study to see what the impact would be if a combustor could be designed to operate with the lower oxygen concentration. The CO_2 capture achieved through the process is 90.5 percent for Cases 1a and 1c and 90.4 percent for Case 1b.

Comparing overall plant efficiencies of this study to Case Ref2 reveals a slight improvement. The efficiencies for Cases 1a, 1b, and 1c are 43.2%, 43.5%, and 44.0% (HHV) respectively, compared to 42.2% % (HHV) for Case Ref2. The differences are the addition of the EGR (Cases

1a, 1b and 1c) and the lower amine system steam requirement (Case 1c only). Exhibit 3-16 shows a comparison between Ref2 and the three configurations of Case 1.

Exhibit 3-16 Cases Ref2, 1a, 1b, and 1c Comparison

Case	Efficiency, % HHV	Amine System Inlet		Amine System Auxiliary Power (kWe)	Reboiler Duty (Btu/lbCO$_2$)	Steam Turbine Power (kWe)
		%O$_2$	%CO$_2$			
Ref2	42.2%	12.1	4	16,364	1,605	148,800
1a	43.2%	8.3	6.7	10,637	1,577	151,400
1b	43.5%	4.5	8.8	8,182	1,561	155,700
1c	44.0%	8.3	6.7	10,367	1,310	160,200

The results show that increasing the amount of EGR and reducing the Econamine reboiler duty both increase the system efficiency.

Case 1a produces a net output of 475 MW at a net plant efficiency of 43.2 percent (HHV), while Case 1b produces a net output of 480 MW at a net plant efficiency of 43.5 percent (HHV). Case 1c's net output is 483 MW at a net plant efficiency of 44.0 percent (HHV). Overall plant performance for all cases is summarized in Exhibit 3-21. The summary includes auxiliary power requirements. Back pressure on the CT was maintained by an ID fan that is part of the amine unit, refer to Exhibit 3-8.

The amine system auxiliaries for all cases (1a, 1b, and 1c) were determined using the results obtained from Case Ref2. Case Ref2's amine system auxiliaries were estimated using Thermoflow's GT Pro. The result for Case Ref2 was scaled to obtain the amine system auxiliaries for Cases 1a, 1b, and 1c. The auxiliary loads were determined by the following calculation which accounts for a reduced gas flow rate entering the amine system with EGR:

Amine System Auxiliaries (kW$_e$) = Case Ref2 Amine System Auxiliaries*((1-EGR)/100)

Therefore, using 35 percent EGR (Cases 1a and 1c):

Amine System Auxiliaries = 16,364*((1-35)/100) = 10,637 kW$_e$

For Case 1b, the scaled amine system auxiliaries were 8,182 kW.

The gas turbine (GT) power for Cases 1a, 1b, and 1c ranges from 360,000-362,000 kW compared to 362,200 kW for Ref2. The difference between the cases is minimal (approximately one percent), and several possibilities may account for this:

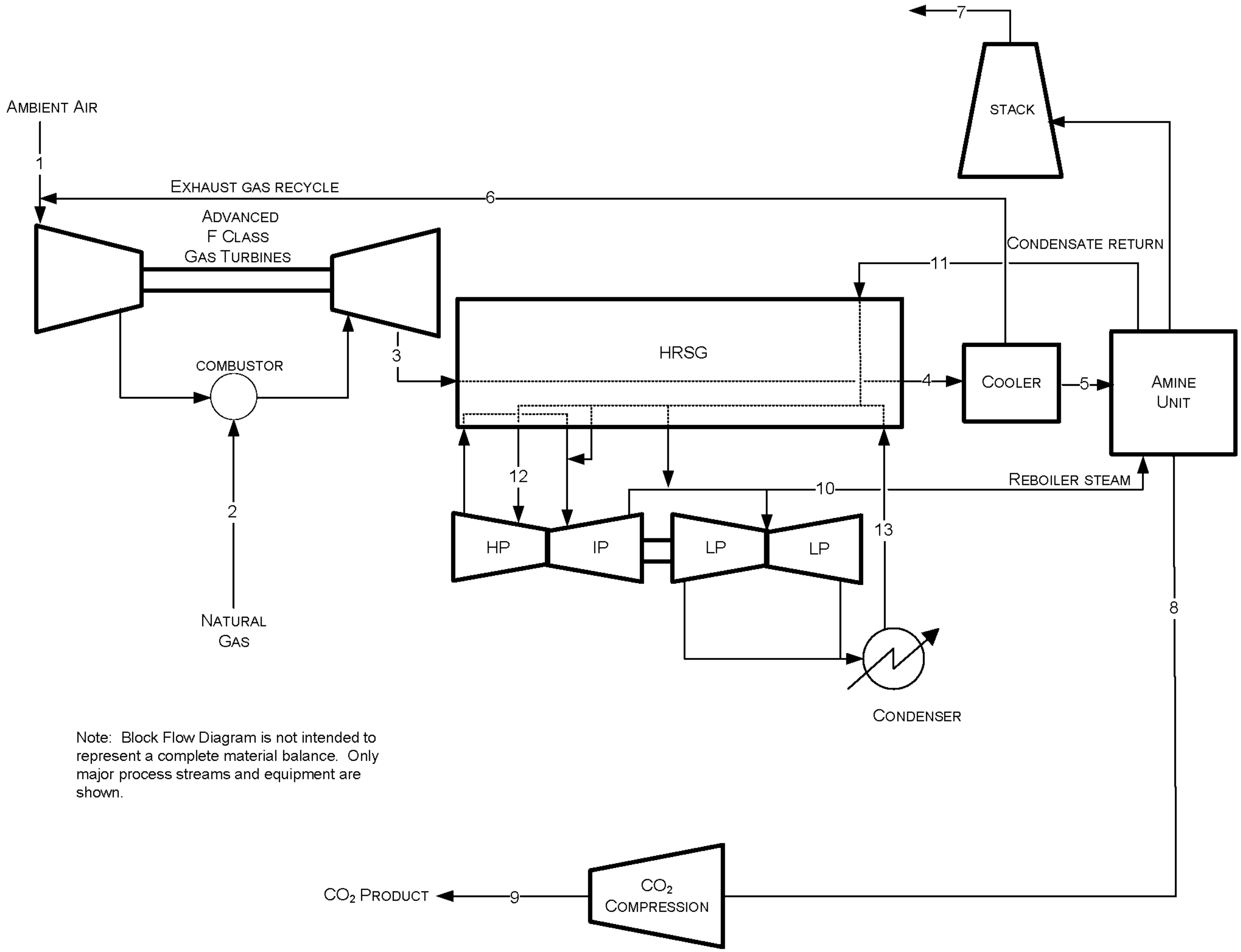
- Working fluid

  For Case Ref2, there was no recycle returned to the CT. Therefore, the working fluid was combusted fuel and air. For all other cases (1a, 1b, and 1c), the EGR (35 percent or 50 percent) is returned to the CT and mixed with the air. In these cases, the working fluid is a combination of the EGR, combusted fuel and air, therefore producing more CO_2 and H_2O, but less O_2. This new working fluid would have slightly different thermal and physical properties than the working fluid of Case Ref2.

- Modeling anomalies

  Design specification tolerances may not be sufficiently tight to distinguish the magnitude of difference between the reference case and the EGR cases.

Exhibit 3-17 Cases 1a, 1b, and 1c Block Flow Diagram, Exhaust Gas Recycle and Amine Absorption



Note: Actual process consists of 2 Advanced F-class CTGs, 2 HRSGs, and 1 STG in a multi-shaft 2x2x1 configuration

Exhibit 3-18 Case 1a Stream Table, 35% Exhaust gas recycle and amine absorption

	1	2	3	4	5	6	7	8	9	10	11	12	13
V-L Mole Fraction													
Ar	0.0092	0.0000	0.0089	0.0089	0.0095	0.0095	0.0102	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.9310	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0100	0.0631	0.0631	0.0668	0.0668	0.0072	0.9921	1.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0099	0.0000	0.1006	0.1006	0.0480	0.0480	0.0383	0.0079	0.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.7732	0.0160	0.7492	0.7492	0.7930	0.7930	0.8551	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2074	0.0000	0.0782	0.0782	0.0827	0.0827	0.0892	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	69,898	4,350	111,116	111,116	68,236	36,742	63,284	4,133	4,100	17,404	17,404	21,984	13,910
V-L Flowrate (kg/hr)	2,017,037	75,374	3,159,551	3,159,551	1,981,833	1,067,140	1,786,041	181,031	180,441	313,543	313,543	396,053	250,589
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	38	635	141	32	32	32	21	51	152	151	566	38
Pressure (MPa, abs)	0.10	3.10	0.11	0.10	0.10	0.10	0.10	0.16	15.27	0.51	0.49	16.65	0.01
Enthalpy (kJ/kg)[A]	30.22	46.30	871.89	309.68	107.20	107.20	93.61	23.68	-164.90	2,746.79	635.72	3,472.36	160.61
Density (kg/m^3)	1.2	22.2	0.4	0.8	1.2	1.2	1.1	2.9	653.5	2.7	915.8	47.7	992.9
V-L Molecular Weight	28.857	17.328	28.435	28.435	29.044	29.044	28.223	43.804	44.010	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	154,098	9,590	244,969	244,969	150,434	81,003	139,517	9,111	9,039	38,370	38,370	48,467	30,666
V-L Flowrate (lb/hr)	4,446,805	166,172	6,965,618	6,965,618	4,369,193	2,352,641	3,937,546	399,105	397,804	691,243	691,243	873,147	552,453
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	100	1,174	287	90	90	89	69	124	306	304	1,050	101
Pressure (psia)	14.7	450.0	15.2	14.7	14.6	14.6	14.7	23.5	2,214.7	73.5	71.0	2,414.7	1.0
Enthalpy (Btu/lb)[A]	13.0	19.9	374.8	133.1	46.1	46.1	40.2	10.2	-70.9	1,180.9	273.3	1,492.8	69.1
Density (lb/ft^3)	0.076	1.384	0.025	0.052	0.072	0.072	0.071	0.183	40.800	0.169	57.172	2.977	61.982

Note: Flows shown are totals for all process trains

Exhibit 3-19 Case 1b Stream Table, 50% Exhaust Gas Recycle and Amine Absorption

	1	2	3	4	5	6	7	8	9	10	11	12	13
V-L Mole Fraction													
Ar	0.0092	0.0000	0.0091	0.0091	0.0097	0.0097	0.0106	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.9310	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0100	0.0829	0.0829	0.0883	0.0883	0.0097	0.9940	1.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0099	0.0000	0.1068	0.1068	0.0480	0.0480	0.0383	0.0060	0.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.7732	0.0160	0.7591	0.7591	0.8091	0.8091	0.8918	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2074	0.0000	0.0421	0.0421	0.0449	0.0449	0.0495	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	54,030	4,366	110,239	110,239	51,716	51,716	46,920	4,135	4,110	17,273	17,273	22,709	14,415
V-L Flowrate (kg/hr)	1,559,150	75,648	3,146,979	3,146,979	1,512,180	1,512,182	1,318,934	181,335	180,888	311,175	311,175	409,105	259,682
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	38	641	140	32	32	32	21	51	152	151	566	38
Pressure (MPa, abs)	0.10	3.10	0.11	0.10	0.10	0.10	0.10	0.16	15.27	0.51	0.49	16.65	0.01
Enthalpy (kJ/kg)[A]	30.22	46.30	892.51	318.02	106.64	106.64	93.99	21.68	-164.90	2,746.79	635.72	3,472.36	160.61
Density (kg/m^3)	1.2	22.2	0.4	0.8	1.2	1.2	1.1	2.9	653.5	2.7	915.8	47.7	992.9
V-L Molecular Weight	28.857	17.328	28.547	28.547	29.240	29.240	28.110	43.854	44.010	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	119,116	9,625	243,035	243,035	114,015	114,015	103,441	9,116	9,061	38,080	38,080	50,064	31,779
V-L Flowrate (lb/hr)	3,437,337	166,774	6,937,902	6,937,902	3,333,786	3,333,791	2,907,751	399,776	398,789	686,022	686,022	901,922	572,500
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	100	1,186	284	90	90	89	69	124	306	304	1,050	101
Pressure (psia)	14.7	450.0	15.2	14.7	14.6	14.6	14.7	23.5	2,214.7	73.5	71.0	2,414.7	1.0
Enthalpy (Btu/lb)[A]	13.0	19.9	383.7	136.7	45.8	45.8	40.4	9.3	-70.9	1,180.9	273.3	1,492.8	69.1
Density (lb/ft^3)	0.076	1.384	0.025	0.053	0.073	0.073	0.070	0.184	40.800	0.169	57.172	2.977	61.982

Note: Flows shown are totals for all process trains

Exhibit 3-20 Case 1c Stream Table, 35% Exhaust Gas Recycle, Lower Amine Steam Requirement, & Amine Absorption

	1	2	3	4	5	6	7	8	9	10	11	12	13
V-L Mole Fraction													
Ar	0.0092	0.0000	0.0089	0.0089	0.0095	0.0095	0.0102	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.9310	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0100	0.0631	0.0631	0.0668	0.0668	0.0072	0.9921	1.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0099	0.0000	0.1006	0.1006	0.0480	0.0480	0.0383	0.0079	0.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.7732	0.0160	0.7492	0.7492	0.7930	0.7930	0.8551	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2074	0.0000	0.0782	0.0782	0.0827	0.0827	0.0892	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	69,898	4,350	111,116	111,116	68,236	36,742	63,284	4,133	4,100	14,461	14,461	21,984	16,388
V-L Flowrate (kg/hr)	2,017,037	75,374	3,159,551	3,159,551	1,981,833	1,067,140	1,786,041	181,031	180,441	260,516	260,516	396,053	295,240
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	38	635	135	32	32	32	21	51	152	151	566	38
Pressure (MPa, abs)	0.10	3.10	0.11	0.10	0.10	0.10	0.10	0.16	15.27	0.51	0.49	16.65	0.01
Enthalpy (kJ/kg)[A]	30.22	46.30	871.89	302.98	107.20	107.20	93.61	23.68	-164.90	2,746.79	635.72	3,472.36	160.61
Density (kg/m^3)	1.2	22.2	0.4	0.9	1.2	1.2	1.1	2.9	653.5	2.7	915.8	47.7	992.9
V-L Molecular Weight	28.857	17.328	28.435	28.435	29.044	29.044	28.223	43.804	44.010	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	154,098	9,590	244,969	244,969	150,434	81,003	139,517	9,111	9,039	31,881	31,881	48,467	36,130
V-L Flowrate (lb/hr)	4,446,805	166,172	6,965,618	6,965,618	4,369,193	2,352,641	3,937,546	399,105	397,804	574,340	574,340	873,147	650,893
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	100	1,174	275	90	90	89	69	124	306	304	1,050	101
Pressure (psia)	14.7	450.0	15.2	14.7	14.6	14.6	14.7	23.5	2,214.7	73.5	71.0	2,414.7	1.0
Enthalpy (Btu/lb)[A]	13.0	19.9	374.8	130.3	46.1	46.1	40.2	10.2	-70.9	1,180.9	273.3	1,492.8	69.1
Density (lb/ft^3)	0.076	1.384	0.025	0.053	0.072	0.072	0.071	0.183	40.800	0.169	57.172	2.977	61.982

Note: Flows shown are totals for all process trains

The estimated air emissions are shown in Exhibit 3-22. Operation of the modern, state-of-the-art gas turbine fueled by natural gas, coupled to a HRSG, results in very low NOx emissions and negligible amounts of particulate and SO_2. There are no mercury emissions in an NGCC plant.

The low level of NOx production (2.5 ppmvd at 15 percent O_2) is achieved by utilizing Selective Catalytic Reduction (SCR).

The Econamine FG+ system removed 90 percent of the CO_2 in the flue gas.

The carbon balance is shown in Exhibit 3-23. The carbon input to the plant consists of carbon in the air and the carbon in the natural gas. Carbon leaves the plant as CO_2 in the stack gas and CO_2 product. The percent of total carbon sequestered is defined as the amount of carbon product produced divided by the carbon in the natural gas feedstock, expressed as a percentage.

% Captured = Carbon in Product for Sequestration / Carbon in the Natural gas

or

49,245/54,442 *100 = 90.5% (1a)

49,367/54,639 *100 = 90.4% (1b)

49,245/54,442 *100 = 90.5% (1c)

An overall water balance for the plant is shown in Exhibit 3-24. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process and is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

An overall plant energy balance is provided in tabular form in Exhibit 3-25. The power out is the combined combustion turbine and steam turbine power after generator losses.

Exhibit 3-21 Case 1a, 1b, and 1c Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)			
Case	**1a**	**1b**	**1c**
Gas Turbine Power	361,600	359,800	361,600
Steam Turbine Power	151,400	155,700	160,200
TOTAL POWER, kWe	**513,000**	**515,500**	**521,800**
AUXILIARY LOAD SUMMARY, kWe			
Condensate Pumps	80	90	100
Boiler Feedwater Pumps	2,750	2,840	2,750
Amine System (MEA) Auxiliaries	10,637	8,182	10,637
CO_2 Compression	14,910	14,950	14,910
Circulating Water Pump	4,110	4,150	4,310
Ground Water Pumps	340	340	360
Cooling Tower Fans	2,120	2,140	2,230
SCR	10	10	10
Gas Turbine Auxiliaries	1,000	1,000	1,000
Steam Turbine Auxiliaries	100	100	100
Miscellaneous Balance of Plant[1]	500	500	500
Transformer Losses	1,640	1,640	1,670
TOTAL AUXILIARIES, kWe	**38,197**	**35,942**	**38,577**
Plant Performance			
NET POWER, kWe	**474,803**	**479,558**	**483,223**
Net Plant Efficiency (HHV)	43.2%	43.5%	44.0%
Net Plant Efficiency (LHV)	48.0%	48.3%	48.8%
Net Plant Heat Rate (HHV), kJ/kWh (Btu/kWh)	8,326 (7,892)	8,274 7,842)	8,181 (7,754)
Net Plant Heat Rate (LHV), kJ/kWh (Btu/kWh)	7,507 (7,115)	7,460 (7,070)	7,376 (6,991)
CONDENSER COOLING DUTY, GJ/hr (10^6 Btu/hr)	549 (520)	570 (540)	654 (620)
CONSUMABLES			
Natural Gas Feed Flow, kg/hr (lb/hr)	75,374 (166,172)	75,648 (166,774)	75,374 (166,172)
Thermal Input (HHV), kW_{th}	1,098,140	1,102,121	1,098,140
Thermal Input (LHV) , kW_{th}	990,114	993,704	990,114
Raw Water Withdrawal, m^3/min (gpm)	14.2 (3,743)	14.1 (3,731)	14.9 (3,946)
Raw Water Consumption, m^3/min (gpm)	10.6 (2,804)	10.5 (2,783)	11.2 (2,961)
CO_2 Capture	90.5%	90.4%	90.5%
CO_2 emissions tonne/yr (ton/yr)[2]	149,285 (164,558)	149,654 (164,966)	149,285 (164,558)
CO_2 emissions kg/MWh (lb/MWh)	39 (86)	39 (86)	38 (85)

1 Includes plant control systems, lighting, HVAC and miscellaneous low voltage loads
2 Based on 85% capacity factor

Exhibit 3-22 Cases 1a, 1b, and 1c Estimated Air Emissions

	kg/GJ ($lb/10^6$ Btu)			Tonne/year (ton/year) 85% capacity factor			kg/MWh (lb/MWh)		
Case	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**
SO_2	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible
NOx	0.003 (0.008)	0.003 (0.008)	0.003 (0.008)	102 (112)	102 (113)	102 (112)	0.027 (0.059)	0.027 (0.059)	0.026 (0.058)
Particulate	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible
Hg	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible	Negligible
CO_2	5.1 (11.8)	5.1 (11.8)	5.1 (11.8)	149,285 (164,558)	149,654 (164,966)	149,285 (164,558)	39 (86)	39 (86)	38 (85)
CO_2 net[1]							42 (93)	42 (92)	41 (91)

[1] CO_2 emissions based on net power instead of gross power

Exhibit 3-23 Cases 1a, 1b, and 1c Carbon Balances

Carbon In, kg/hr (lb/hr)				Carbon Out, kg/hr (lb/hr)			
Case	**1a**	**1b**	**1c**		**1a**	**1b**	**1c**
Natural Gas	54,442 (120,024)	54,639 (120,459)	54,442 (120,024)	**Stack Gas**	5,472 (12,063)	5,485 (12,093)	5,472 (12,063)
Air (CO_2)	275 (607)	213 (469)	275 (607)	**CO_2 Product**	49,245 (108,567)	49,367 (108,836)	49,245 (108,567)
Total	54,717 (120,631)	54,852 (120,928)	54,717 (120,631)	**Total**	54,717 (120,631)	54,852 (120,928)	54,717 (120,631)

Exhibit 3-24 Cases 1a, 1b, and 1c Water Balances

Water Use	Water Demand, m^3/min (gpm)			Internal Recycle, m^3/min (gpm)			Raw Water Withdrawal, m^3/min (gpm)			Process Water Discharge, m^3/min (gpm)			Raw Water Consumption, m^3/min (gpm)		
	1a	**1b**	**1c**	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**
Econamine	0.04 (12)	0.04 (12)	0.04 (12)	0 (0)	0 (0)	0 (0)	0.04 (12)	0.04 (12)	0.04 (12)	0 (0)	0 (0)	0 (0)	0.04 (12)	0.04 (12)	0.04 (12)
Condenser Makeup	0.09 (23)	0.09 (23)	0.09 (23)	0 (0)	0 (0)	0 (0)	0.09 (23)	0.09 (23)	0.09 (23)	0 (0)	0 (0)	0 (0)	0.09 (23)	0.09 (23)	0.09 (23)
BFW Makeup	0.09 (23)	0.09 (23)	0.09 (23)	0 (0)	0 (0)	0 (0)	0.09 (23)	0.09 (23)	0.09 (23)				0 (0)	0 (0)	0 (0)
Cooling Tower	16.01 (4,229)	16.16 (4,270)	16.78 (4,432)	1.9 (510)	2.1 (564)	1.9 (510)	14.08 (3,719)	14.0 (3,706)	14.84 (3,921)	3.6 (951)	3.6 (960)	3.8 (997)	10.48 (2,767)	10.4 (2,746)	11.07 (2,925)
BFW Blowdown	0 (0)	0 (0)	0 (0)	0.1 (23)	0.09 (23)	0.1 (23)	-0.09 (-23)	-0.09 (-23)	-0.09 (-23)				0 (0)	0 (0)	0 (0)
Flue Gas Condensate	0 (0)	0 (0)	0 (0)	1.8 (488)	2.0 (541)	1.8 (488)	-1.85 (-488)	-2.0 (-541)	-1.85 (-488)				0 (0)	0 (0)	0 (0)
CO_2 Product Condensate	0 (0)	0 (0)	0 (0)	0 (3)	0 (2)	0 (3)	-0.01 (-3)	-0 (-2)	-0.01 (-3)				0 (0)	0 (0)	0 (0)
Total	**16.1 (4,251)**	**16.3 (4,293)**	**16.9 (4,454)**	**1.93 (510)**	**2.13 (564)**	**1.93 (510)**	**14.2 (3,741)**	**14.1 (3,729)**	**14.9 (3,944)**	**3.6 (951)**	**3.6 (960)**	**3.8 (997)**	**10.61 (2,802)**	**10.5 (2,781)**	**11.20 (2,959)**

Exhibit 3-25 Cases 1a, 1b, and 1c Overall Energy Balances

	HHV			Sensible + Latent			Power			Total		
Case	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**	**1a**	**1b**	**1c**
Heat In GJ/hr (MMBtu/hr)												
Natural Gas	3,953 (3,747)	3,968 (3,761)	3,953 (3,747)	2.6 (2.5)	2.7 (2.5)	2.6 (2.5)				3,956 (3,750)	3,970 (3,763)	3,956 (3,750)
GT Air				61.0 (57.8)	47.1 (44.7)	61.0 (57.8)				61 (58)	47 (45)	61 (58)
Raw Water Makeup				53.3 (50.5)	53.1 (50.3)	56.1 (53.2)				53 (50)	53 (50)	56 (53)
Auxiliary Power							138 (130)	129 (123)	139 (132)	138 (130)	129 (123)	139 (132)
TOTAL	**3,953 (3,747)**	**3,968 (3,761)**	**3,953 (3,747)**	**116.9 (110.8)**	**102.9 (97.5)**	**119.7 (113.5)**	**138 (130)**	**129 (123)**	**139 (132)**	**4,208 (3,988)**	**4,200 (3,981)**	**4,212 (3,992)**
Heat Out GJ/hr (MMBtu/hr)												
CO_2				-29.8 (-28.2)	-29.8 (-28.3)	-29.8 (-28.2)				-30 (-28)	-30 (-28)	-30 (-28)
Cooling Tower Blowdown				26.7 (25.4)	27.0 (25.6)	28.0 (26.6)				27 (25)	27 (26)	28 (27)
Combustion Turbine Heat Loss				54.9 (52.0)	54.9 (52.0)	54.9 (52.0)				55 (52)	55 (52)	55 (52)
Recycle/Econamine Precooler				636.6 (603.3)	661.6 (627.1)	615.4 (583.3)				637 (603)	662 (627)	615 (583)
Econamine Losses				726.2 (688.3)	712.0 (674.9)	614.2 (582.2)				726 (688)	712 (675)	614 (582)
CO_2 Compression Intercooling				83.6 (79.2)	83.8 (79.4)	83.6 (79.2)				84 (79)	84 (79)	84 (79)
Stack Gas				167 (158)	124 (118)	167 (158)				167 (158)	124 (118)	167 (158)
Condenser				549 (520)	569 (540)	649 (615)				549 (520)	569 (540)	649 (615)
Process Losses*				147 (139)	141 (134)	151 (143)				147 (139)	141 (134)	151 (143)
Power							1,847 (1,750)	1,856 (1,759)	1,878 (1,780)	1,847 (1,750)	1,856 (1,759)	1,878 (1,780)
TOTAL				**2,361 (2,238)**	**2,344 (2,222)**	**2,333 (2,212)**	**1,847 (1,750)**	**1,856 (1,759)**	**1,878 (1,780)**	**4,208 (3,988)**	**4,200 (3,981)**	**4,212 (3,992)**

*Process losses including steam turbine, combustion reactions, HRSG, gas turbine, and gas cooling are estimated to match the heat input to the plant.

3.3 CASE 2-PRE-COMBUSTION AUTO-THERMAL REFORMING WITH AMINE ABSORPTION CO_2 SEPARATION

This case is an NGCC plant with an auto-thermal reformer and pre-combustion CO_2 capture via amine absorption. Auto thermal reforming is partial oxidation followed by thermal and catalytic-steam reforming.

The process is shown by the following reactions: [10,11]

Partial oxidation:

$$CH_4 + \tfrac{1}{2} O_2 \leftrightarrow CO + 2H_2$$

$$C_nH_{2n+2} + \frac{n}{2} O_2 \leftrightarrow nCO + (n+1)H_2$$

Reforming process:

$$CH_4 + H_2O \leftrightarrow CO + 3H_2$$

$$C_nH_{2n+2} + nH_2O \leftrightarrow nCO + (2n+1)H_2$$

$$C_nH_{2n+2} + 2nH_2O \leftrightarrow nCO_2 + (2n+2)H_2$$

Water Gas Shift (WGS):

$$CO + H_2O \leftrightarrow CO_2 + H_2 \text{ (CO shift reaction)}$$

Where n = an integer

The system consists of two gas turbines, two HRSGs, and one steam turbine. It was modeled after an air-fired Integrated Gasification Combined Cycle (IGCC) system because the ATR acts like a gasifier in an IGCC system. The amine system used is methyl diethanolamine (MDEA). To improve the kinetics of the CO_2 reaction with MDEA, activators, such as piperazine, are mixed with the MDEA. To predict the performance of an MDEA system, use of Optimized Gas Treating’s (OGT) ProTreat software was attempted, but found to be inadequate. Performance predictions using ProTreat with MDEA alone were not in line with published results. After several attempts to get OGT to resolve the problem, the MDEA system performance based on an IEA publication was used [12]. Values used include CO_2 removal of 91.7 percent, reboiler steam of 198 Btu/lb acid gas, cooling water of 8.30 lb H_2O/lb acid gas, and auxiliary load of 16.5 kWh/ton acid gas.

For this case, two published studies were used as a point of reference: Kvamsdal, et al. [11] and Corradetti, et al. [10]. Corradetti, et al. investigated three different plant variations. The difference between the three is the way the streams entering the reformer are preheated. The “Reference Case” is provided with only one heat exchanger. GT exhaust gases are used to preheat the natural gas and the steam. In the supplementary firing configuration, exhaust gas temperature is increased by supplementary firing and two heat exchangers are employed. For the gas-gas configuration, the heat exchangers utilize the hot syngas leaving the reformer. Therefore, this hot syngas preheats the natural gas and steam instead of the exhaust gas. Also they ran each variation at two different temperatures: 1,562°F and 1,742°F. The system described in their study consisted of no extra pressurization of the air to the ATR and a compressor added on the fuel stream to the GT. Kvamsdal et al. compared this system to three others: a base case which was a natural gas fired combined cycle plant operating at ISO conditions, using 59°Fcooling water and without CO_2 capture; another case that had a compressor on the air supplied to the ATR; and a final case that was like the case described in the study, but the medium pressure steam stream was split and partly supplied to both upstream of the reformer and downstream of the ATR.

Case 2's performance was based on the published data given in these two papers. A table comparing these two documents along with Case 2 and other studies is shown in Exhibit 3-26. A BFD and stream tables are shown in Exhibit 3-27 and Exhibit 3-28, respectively.

Exhibit 3-26 Case 2 Comparison with Other Studies

Case/Study/ System	AGR Type	CO_2 Capture	Efficiency % LHV	Reboiler Duty (Btu/lb$_{CO2}$)	Lower Heating Value (Btu/scf)	CT	Air Extraction
Case 2	MDEA	89.6%	47.0%	217	141	Advanced F-Class	Yes
Kvamsdal (several studies)	N/A	90.5%	46%-47%	Not Specified	157	GE 7FA	Yes
Corradetti (several studies)	MDEA with DEA as activator	92%-97%	47%-48%	356-423	152	Siemens V94.3A	Yes
Ref2	MEA	90.7%	46.8%	1,605	932	Advanced F-Class	No

Because the methane reforming reaction requires steam and is endothermic, conversion is optimal at temperatures between 1,292-1,742°F. Methane conversion is enhanced by excess steam and high temperatures.

The temperature must be controlled in order to achieve the carbon capture (90%) desired. To maintain values close to these numbers, the ATR is operated at 1,742°F. The CO_2 capture achieved through the process is 89.6 percent.

The system is air fired so there is nitrogen mixed with the fuel gas. This gives a fuel gas lower heating value of 141 Btu/scf. For this case, no humidification was used to reduce the heating value because the heating values of the fuel gas from the published studies [10, 11] are 152 Btu/scf and 157 Btu/scf, respectively, and neither case provided additional dilution.

The combustion turbine model used for this study is based on the Advanced F-Class. The Advanced F-Class CT requires fuel at ~1.5 – 2.0 times the compressor exhaust pressure, and 450 psia was used in this case. This number was based on published performance data. This combustion turbine is used for all the NETL baseline studies.

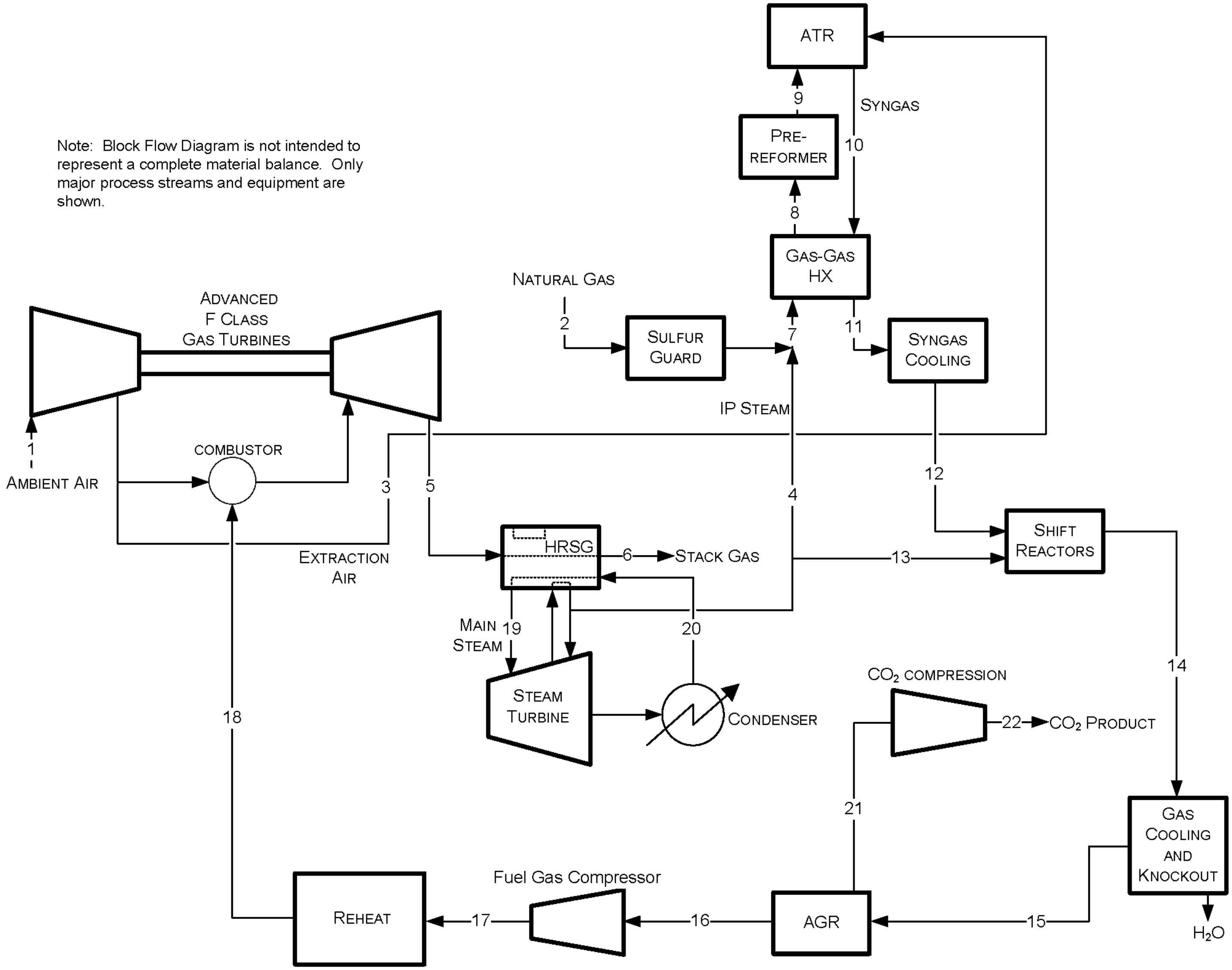
The study by Kvamsdal, et al. [11] is based on a GE 7FA which is older and smaller than the Advanced F-Class and requires a fuel pressure of only 300 psia. The study by Corradetti, et al. [10] uses a Siemens V94.3A combustion turbine but no fuel pressure is listed.

For this study, extraction air for the ATR is 16 percent of the CT compressor capacity. This amount is somewhat higher when compared to the maximum normally allowed for a 7F combustion turbine but was assumed to be within possible limits for advanced turbines operating with hydrogen rich fuels. The ATR outlet stream is at a pressure of 253 psia. After the water gas shift and the MDEA carbon dioxide removal sections, the clean hydrogen rich fuel gas stream is at a pressure of 178 psia. In order to provide fuel gas at the necessary pressure for the CT combustor, a fuel compressor was added to the fuel gas entering the CT to increase the pressure to 450 psia.

It is worth noting that the two published studies obtained all of the ATR air from CT compressor extraction. Corradetti [10] pressurized the fuel gas to achieve the required pressure while Kvamsdal [11] looked at both boosting the air pressure to the ATR and boosting the fuel gas pressure. Boosting the fuel gas pressure resulted in an efficiency of 1.2 absolute percentage points higher efficiency than boosting the extraction air for combustion. Kvamsdal used 13.2 percent extraction, which is close to the design value used in this study.

Comparing the overall plant efficiency of this study to the published studies reveals similar results. The efficiency in this study is 47.0 percent (LHV basis), whereas published data ranges from 46-48 percent.

Exhibit 3-27 Case 2 Block Flow Diagram, Auto-Thermal Reforming and Amine CO_2 Separation



Note: Actual process consists of 2 Advanced F-class CTGs, 2 HRSGs, and 1 STG in a multi-shaft 2x2x1 configuration

Exhibit 3-28 Case 2 Stream Table, Auto-Thermal Reforming and Amine CO_2 Separation

	1	2	3	4	5	6	7	8	9	10
V-L Mole Fraction										
H_2O	0.0099	0.0000	0.0099	1.0000	0.1389	0.1389	0.5494	0.5494	0.4766	0.1846
Ar	0.0092	0.0000	0.0092	0.0000	0.0089	0.0089	0.0000	0.0000	0.0000	0.0044
CO_2	0.0003	0.0100	0.0003	0.0000	0.0055	0.0055	0.0045	0.0045	0.0275	0.0461
O_2	0.2074	0.0000	0.2074	0.0000	0.1016	0.1016	0.0000	0.0000	0.0000	0.0000
N_2	0.7732	0.0160	0.7732	0.0000	0.7451	0.7451	0.0072	0.0072	0.0067	0.3673
CH_4	0.0000	0.9310	0.0000	0.0000	0.0000	0.0000	0.4195	0.4195	0.4194	0.0011
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.1057
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0690	0.2904
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003
C_2H_6	0.0000	0.0320	0.0000	0.0000	0.0000	0.0000	0.0144	0.0144	0.0000	0.0000
C_3H_8	0.0000	0.0070	0.0000	0.0000	0.0000	0.0000	0.0032	0.0032	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0000	0.0000	0.0000	0.0018	0.0018	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	109,323	5,481	17,647	6,682	113,477	113,477	12,163	12,163	12,764	37,367
V-L Flowrate (kg/hr)	3,154,735	94,971	509,228	120,377	3,089,303	3,089,303	215,348	215,348	215,348	724,576
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	38	417	515	632	105	278	482	425	950
Pressure (MPa, abs)	0.10	3.10	1.84	1.84	0.11	0.10	1.84	1.84	1.81	1.74
Enthalpy (kJ/kg)[A]	30.23	46.30	447.65	3,499.49	955.35	344.04	1,977.17	2,519.40	2,255.58	2,043.22
Density (kg/m^3)	1.2	22.2	9.2	5.2	0.4	0.9	7.3	5.2	5.3	3.3
V-L Molecular Weight	28.857	17.328	28.857	18.015	27.224	27.224	17.705	17.705	16.872	19.391
V-L Flowrate (lb_{mol}/hr)	241,016	12,083	38,904	14,731	250,175	250,175	26,815	26,815	28,140	82,379
V-L Flowrate (lb/hr)	6,955,000	209,375	1,122,656	265,386	6,810,747	6,810,747	474,760	474,760	474,760	1,597,416
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	100	783	959	1,169	220	532	900	797	1,742
Pressure (psia)	14.7	450.0	267.5	267.5	15.2	14.7	267.5	267.5	262.5	252.5
Enthalpy (Btu/lb)[A]	13.0	19.9	192.5	1,504.5	410.7	147.9	850.0	1,083.1	969.7	878.4
Density (lb/ft^3)	0.076	1.384	0.576	0.322	0.024	0.055	0.456	0.327	0.330	0.207

Note: Flows shown are totals for all process trains

Exhibit 3-28 Case 2 Stream Table, Auto-Thermal Reforming and Amine CO_2 Separation (continued)

	11	12	13	14	15	16	17	18	19	20	21	22
V-L Mole Fraction												
H_2O	0.1846	0.1846	1.0000	0.1065	0.0047	0.0044	0.0036	0.0036	1.0000	1.0000	0.0062	0.0000
Ar	0.0044	0.0044	0.0000	0.0043	0.0047	0.0055	0.0055	0.0055	0.0000	0.0000	0.0004	0.0000
CO_2	0.0461	0.0461	0.0000	0.1444	0.1609	0.0144	0.0144	0.0144	0.0000	0.0000	0.9821	1.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.3673	0.3673	0.0000	0.3577	0.3986	0.4678	0.4682	0.4682	0.0000	0.0000	0.0104	0.0000
CH_4	0.0011	0.0011	0.0000	0.0011	0.0012	0.0014	0.0014	0.0014	0.0000	0.0000	0.0002	0.0000
CO	0.1057	0.1057	0.0000	0.0035	0.0039	0.0045	0.0045	0.0045	0.0000	0.0000	0.0004	0.0000
H_2	0.2904	0.2904	0.0000	0.3823	0.4260	0.5019	0.5023	0.5023	0.0000	0.0000	0.0003	0.0000
NH_3	0.0003	0.0003	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	37,367	37,367	1,003	38,370	34,436	29,223	29,200	29,200	34,254	41,630	5,213	5,120
V-L Flowrate (kg/hr)	724,576	724,576	18,069	742,645	671,783	444,216	443,796	443,796	617,098	749,975	227,567	225,312
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	863	232	515	222	36	31	134	302	566	38	16	50
Pressure (MPa, abs)	1.71	1.60	1.84	1.43	1.29	1.22	3.10	3.07	16.65	0.01	0.14	15.27
Enthalpy (kJ/kg)[A]	1,882.06	792.92	3,499.49	604.82	61.44	65.80	267.81	596.89	3,472.33	160.62	17.96	-169.59
Density (kg/m^3)	3.5	7.4	5.2	6.7	9.8	7.4	13.8	9.7	47.7	992.9	2.5	667.3
V-L Molecular Weight	19.391	19.391	18.015	19.355	19.508	15.201	15.198	15.198	18.015	18.015	43.657	44.010
V-L Flowrate (lb_{mol}/hr)	82,379	82,379	2,211	84,591	75,918	64,426	64,375	64,375	75,518	91,778	11,492	11,287
V-L Flowrate (lb/hr)	1,597,416	1,597,416	39,836	1,637,253	1,481,028	979,329	978,403	978,403	1,360,469	1,653,412	501,699	496,727
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	1,585	450	959	431	96	87	274	575	1,050	101	60	121
Pressure (psia)	247.5	232.5	267.5	207.5	187.5	177.5	450.0	445.0	2,415.0	1.0	20.0	2,214.7
Enthalpy (Btu/lb)[A]	809.1	340.9	1,504.5	260.0	26.4	28.3	115.1	256.6	1,492.8	69.1	7.7	-72.9
Density (lb/ft^3)	0.218	0.462	0.322	0.419	0.614	0.459	0.859	0.603	2.977	61.982	0.158	41.660

Note: Flows shown are totals for all process trains

The auto thermal reformer plant produces a net output of 586 MW at a net plant efficiency of 42.4 percent (HHV basis). Overall plant performance is summarized in Exhibit 3-29. The summary includes auxiliary power requirements. The steam turbine power is higher than the reference case due to additional steam available from the shift reactors (interstage heat recovery) and other sources.

Exhibit 3-29 Case 2 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	384,227
Steam Turbine Power	260,973
TOTAL POWER, kWe	**645,200**
AUXILIARY LOAD SUMMARY, kWe	
Fuel Gas Compression	25,680
CO_2 Compression	19,110
Condensate Pumps	250
Boiler Feedwater Pumps	4,220
Amine System (MDEA) Auxiliaries	1,000
Circulating Water Pump	3,430
Ground Water Pumps	400
Cooling Tower Fans	1,770
SCR	10
Gas Turbine Auxiliaries	700
Steam Turbine Auxiliaries	100
Miscellaneous Balance of Plant[1]	500
Transformer Losses	2,030
TOTAL AUXILIARIES, kWe	**59,200**
Plant Performance	
NET POWER, kWe	**586,000**
Net Plant Efficiency (HHV)	42.4%
Net Plant Efficiency (LHV)	47.0%
Net Plant Heat Rate (HHV), kJ/kWh (Btu/kWh)	8,500 (8,057)
Net Plant Heat Rate (LHV), kJ/kWh (Btu/kWh)	7,664 (7,264)
CONDENSER COOLING DUTY, GJ/hr (10^6 Btu/hr)	1,340 (1,270)
CONSUMABLES	
Natural Gas Feed Flow, kg/hr (lb/hr)	94,971 (209,375)
Thermal Input (HHV), kW_{th}	1,383,644
Thermal Input (LHV) , kW_{th}	1,247,533
Raw Water Withdrawal, m^3/min (gpm)	16.8 (4,430)
Raw Water Consumption, m^3/min (gpm)	13.8 (3,638)
CO_2 Capture	89.6%
CO_2 emissions tonne/yr (ton/yr)[2]	204,492 (225,414)
CO_2 emissions kg/MWh (lb/MWh)	43 (94)

1 Includes plant control systems, lighting, HVAC and miscellaneous low voltage loads
2 Based on 85% capacity factor

The estimated air emissions are shown in Exhibit 3-30. Operation of the modern, state-of-the-art gas turbine fueled by natural gas, coupled to a HRSG, results in very low NOx emissions and negligible amounts of particulate and SO_2. There are no mercury emissions in an NGCC plant.

The low level of NOx production (2.5 ppmvd at 15 percent O_2) is achieved by utilizing Selective Catalytic Reduction (SCR).

Exhibit 3-30 Case 2 Estimated Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (ton/year) 85% capacity factor	kg/MWh (lb/MWh)
SO_2	Negligible	Negligible	Negligible
NOx	0.021 (0.048)	770 (848)	0.160 (0.353)
Particulate	Negligible	Negligible	Negligible
Hg	Negligible	Negligible	Negligible
CO_2	5.5 (12.8)	204,492 (225,414)	43 (94)
CO_2 net[1]			47 (103)

[1] CO_2 emissions based on net power instead of gross power

The carbon balance is shown in Exhibit 3-31. The carbon input to the plant consists of carbon in the air and the carbon in the natural gas. Carbon leaves the plant as CO_2 in the stack gas and CO_2 product. The percent of total carbon sequestered is defined as the amount of carbon product produced divided by the carbon in the natural gas feedstock, expressed as a percentage.

% Captured = Carbon in Product for Sequestration / Carbon in the Natural gas

or

61,491/68,596 *100 = 89.6%

Exhibit 3-31 Case 2 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	68,596 (151,229)	**Stack Gas**	7,495 (16,524)
Air (CO_2)	498 (1,099)	**CO_2 Product**	61,491 (135,565)
		Convergence Tolerance*	108 (238)
Total	69,095 (152,327)	**Total**	69,095 (152,327)

*by difference

An overall water balance for the plant is shown in Exhibit 3-32. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process and is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

Exhibit 3-32 Case 2 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Condenser Makeup	4.7 (1,253)	0 (0)	4.7 (1,253)	0 (0)	4.7 (1,253)
Shift Steam	0.3 (80)	0 (0)	0.3 (80)		
ATR Steam	2.01 (531)	0 (0)	2.01 (531)		
BFW Makeup	2.43 (643)	0 (0)	2.43 (643)		
Cooling Tower	13.3 (3,523)	1.3 (346)	12.0 (3,177)	3.0 (792)	9.0 (2,385)
BFW Blowdown	0 (0)	0.12 (32)	-0.12 (-32)		
Flue Gas Condensate	0 (0)	1.18 (312)	-1.18 (-312)		
CO_2 Product Condensate	0 (0)	0.01 (2)	-0.01 (-2)		
Total	**18.1 (4,776)**	**1.3 (346)**	**16.8 (4,430)**	**3.0 (792)**	**13.8 (3,638)**

An overall plant energy balance is provided in tabular form in Exhibit 3-33. The power out is the combined combustion turbine and steam turbine power after generator losses.

Exhibit 3-33 Case 2 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Natural Gas	4,981 (4,721)	3.3 (3.2)		4,984 (4,724)
Ambient Air		323.3 (306.4)		323 (306)
Raw Water Makeup		63.1 (59.8)		63 (60)
Auxiliary Power			213 (202)	213 (202)
TOTAL	**4,981 (4,721)**	**389.7 (369.4)**	**213 (202)**	**5,584 (5,293)**
Heat Out GJ/hr (MMBtu/hr)				
CO_2		-38.2 (-36.2)		-38 (-36)
Cooling Tower Blowdown		22.3 (21.1)		22 (21)
Combustion Turbine Heat Loss		54.9 (52.0)		55 (52)
Stack Gas		1,063 (1,007)		1,063 (1,007)
Condenser		1,345 (1,274)		1,345 (1,274)
Non-Condenser Cooling Tower Loads		284 (269)		284 (269)
Process Losses*		531 (503)		531 (503)
Power			2,323 (2,202)	2,323 (2,202)
TOTAL		**3,261 (3,091)**	**2,323 (2,202)**	**5,584 (5,293)**

*Process losses including steam turbine, combustion reactions, HRSG, gas turbine, and gas cooling are estimated to match the heat input to the plant.

3.4 CASE 3-PRE-COMBUSTION PARTIAL OXIDATION WITH AMINE ABSORPTION CO_2 SEPARATION

This case is an NGCC plant with high pressure partial oxidation (POX) and pre-combustion CO_2 capture via amine absorption. The POX reactor consists of fast exothermic partial oxidation reactions and slow endothermic steam reforming reactions. The reactions are [10, 11]:

Partial oxidation:

$$CH_4 + \tfrac{1}{2} O_2 \leftrightarrow CO + 2\ H_2$$

$$C_nH_{2n+2} + \frac{n}{2} O_2 \leftrightarrow nCO + (n+1)H_2$$

Water Gas Shift:

$$CO + H_2O \leftrightarrow CO_2 + H_2 \text{ (CO shift reaction)}$$

Where n = an integer

This case is similar to Case 2. The system consists of two gas turbines, two HRSGs, and one steam turbine. The amine system used is methyl diethanolamine (MDEA). A BFD and stream tables are shown in Exhibit 3-34 and Exhibit 3-35, respectively.

A study by GE Global (Hoffman et al.) [13] introduces two novel cycles that include a high-pressure syngas generation island where an air-blown, partial oxidation reformer is used to generate syngas from natural gas. CO_2 is removed from the shifted syngas using either CO_2 absorbing solvents or a CO_2 membrane.

Since the POX reaction is less pressure dependent than steam-methane reforming reaction, the standard POX reformer can be used at very high pressures without excessive methane-slip. Operation at high-pressure allows smaller downstream units and higher driving forces for CO_2 removal [13]. The POX reactor in this study is operated at 2,102°F and 1,405 psia, which is consistent with Hoffman, et al.'s operating pressure range of 1,160-1,450 psia and operating temperature of 2,102°F. Since natural gas is delivered to the plant at a pipeline pressure of 450 psia, a compressor is used to increase the fuel pressure to 1,430 psia prior to the POX reactor.

Steam to the POX reactor must be controlled to achieve a specified steam to carbon ratio (0.1) and carbon capture level (90 percent). Since there was no published data showing how the steam ratio is affected using a POX, various steam ratios were tested until the amount of methane in the syngas stream was less than one percent. The CO_2 capture achieved through the process is 88.9 percent. The systems in the Hoffman, et al. [13] study were designed to separate 80 percent of the overall CO_2 produced in the cycle.

For this study, extraction air for the POX is 20 percent of the CT compressor capacity. Since the extraction air flow is higher than normally allowed for the Advanced F-Class combustion turbine, it was assumed that the CT could be redesigned to accommodate 20 percent extraction air.

After CO_2 is removed using the MDEA process the fuel gas has a heating value of 129 Btu/scf (LHV), which is in an acceptable range for the CT. However the mass flow rate is insufficient to fully load the F-class turbine, so additional dilution is achieved through humidification. The humidified fuel gas stream has a heating value of 111 Btu/scf (LHV).

Before entering the syngas quench, an expander was used on the outlet stream of the POX reactor to lower the pressure from 1,405 psia to 500 psia to accommodate the CT fuel gas pressure requirement and to recover 98 MW of power. Hoffman, et al. used a low pressure expander (inlet temperature of 644°F) to recover the energy in the syngas prior to combustion.

Comparing overall plant efficiencies of this study to other published studies reveals a slight difference. The efficiency in this study is 45.8 percent (LHV basis); whereas the published value is 47.5 percent (LHV) with a solvent based CO_2 separation. The difference is attributed to the following:

- Cooling Water

The Hoffman study assumes 46°F cooling water (seawater) compared to 60°F assumed in case 3 [13]. The lower water temperature in the Hoffman study would result in a lower condenser pressure and hence greater energy recovery in the steam turbine.

- Steam conditions

The steam conditions used in this study were 1,050°F and 2,400 psia. Hoffman, et al. [13] did not provide steam conditions for their study. A possible difference in steam conditions could also contribute to the efficiency difference.

- Amount of compressor power

The power required for the natural gas compressor and air compressor are significant, 43,620 kW. The power requirements for the natural gas compressor and air compressor are 6,310 kW and 37,300 kW, respectively, which accounts for 56 percent of the total auxiliary loads. If these values were zero, the efficiency (LHV basis) would increase from 45.8 percent to 48.9 percent (both LHV basis). In the Hoffman study, the natural gas supply pressure was 725 psi compared to 450 psia used in this study. The higher natural gas supply pressure would result in a 0.2 percentage point increase in overall net efficiency (45.8 to 46.0 percent, LHV basis). Compressor efficiency was not provided in the Hoffman study and could potentially account for more of the efficiency difference.

- Combustion Turbine Frame Size

The Hoffman study used a 9FB gas turbine in the power island while the current study uses a 7FB. The larger 9FB machine results in 0.5 to 0.75 percentage point higher efficiency using natural gas in a simple cycle or combined cycle configuration.

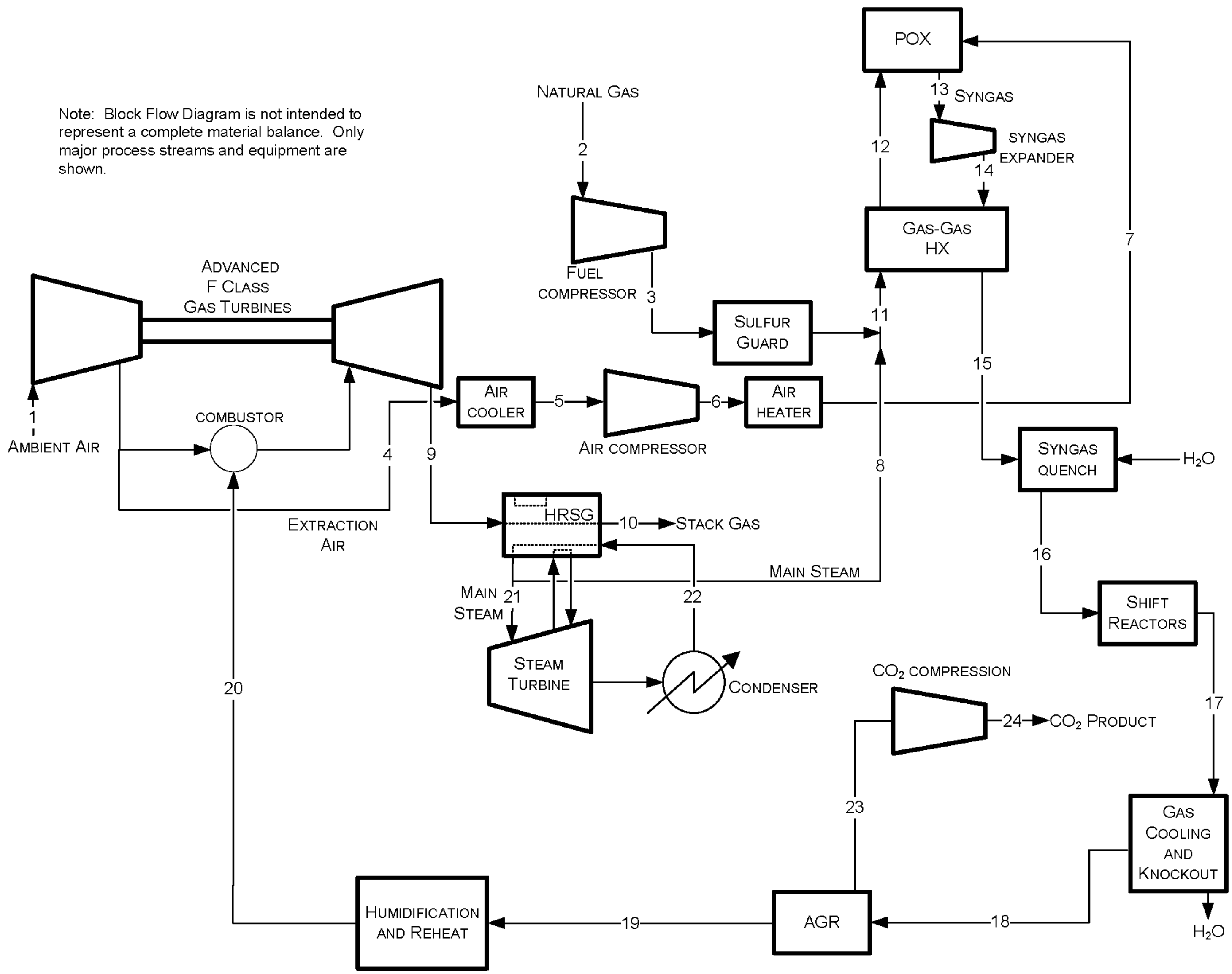
- Type of solvent

In Hoffman, et al.'s study, there were two types of CO_2 separation methods used: a solvent based CO_2 separation unit and a CO_2 membrane. The most efficient method was the CO_2 membrane. The net efficiency (LHV basis) was 50.7 percent compared to 47.5 percent when chemical absorption was used. However, the type of chemical solvent used was not specified.

For this study, the solvent used was MDEA. Alternative solvents may impact efficiency, but investigation of alternatives was beyond the scope of this study.

The high pressure partial oxidation plant produces a net output of 650 MW at a net plant efficiency of 41.3 percent (HHV basis). Overall plant performance is summarized in Exhibit 3-36. The summary includes auxiliary power requirements.

Exhibit 3-34 Case 3 Block Flow Diagram, Partial Oxidation and Amine CO_2 Separation



Note: Actual process consists of 2 Advanced F-class CTGs, 2 HRSGs, and 1 STG in a multi-shaft 2x2x1 configuration

Exhibit 3-35 Case 3 Stream Tables, Partial Oxidation and Amine CO_2 Separation

	1	2	3	4	5	6	7	8	9	10	11	12
V-L Mole Fraction												
H_2O	0.0099	0.0000	0.0000	0.0099	0.0043	0.0015	0.0015	1.0000	0.1899	0.1899	0.0888	0.0888
Ar	0.0092	0.0000	0.0000	0.0092	0.0093	0.0093	0.0093	0.0000	0.0085	0.0085	0.0000	0.0000
CO_2	0.0003	0.0100	0.0100	0.0003	0.0003	0.0003	0.0003	0.0000	0.0064	0.0064	0.0091	0.0091
O_2	0.2074	0.0000	0.0000	0.2074	0.2086	0.2092	0.2092	0.0000	0.0842	0.0842	0.0000	0.0000
N_2	0.7732	0.0160	0.0160	0.7732	0.7775	0.7797	0.7797	0.0000	0.7110	0.7110	0.0146	0.0146
CH_4	0.0000	0.9310	0.9310	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.8483	0.8483
CO	0.0000	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0320	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0292	0.0292
C_3H_8	0.0000	0.0070	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0064	0.0064
C_4H_{10}	0.0000	0.0040	0.0040	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0036	0.0036
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	109,323	6,234	6,234	22,371	22,245	22,183	22,183	608	118,906	118,906	6,842	6,842
V-L Flowrate (kg/hr)	3,154,735	108,022	108,022	645,547	643,278	642,160	642,160	10,948	3,169,251	3,169,251	118,970	118,970
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	38	89	417	38	104	371	482	614	121	142	427
Pressure (MPa, abs)	0.10	3.10	9.86	1.84	1.74	10.00	9.94	10.00	0.11	0.10	9.79	9.76
Enthalpy (kJ/kg)[A]	30.23	46.30	118.78	447.65	40.22	93.72	383.24	3,316.88	1,046.92	457.33	412.74	1,343.80
Density (kg/m^3)	1.2	22.2	62.0	9.2	19.6	90.9	52.0	31.9	0.4	0.8	54.2	28.8
V-L Molecular Weight	28.857	17.328	17.328	28.857	28.918	28.949	28.949	18.015	26.653	26.653	17.389	17.389
V-L Flowrate (lb_{mol}/hr)	241,016	13,744	13,744	49,319	49,041	48,904	48,904	1,340	262,143	262,143	15,084	15,084
V-L Flowrate (lb/hr)	6,955,000	238,148	238,148	1,423,186	1,418,185	1,415,720	1,415,720	24,137	6,987,003	6,987,003	262,285	262,285
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	100	193	783	100	220	700	900	1,137	250	288	800
Pressure (psia)	14.7	450.0	1,430.0	267.5	252.5	1,450.0	1,442.0	1,450.0	15.2	14.7	1,420.0	1,415.0
Enthalpy (Btu/lb)[A]	13.0	19.9	51.1	192.5	17.3	40.3	164.8	1,426.0	450.1	196.6	177.4	577.7
Density (lb/ft^3)	0.076	1.384	3.870	0.576	1.224	5.676	3.248	1.993	0.024	0.051	3.383	1.801

Note: Flows shown are totals for all process trains

Exhibit 3-35 Case 3 Stream Tables, Partial Oxidation and Amine CO_2 Separation (continued)

	13	14	15	16	17	18	19	20	21	22	23	24
V-L Mole Fraction												
H_2O	0.0800	0.0800	0.0800	0.2485	0.1269	0.0110	0.0104	0.1534	1.0000	1.0000	0.0148	0.0000
Ar	0.0056	0.0056	0.0056	0.0046	0.0046	0.0052	0.0060	0.0051	0.0000	0.0000	0.0004	0.0005
CO_2	0.0202	0.0202	0.0202	0.0165	0.1381	0.1565	0.0140	0.0120	0.0000	0.0000	0.9720	0.9866
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.4707	0.4707	0.4707	0.3845	0.3845	0.4358	0.5100	0.4363	0.0000	0.0000	0.0115	0.0117
CH_4	0.0040	0.0040	0.0040	0.0033	0.0033	0.0037	0.0042	0.0036	0.0000	0.0000	0.0007	0.0007
CO	0.1519	0.1519	0.1519	0.1241	0.0025	0.0028	0.0033	0.0028	0.0000	0.0000	0.0003	0.0003
H_2	0.2669	0.2669	0.2669	0.2180	0.3396	0.3849	0.4522	0.3868	0.0000	0.0000	0.0003	0.0003
NH_3	0.0007	0.0007	0.0007	0.0006	0.0006	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	36,931	36,931	36,931	45,207	45,207	39,883	33,948	39,684	17,856	27,956	5,935	5,847
V-L Flowrate (kg/hr)	761,130	761,130	761,130	910,235	910,235	814,326	556,731	660,063	321,679	503,636	257,595	256,010
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	1,150	881	594	232	220	35	31	342	566	38	21	53
Pressure (MPa, abs)	9.69	3.45	3.41	3.34	3.17	3.03	2.96	2.79	16.65	0.01	0.14	15.27
Enthalpy (kJ/kg)[A]	1,993.12	1,529.83	1,064.11	899.89	621.70	47.62	53.31	1,038.36	3,472.33	160.61	31.72	-153.69
Density (kg/m^3)	16.6	7.4	9.7	16.1	15.5	24.5	19.3	9.0	47.7	992.9	2.5	616.2
V-L Molecular Weight	20.610	20.610	20.610	20.135	20.135	20.418	16.399	16.633	18.015	18.015	43.404	43.786
V-L Flowrate (lb_{mol}/hr)	81,418	81,418	81,418	99,665	99,665	87,927	74,843	87,488	39,365	61,633	13,084	12,890
V-L Flowrate (lb/hr)	1,678,005	1,678,005	1,678,005	2,006,726	2,006,726	1,795,280	1,227,381	1,455,189	709,180	1,110,327	567,899	564,406
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	2,102	1,617	1,101	450	428	95	87	648	1,050	101	70	127
Pressure (psia)	1,405.0	500.0	495.0	485.0	460.0	440.0	430.0	405.0	2,415.0	1.0	20.0	2,214.7
Enthalpy (Btu/lb)[A]	856.9	657.7	457.5	386.9	267.3	20.5	22.9	446.4	1,492.8	69.1	13.6	-66.1
Density (lb/ft^3)	1.036	0.459	0.604	1.004	0.969	1.530	1.205	0.563	2.977	61.982	0.154	38.470

Note: Flows shown are totals for all process trains

Exhibit 3-36 Case 3 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	456,776
Steam Turbine Power	172,422
Fuel Gas Expander	97,950
TOTAL POWER, kWe	**727,148**
AUXILIARY LOAD SUMMARY, kWe	
CO_2 Compression	21,820
Air Compression	37,300
Natural Gas Compression	6,310
Condensate Pumps	160
Boiler Feedwater Pumps	2,280
Amine System (MDEA) Auxiliaries	1,200
Circulating Water Pump	2,900
Ground Water Pumps	280
Cooling Tower Fans	1,500
SCR	10
Gas Turbine Auxiliaries	700
Steam Turbine Auxiliaries	100
Miscellaneous Balance of Plant[1]	500
Transformer Losses	2,390
TOTAL AUXILIARIES, kWe	**77,450**
Plant Performance	
NET POWER, kWe	**649,698**
Net Plant Efficiency (HHV)	41.3%
Net Plant Efficiency (LHV)	45.8%
Net Plant Heat Rate (HHV), kJ/kWh (Btu/kWh)	8,720 (8,265)
Net Plant Heat Rate (LHV), kJ/kWh (Btu/kWh)	7,863 (7,452)
CONDENSER COOLING DUTY, GJ/hr (10^6 Btu/hr)	1,087 (1,030)
CONSUMABLES	
Natural Gas Feed Flow, kg/hr (lb/hr)	108,022 (238,148)
Thermal Input (HHV), kW_{th}	1,573,791
Thermal Input (LHV) , kW_{th}	1,418,975
Raw Water Withdrawal, m^3/min (gpm)	14.2 (3,762)
Raw Water Consumption, m^3/min (gpm)	11.7 (3,091)
CO_2 Capture	88.9%
CO_2 emissions tonne/yr (ton/yr)[2]	247,961 (273,331)
CO_2 emissions kg/MWh (lb/MWh)	46 (101)

1 Includes plant control systems, lighting, HVAC and miscellaneous low voltage loads
2 Based on 85% capacity factor

The estimated air emissions are shown in Exhibit 3-37. Operation of the modern, state-of-the-art gas turbine fueled by natural gas, coupled to a HRSG, results in very low NOx emissions and negligible amounts of particulate and SO_2. There are no mercury emissions in an NGCC plant.

The low level of NOx production (2.5 ppmvd at 15 percent O_2) is achieved by utilizing Selective Catalytic Reduction (SCR).

Exhibit 3-37 Case 3 Estimated Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (ton/year) 85% capacity factor	kg/MWh (lb/MWh)
SO_2	Negligible	Negligible	Negligible
NOx	0.021 (0.048)	874 (964)	0.161 (0.356)
Particulate	Negligible	Negligible	Negligible
Hg	Negligible	Negligible	Negligible
CO_2	5.9 (13.7)	247,961 (273,331)	46 (101)
CO_2 net[1]			51 (113)

[1] CO_2 emissions based on net power instead of gross power

The carbon balance is shown in Exhibit 3-38. The carbon input to the plant consists of carbon in the air and the carbon in the natural gas. Carbon leaves the plant as CO_2 in the stack gas and CO_2 product. The percent of total carbon sequestered is defined as the amount of carbon product produced divided by the carbon in the natural gas feedstock, expressed as a percentage.

% Captured = Carbon in Product for Sequestration / Carbon in the Natural gas

or

69,355/78,023 *100 = 88.9%

Exhibit 3-38 Case 3 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	78,023 (172,012)	**Stack Gas**	9,088 (20,037)
Air (CO_2)	517 (1,139)	**CO_2 Product**	69,355 (152,901)
		Convergence Tolerance*	97 (214)
Total	78,540 (173,151)	**Total**	78,540 (173,151)

*by difference

An overall water balance for the plant is shown in Exhibit 3-39. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process and is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

Exhibit 3-39 Case 3 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Humidifier	1.8 (488)	0 (0)	1.8 (488)	0 (0)	1.8 (488)
Quench	2.5 (657)	0 (0)	2.5 (657)	0 (0)	2.5 (657)
Condenser Makeup	0.5 (122)	0 (0)	0.5 (122)	0 (0)	0.5 (122)
POX Steam	0.18 (48)	0 (0)	0.18 (48)		
BFW Makeup	0.28 (73)	0 (0)	0.28 (73)		
Cooling Tower	11.3 (2,981)	1.8 (487)	9.4 (2,495)	2.5 (670)	6.9 (1,824)
BFW Blowdown	0 (0)	0.09 (25)	-0.09 (-25)		
Flue Gas Condensate	0 (0)	1.72 (455)	-1.72 (-455)		
CO_2 Product Condensate	0 (0)	0.02 (6)	-0.02 (-6)		
Total	**16.1 (4,248)**	**1.8 (487)**	**14.2 (3,762)**	**2.5 (670)**	**11.7 (3,091)**

An overall plant energy balance is provided in tabular form in Exhibit 3-40. The power out is the combined combustion turbine and steam turbine power after generator losses.

Exhibit 3-40 Case 3 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Natural Gas	5,666 (5,370)	3.8 (3.6)		5,669 (5,374)
Ambient Air		384.3 (364.3)		384 (364)
Raw Water Makeup		44.2 (41.9)		44 (42)
Auxiliary Power			279 (264)	279 (264)
TOTAL	**5,666 (5,370)**	**432.3 (409.7)**	**279 (264)**	**6,377 (6,044)**
Heat Out GJ/hr (MMBtu/hr)				
CO_2		-39.3 (-37.3)		-39 (-37)
Cooling Tower Blowdown		18.9 (17.9)		19 (18)
Combustion Turbine Heat Loss		63.3 (60.0)		63 (60)
Stack Gas		1,449 (1,374)		1,449 (1,374)
Condenser		1,085 (1,028)		1,085 (1,028)
Non-Condenser Cooling Tower Loads		265 (251)		265 (251)
Process Losses*		917 (870)		917 (870)
Power			2,618 (2,481)	2,618 (2,481)
TOTAL		**3,759 (3,563)**	**2,618 (2,481)**	**6,377 (6,044)**

*Process losses including steam turbine, combustion reactions, HRSG, gas turbine, and gas cooling are estimated to match the heat input to the plant.

3.5 CASE 4 OXY-COMBUSTION WITH CO_2 RECYCLE

This case is an NGCC plant with oxy-combustion and CO_2 capture. It is modeled after Case 1, except the EGR was replaced with CO_2 recycle. The system consists of two gas turbines, two HRSGs, and one steam turbine. A BFD and stream tables are shown in Exhibit 3-41 and Exhibit 3-42, respectively. Since there is approximately two percent O_2 in the CO_2 product stream (stream 9), CO_2 purification may be needed because this result does not meet the currently accepted CO_2 pipeline specification. (This could be accomplished by adding a catalytic combustor using a small amount of natural gas prior to the CO_2 compression section).

For this case, two published studies were used as points of reference: Kvamsdal, et al., [14] and Lozza, et al. [15]. In the study completed by Kvamsdal, et al., various gas turbine cycles with CO_2 capture were explored. The main focus for this current NGCC case was the oxy-combustion cycle. Kvamsdal, et al.'s study supplied an exhaust gas containing mostly H_2O and CO_2 to the HRSG. A large fraction of the CO_2, 90 percent, was recycled back to the combustor in order to keep the TIT at a required level. The overall net efficiency was 47.0 percent (LHV). The study done by Lozza, et al. examined three types of CO_2 capture methods: pre-combustion, post-combustion, and oxy-combustion. The main focus for this case is the oxy-combustion cycle. A higher pressure ratio, 45 vs. 18.5 (used in other cases), was used to keep the temperature profile (i.e. TIT and TEX) similar to the 9FB turbine used in the other cases. Lozza, et al.'s study produces a net plant efficiency of 46.2 percent (LHV basis), 96.9 percent CO_2 capture, and negligible amounts of CO, unburned hydrocarbons, and NO_x.

For Case 4, the working fluid entering the CT is primarily CO_2 (due to the high recycle percentage of ~90 percent) and the pressure ratio across the CT is 45.5. Due to this pressure ratio and the working fluid thermodynamic characteristics a new CT would be required. Additionally, the O_2 supplied from the ASU needs to be compressed to this high level (~660 psia). The pressure ratio was based on the Lozza, et al. [15] study discussed above after determining that using a pressure ratio corresponding to a 7F frame would result in an exhaust temperature greater than 1,500°F. This notion will be discussed more in the paragraph below. To get the desired total CT net work, ~362 MWe, the amount of CO_2 recycle was varied. The turbine inlet temperature (TIT) was set to match the previous cases, TIT = 2,500°F, by adjusting the fuel flow rate while maintaining a fixed combustor heat loss. However, even with a high pressure ratio, the resulting exhaust temperature due to the primarily CO_2 working fluid was 1,280°F. To match the 7F CT exhaust temperature (~1,163 °F) it was estimated that the pressure ratio would need to be increased to ~66 which is beyond any practical design.

As discussed above, a study by Lozza, et al. [15] investigated the pressure ratio effect on cycle efficiency. The same temperature profile must be reproduced by means of a much higher pressure ratio in order to keep cycle efficiency close to conventional machines' efficiency. Lozza, et al. [15] used a pressure ratio of 44 versus 18 for a conventional machine. However, it was stated that a full re-design of the machine was required to achieve a high pressure ratio. The pressure ratio used for this study was 45.5 (pressure rise across air compressor). This high pressure ratio resulted in a CT outlet temperature of 1,281°F, which is too high for a conventional HRSG but used nonetheless. The CT outlet temperature in the Lozza, et al. [15] study was 1,137°F. The current study produces steam at a temperature of 1,200°F. A conventional (commercial) steam turbine (ST) can accommodate temperatures up to about 1,100°F. Because of the high temperature, the ST would have to be designed with high temperature alloys.

The CO_2 capture achieved through the process is greater than 99 percent, typical of oxy-combustion processes.

Alternatives for this study include:

- Two-stage compressor

  In this study, a one-stage compressor was used in the CT. A two-stage compressor, with an intercooler, could have been used to reduce the exhaust gas temperature. The compressor's first stage could have included a 3.1 pressure ratio, while the second stage incorporated the balance.

- Pressure ratio

  Instead of having such a high pressure ratio across the CT, the pressure ratio could have been maintained at 18.5, the ratio used in all other cases. If this smaller pressure ratio is used, the firing temperature would be reduced so the exhaust gas temperature is approximately 1,100°F. This option was not considered in this study because it would derate the CT, thus causing a decrease in overall efficiency.

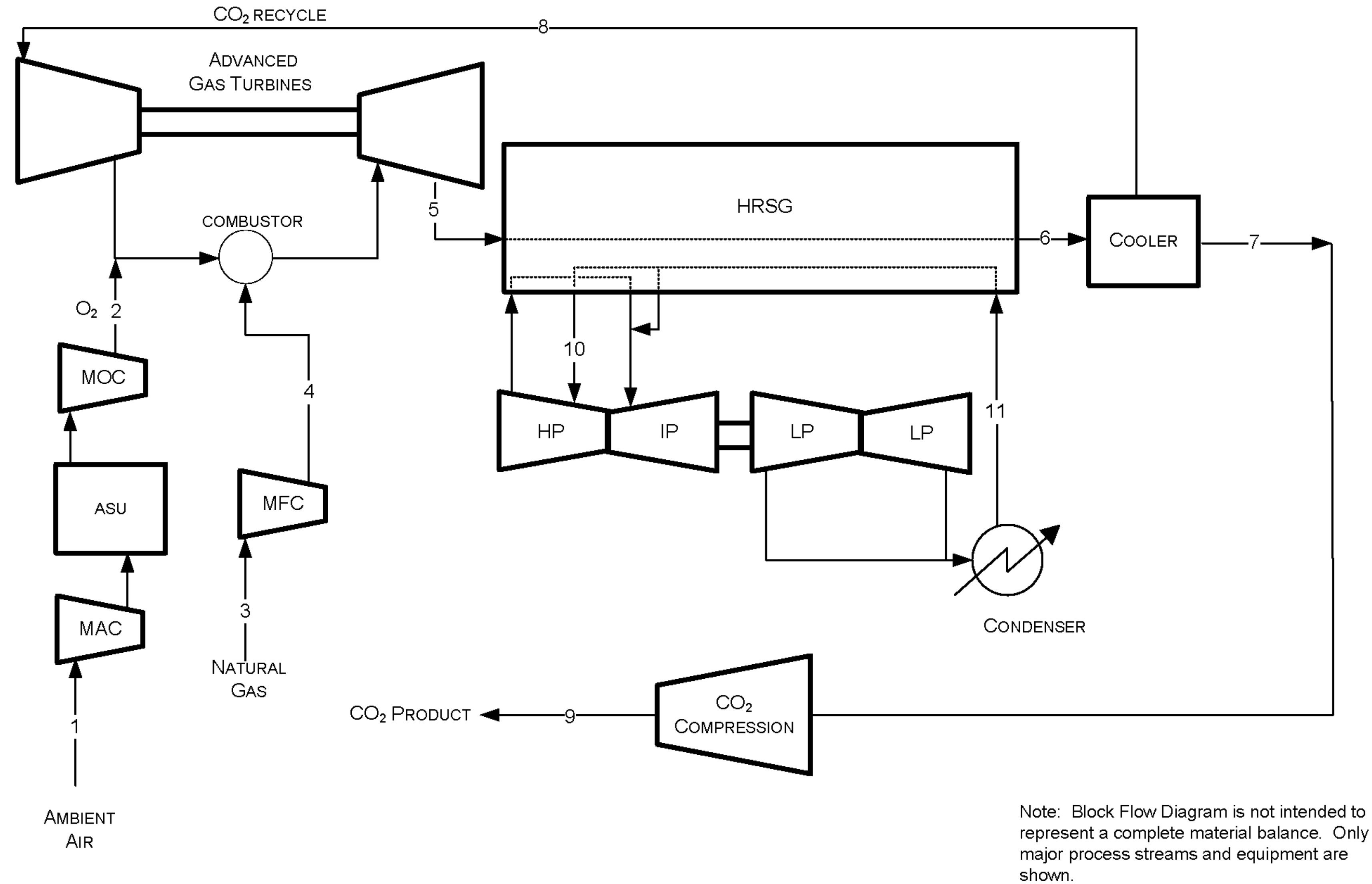
- Recuperative cycle

  Use a pressure ratio of 18.5 which would result in a high outlet temperature. Before the turbine exhaust stream enters the HRSG, a recuperator could be used to preheat the stream exiting the compressor and reducing the fuel requirement in the combustor. After the recuperator, the lower temperature exhaust stream would enter the HRSG. This option was not evaluated but may be looked at in the future.

The overall plant efficiency of this study is comparable to published data. The efficiency in this study is 46.2 percent (LHV), whereas published data show 46.2 percent (LHV) [15] and 47.0 percent (LHV) [14].

There are no estimated air emissions for Case 4 because all the exit gas streams except the ASU vent are included in the CO_2 Product for sequestration.

Exhibit 3-41 Case 4 Block Flow Diagram, Oxy-combustion with CO_2 Recycle



Note: Actual process consists of 2 Advanced F-class CTGs, 2 HRSGs, and 1 STG in a multi-shaft 2x2x1 configuration

Exhibit 3-42 Case 4 Stream Tables, Oxy-combustion with CO_2 Recycle

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0092	0.0100	0.0000	0.0000	0.0176	0.0176	0.0200	0.0200	0.0200	0.0000	0.0000
CH_4	0.0000	0.0000	0.9310	0.9310	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0320	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0070	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0040	0.0040	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	0.0000	0.0000	0.0000	0.0000
CO_2	0.0005	0.0000	0.0100	0.0100	0.8332	0.8332	0.9466	0.9466	0.9471	0.0000	0.0000
H_2O	0.0101	0.0000	0.0000	0.0000	0.1203	0.1203	0.0006	0.0006	0.0000	1.0000	1.0000
N_2	0.7729	0.0000	0.0160	0.0160	0.0131	0.0131	0.0148	0.0149	0.0148	0.0000	0.0000
O_2	0.2074	0.9900	0.0000	0.0000	0.0159	0.0159	0.0180	0.0180	0.0180	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	43,629	8,877	4,275	4,275	71,594	71,594	4,706	58,314	4,703	23,401	27,523
V-L Flowrate (kg/hr)	1,258,955	284,752	74,083	74,083	2,893,317	2,893,317	204,517	2,534,338	204,469	421,581	495,838
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	114	38	55	694	97	32	32	135	649	38
Pressure (MPa, abs)	0.10	4.55	3.10	4.55	0.11	0.10	0.10	0.10	15.27	16.65	0.01
Enthalpy (kJ/kg)[A]	30.57	97.41	46.30	71.73	901.58	222.58	26.56	26.54	36.31	3,692.36	160.61
Density (kg/m^3)	1.2	45.6	22.2	31.2	0.5	1.3	1.7	1.7	248.6	41.8	992.9
V-L Molecular Weight	28.856	32.079	17.328	17.328	40.413	40.413	43.460	43.460	43.475	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	96,185	19,570	9,426	9,426	157,837	157,837	10,375	128,560	10,369	51,591	60,678
V-L Flowrate (lb/hr)	2,775,521	627,771	163,325	163,325	6,378,672	6,378,672	450,883	5,587,258	450,777	929,428	1,093,135
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	238	100	130	1,281	206	90	90	274	1,200	101
Pressure (psia)	14.7	660.0	450.0	660.0	15.2	14.7	14.6	14.6	2,214.7	2,414.7	1.0
Enthalpy (Btu/lb)[A]	13.1	41.9	19.9	30.8	387.6	95.7	11.4	11.4	15.6	1,587.4	69.1
Density (lb/ft^3)	0.076	2.850	1.384	1.948	0.033	0.084	0.108	0.108	15.522	2.610	61.982

Note: Flows shown are for total values for all process trains

The oxy-combustion, CO_2 recycle plant produces a net output of 449 MW at a net plant efficiency of 41.6 percent (HHV basis). Overall plant performance is summarized in Exhibit 3-43. The summary includes auxiliary power requirements.

Exhibit 3-43 Case 4 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	362,100
Steam Turbine Power	223,800
TOTAL POWER, kWe	**585,900**
AUXILIARY LOAD SUMMARY, kWe	
ASU Compression	70,740
Oxygen Compression	28,520
Natural Gas Compression	1,410
CO_2 Compression	20,880
Condensate Pumps	160
Boiler Feedwater Pumps	2,900
Amine System (MDEA) Auxiliaries	N/A
Circulating Water Pump	5,260
Ground Water Pumps	440
Cooling Tower Fans	2,720
SCR	0
Gas Turbine Auxiliaries	1,000
Steam Turbine Auxiliaries	100
Miscellaneous Balance of Plant[1]	500
Transformer Losses	1,850
TOTAL AUXILIARIES, kWe	**136,480**
Plant Performance	
NET POWER, kWe	**449,420**
Net Plant Efficiency (HHV)	41.6%
Net Plant Efficiency (LHV)	46.2%
Net Plant Heat Rate (HHV), kJ/kWh (Btu/kWh)	8,646 (8,195)
Net Plant Heat Rate (LHV), kJ/kWh (Btu/kWh)	7,795 (7,388)
CONDENSER COOLING DUTY, GJ/hr (10^6 Btu/hr)	1,139 (1,080)
CONSUMABLES	
Natural Gas Feed Flow, kg/hr (lb/hr)	74,083 (163,325)
Thermal Input (HHV), kW_{th}	1,079,327
Thermal Input (LHV) , kW_{th}	973,152
Raw Water Withdrawal, m^3/min (gpm)	12.7 (3,344)
Raw Water Consumption, m^3/min (gpm)	9.3 (2,454)
CO_2 Capture	>99%
CO_2 emissions tonne/yr (ton/yr)[2]	0
CO_2 emissions kg/MWh (lb/MWh)	0

1 Includes plant control systems, lighting, HVAC and miscellaneous low voltage loads
2 Based on 85% capacity factor

The carbon balance is shown in Exhibit 3-44. The carbon input to the plant consists of carbon in the air and the carbon in the natural gas. Carbon leaves the plant as CO_2 product. The percent of total carbon sequestered is defined as the amount of carbon product divided by the carbon in the natural gas feedstock, expressed as a percentage.

% Captured = Carbon in Product for Sequestration / Carbon in the Natural gas

or

53,503/53,509 *100 = >99%

Exhibit 3-44 Case 4 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	53,509 (117,968)	**Stack Gas**	0 (0)
Air (CO_2)	241 (530)	**CO_2 Product**	53,503 (117,955)
		ASU Vent	241 (530)
		Convergence Tolerance*	6 (13)
Total	53,750 (118,498)	**Total**	53,570 (118,498)

*by difference

An overall water balance for the plant is shown in Exhibit 3-45. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process and is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

Exhibit 3-45 Case 4 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Condenser Makeup	0.08 (22)	0 (0)	0.08 (22)	0 (0)	0.08 (22)
BFW Makeup	0.08 (22)	0 (0)	0.08 (22)		
Cooling Tower	15.0 (3,957)	2.4 (635)	12.6 (3,322)	3.4 (890)	9.2 (2,432)
BFW Blowdown	0 (0)	0.1 (22)	-0.08 (-22)		
Flue Gas Condensate	0 (0)	2.3 (613)	-2.32 (-613)		
CO_2 Product Condensate	0 (0)	0 (0)	0 (0)		
Total	**15.1 (3,979)**	**2.4 (635)**	**12.7 (3,344)**	**3.4 (890)**	**9.3 (2,454)**

An overall plant energy balance is provided in tabular form in Exhibit 3-46. The power out is the combined combustion turbine and steam turbine power after generator losses.

Exhibit 3-46 Case 4 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Natural Gas	3,886 (3,683)	2.6 (2.5)		3,888 (3,685)
Ambient Air		0 (0)		0 (0)
Raw Water Makeup		47.6 (45.1)		48 (45)
Auxiliary Power			491 (466)	491 (466)
TOTAL	**3,886 (3,683)**	**70.9 (67.2)**	**491 (466)**	**4,448 (4,216)**
Heat Out GJ/hr (MMBtu/hr)				
CO_2		7.4 (7.0)		7 (7)
Cooling Tower Blowdown		34.2 (32.5)		34 (33)
Combustion Turbine Heat Loss		54.9 (52.0)		55 (52)
Stack Gas		550.3 (521.5)		550 (522)
Condenser		69.3 (65.7)		69 (66)
Non-Condenser Cooling Tower Loads		1,135 (1,076)		1,135 (1,076)
Process Losses*		487 (462)		487 (462)
Power			2,109 (1,999)	2,109 (1,999)
TOTAL		**2,339 (2,217)**	**2,109 (1,999)**	**4,448 (4,216)**

*Process losses including steam turbine, combustion reactions, HRSG, gas turbine, and gas cooling are estimated to match the heat input to the plant.

3.6 CASE 5- OXY-COMBUSTION TURBINE WITH WATER/STEAM RECYCLE

This case is an NGCC plant based on an oxy-combustion Clean Energy Systems (CES) design.

A typical CES process includes a high pressure, oxy-combustion gas generator, a high pressure expander, and a reheat combustor followed by IP and LP expanders. A partial condenser following the LP expander is used to recover water while the remaining CO_2 rich stream enters a compression section. Initial CO_2 compressor stages are used to pressurize the CO_2 from 1.9 psia to about atmospheric pressure and to recover most of the remaining water. This is followed by a compression section similar to other cases in this study, which results in a CO_2 product pressure of 2,215 psia. This study is not a typical CES design because it recycles and injects steam and water into the gas generator. The steam is required to increase the mass flow of the working fluid and thereby increase the power generation. The steam is generated by modifying the heat exchanger network. This would require a redesign of the CES gas generator. A BFD and stream tables are shown in Exhibit 3-47 and Exhibit 3-48, respectively.

A low pressure cryogenic air separation unit supplies high purity oxygen required for the gas generator and for the reheat combustor. Auxiliary compressors are used to increase the oxygen pressure to levels required for the gas generator (2,500 psia) and for the reheat combustor (420 psia). Natural gas is assumed to be available at 100°F and 450 psia. This is sufficient for the reheat combustor, but an auxiliary compressor is required to supply the natural gas to the high pressure gas generator.

The gas generator is an oxy-combustion reactor that uses both water and steam injection to moderate temperature and produce a high pressure working fluid consisting of 95 mol% water and 5 mol% CO_2. The gas stream exiting the gas generator enters the high pressure expander at 1,800°F and 2,150 psia and is expanded to 450 psia. After being reheated in an auxiliary combustor, the working fluid goes through a series of IP and LP expanders before a large portion of the water is recovered in a partial condenser. The CO_2 and remaining water enter the 7-stage CO_2 compression train. Knockout water from the CO_2 compressor is combined with water from the partial condenser, passes through a H_2O treatment unit, and is recycled back to the gas generator as liquid water and steam.

An alternative to this study would be to raise the low pressure turbine exit pressure to slightly above atmospheric pressure and introduce a HRSG and ST into the system. This change would cause a lower volumetric flow rate entering the CO_2 compression train and make it easier to condense H_2O. However, the plant cost would increase because of the addition of the HRSG and ST.

Like Case 4, the amount of CO_2 captured was greater than 99 percent, which is expected from an oxy-combustion system.

This case plant produces a net output of 406 MW at a net plant efficiency of 44.7 percent (HHV basis). Overall plant performance is summarized in Exhibit 3-49. The summary includes auxiliary power requirements.

There are no estimated air emissions for Case 5 because all the exit gas streams except the ASU vent are included in the CO_2 Product for sequestration.

Exhibit 3-47 Case 5 Block Flow Diagram, Oxy-combustion Turbine

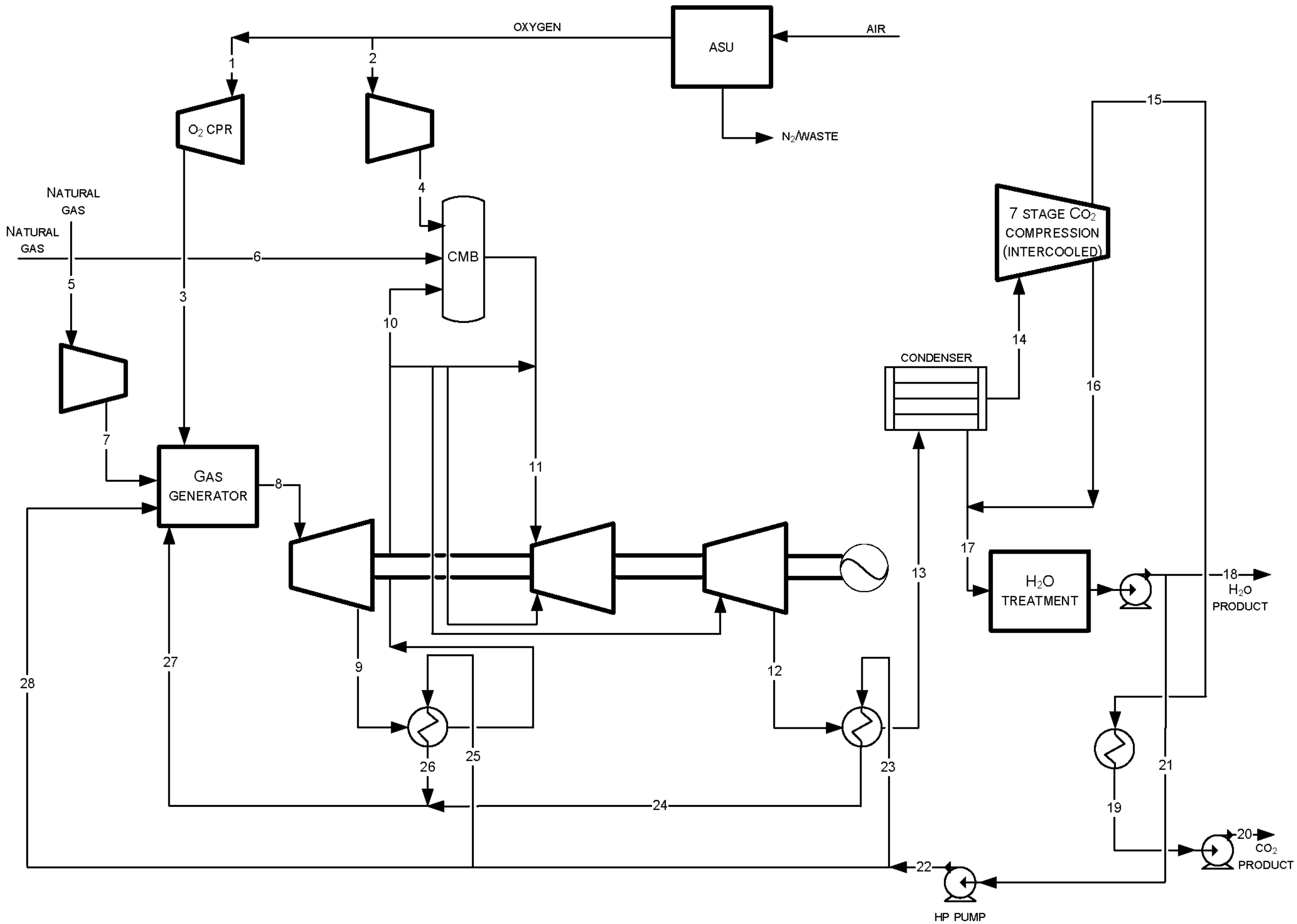


Exhibit 3-48 Case 5 Stream Tables, CES-Based Oxy-combustion Turbine

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0020	0.0020	0.0020	0.0020	0.0000	0.0000	0.0000	0.0002	0.0002	0.0002	0.0004	0.0004	0.0004	0.0019
CH_4	0.0000	0.0000	0.0000	0.0000	0.9310	0.9310	0.9310	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0320	0.0320	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0000	0.0000	0.0070	0.0070	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0000	0.0040	0.0040	0.0040	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0000	0.0000	0.0000	0.0000	0.0100	0.0100	0.0100	0.0514	0.0514	0.0514	0.1075	0.1048	0.1048	0.4776
H_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.9463	0.9463	0.9463	0.8876	0.8905	0.8905	0.5010
N_2	0.0030	0.0030	0.0030	0.0030	0.0160	0.0160	0.0160	0.0011	0.0011	0.0011	0.0023	0.0022	0.0022	0.0102
O_2	0.9950	0.9950	0.9950	0.9950	0.0000	0.0000	0.0000	0.0010	0.0010	0.0010	0.0021	0.0020	0.0020	0.0093
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	2,961	4,463	2,961	4,463	1,433	2,160	1,433	29,045	29,045	26,431	33,988	35,731	35,731	7,841
V-L Flowrate (kg/hr)	94,765	142,815	94,765	142,815	24,839	37,433	24,839	562,947	562,947	512,282	709,419	743,196	743,196	240,744
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	32	32	129	149	38	38	117	982	669	316	1,426	422	59	38
Pressure (MPa, abs)	0.21	0.21	17.24	2.90	3.10	3.10	17.24	14.82	2.76	2.69	2.62	0.01	0.01	0.01
Enthalpy (kJ/kg)[A]	28.97	28.97	103.20	135.37	48.09	48.09	168.64	-6,445,108.64	3,461.26	2,722.28	4,801.97	2,661.09	2,026.47	778.47
Density (kg/m^3)	2.6	2.6	157.0	26.2	21.8	21.8	92.5	27.2	6.9	11.3	3.9	0.1	0.1	0.2
V-L Molecular Weight	32.003	32.003	32.003	32.003	17.327	17.327	17.327	19.382	19.382	19.382	20.873	20.800	20.800	30.703
V-L Flowrate (lb_{mol}/hr)	6,528	9,838	6,528	9,838	3,160	4,763	3,160	64,034	64,034	58,271	74,931	78,773	78,773	17,287
V-L Flowrate (lb/hr)	208,921	314,852	208,921	314,852	54,760	82,525	54,760	1,241,087	1,241,087	1,129,389	1,564,001	1,638,466	1,638,466	530,751
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	90	90	264	300	100	100	243	1,799	1,236	600	2,599	792	139	100
Pressure (psia)	30	30	2,500	420	450	450	2,500	2,150	400	390	380	2	2	2
Enthalpy (Btu/lb)[A]	12.5	12.5	44.4	58.2	20.7	20.7	72.5	-2,770,898.0	1,488.1	1,170.4	2,064.5	1,144.1	871.2	334.7
Density (lb/ft^3)	0.163	0.163	9.804	1.637	1.359	1.359	5.774	1.698	0.429	0.704	0.241	0.003	0.006	0.010

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 3-48 Case 5 Stream Tables, CES-Based Oxy-combustion Turbine (continued)

	15	16	17	18	19	20	21	22	23	24	25	26	27	28
V-L Mole Fraction														
Ar	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	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	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	1.0000	0.0001	0.0000	0.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0000	0.9999	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	3,744	3,918	31,808	7,199	3,744	3,744	24,609	24,609	11,217	11,217	9,936	9,936	21,153	3,456
V-L Flowrate (kg/hr)	164,788	70,599	573,050	129,689	164,788	164,788	443,343	443,343	202,080	202,080	179,006	179,006	381,086	62,257
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	119	38	36	38	27	28	38	39	39	353	39	353	353	39
Pressure (MPa, abs)	14.48	0.04	0.01	0.34	14.20	15.27	0.34	17.93	17.93	17.24	17.93	17.24	17.24	17.93
Enthalpy (kJ/kg)[A]	12.18	153.68	153.62	159.40	-230.40	-228.52	159.40	180.23	180.23	2,510.83	180.23	2,504.21	2,679.93	176.95
Density (kg/m^3)	253.3	663.8	993.6	993.2	757.3	761.2	993.2	1,000.3	1,000.3	126.3	1,000.3	127.1	108.6	747.6
V-L Molecular Weight	44.010	18.017	18.016	18.016	44.010	44.010	18.016	18.016	18.016	18.016	18.016	18.016	18.016	18.016
V-L Flowrate (lb_{mol}/hr)	8,255	8,639	70,125	15,870	8,255	8,255	54,254	54,254	24,729	24,729	21,906	21,906	46,635	7,619
V-L Flowrate (lb/hr)	363,295	155,644	1,263,359	285,915	363,295	363,295	977,405	977,405	445,511	445,511	394,641	394,641	840,152	137,253
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	246	100	98	100	80	83	100	102	102	668	102	668	668	102
Pressure (psia)	2,100	6	2	50	2,060	2,215	50	2,600	2,600	2,500	2,600	2,500	2,500	2,600
Enthalpy (Btu/lb)[A]	5.2	66.1	66.0	68.5	-99.1	-98.2	68.5	77.5	77.5	1,079.5	77.5	1,076.6	1,152.2	76.1
Density (lb/ft^3)	15.816	41.440	62.028	62.006	47.275	47.519	62.006	62.449	62.449	7.882	62.449	7.932	6.782	46.672

Exhibit 3-49 Case 5 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Steam Turbine Power	528,700
TOTAL POWER, kWe	**528,700**
AUXILIARY LOAD SUMMARY, kWe	
ASU Main Air Compressor	50,920
Oxygen Compression	26,740
Natural Gas Compression	2,340
CO_2 Compression	31,940
Condensate Pumps	70
Boiler Feedwater Pumps	2,660
Amine System (MDEA) Auxiliaries	N/A
Circulating Water Pump	3,086
Ground Water Pumps	255
Cooling Tower Fans	1,592
Turbine Auxiliaries	829
Miscellaneous Balance of Plant[1]	500
Transformer Losses	1,840
TOTAL AUXILIARIES, kWe	**122,772**
Plant Performance	
NET POWER, kWe	**405,928**
Net Plant Efficiency (HHV)	44.7%
Net Plant Efficiency (LHV)	49.6%
Net Plant Heat Rate (HHV), kJ/kWh (Btu/kWh)	8,046 (7,626)
Net Plant Heat Rate (LHV), kJ/kWh (Btu/kWh)	7,255 (6,876)
CONDENSER COOLING DUTY, GJ/hr (10^6 Btu/hr)	1,245 (1,180)
CONSUMABLES	
Natural Gas Feed Flow, kg/hr (lb/hr)	62,272 (137,285)
Thermal Input (HHV), kW_{th}	907,255
Thermal Input (LHV) , kW_{th}	818,007
Raw Water Withdrawal, m^3/min (gpm)	23.6 (6,234)
Raw Water Consumption, m^3/min (gpm)	18.3 (4,832)
CO_2 Capture	>99%
CO_2 emissions tonne/yr (ton/yr)[2]	0
CO_2 emissions kg/MWh (lb/MWh)	0

1 Includes plant control systems, lighting, HVAC and miscellaneous low voltage loads
2 Based on 85% capacity factor

The carbon balance is shown in Exhibit 3-50 . The carbon input to the plant consists of carbon in the air and the carbon in the natural gas. Carbon leaves the plant as ASU vent gas, H_2O purification off-gas and CO_2 product. The percent of total carbon sequestered is defined as the amount of carbon product produced divided by the carbon in the natural gas feedstock, expressed as a percentage.

% Captured = Carbon in Product for Sequestration / Carbon in the Natural gas

or

44,974/44,979 *100 = >99%

Exhibit 3-50 Case 5 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	44,979 (99,161)	**Stack Gas**	0 (0)
Air (CO_2)	102 (224)	**CO_2 Product**	44,974 (99,150)
		ASU Vent	102 (224)
		H_2O Purification	5 (11)
		Convergence Tolerance*	6 (13)
Total	45,080 (99,385)	**Total**	45,080 (99,385)

*by difference

An overall water balance for the plant is shown in Exhibit 3-51. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process and is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. The difference between water withdrawal and process water discharge is defined as water consumption and can be represented by the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to the water source from which it was withdrawn. Water consumption represents the net impact of the plant process on the water source balance.

Exhibit 3-51 Case 5 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Cooling Tower	12.55 (3,316)	1.95 (515)	10.60 (2,801)	2.8 (746)	7.78 (2,056)
CO_2 Product Condensate	0 (0)	1.95 (515)	-1.95 (-515)		
Total	**12.6 (3,316)**	**1.9 (515)**	**10.6 (2,801)**	**2.8 (746)**	**7.8 (2,056)**

An overall plant energy balance is provided in tabular form in Exhibit 3-52. The power out is the combined combustion turbine and steam turbine power after generator losses.

Exhibit 3-52 Case 5 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Natural Gas	3,266 (3,096)	2.2 (2.1)		3,268 (3,098)
Raw Water Makeup		39.9 (37.8)		40 (38)
Auxiliary Power			442 (419)	442 (419)
TOTAL	**3,266 (3,096)**	**42.1 (39.9)**	**442 (419)**	**3,750 (3,554)**
Heat Out GJ/hr (MMBtu/hr)				
CO_2		-37.7 (-35.7)		-38 (-36)
Cooling Tower Blowdown		21.0 (19.9)		21 (20)
Compression Intercooling		397.7 (376.9)		398 (377)
Condenser		1,241 (1,177)		1,241 (1,177)
Process Losses*		224 (213)		224 (213)
Power			1,903 (1,804)	1,903 (1,804)
TOTAL		**1,847 (1,750)**	**1,903 (1,804)**	**3,750 (3,554)**

*Process losses including steam turbine, combustion reactions, HRSG, gas turbine, and gas cooling are estimated to match the heat input to the plant.

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4. COST RESULTS

The estimating methodology for capital costs, operations and maintenance costs, and CO_2 TS&M costs are described below. The finance structure, basis for the discounted cash flow analysis, and Cost of Electricity (COE) calculations are also described. More detailed information on the cost estimating methodology, calculations, and assumptions is presented in Section 2 of the Bituminous Baseline study report [1] which contains the reference cases. Capital costs were estimated at four levels: Bare Erected Cost (BEC), Total Plant Cost (TPC), Total Overnight Cost (TOC) and Total As-spent Capital (TASC). BEC, TPC and TOC are “overnight” costs and are expressed in “base-year” dollars. The base year is the first year of capital expenditure, which for this study is assumed to be 2007. TASC is expressed in mixed-year, current-year dollars over the entire capital expenditure period, which is assumed to last three years for natural gas plants (2007 to 2009).

4.1 EQUIPMENT CAPITAL COSTING

The Bare Erected Cost (BEC) comprises the cost of process equipment, on-site facilities and infrastructure that support the plant (e.g., shops, offices, labs, road), and the direct and indirect labor required for its construction and/or installation. The cost of EPC services and contingencies is not included in BEC. BEC is an overnight cost expressed in base-year (2007) dollars.

The equipment cost estimating methodology for Cases 1 through 5 uses reference costs established in Bituminous Baseline Study [1]. WorleyParsons Group Inc. provided the Total Plant Cost (TPC) and Operation and Maintenance (O&M) costs for cases in the referenced study. Those cost data were employed for scaling conventional systems in the present cost estimates. All costs are in June 2007 dollars, similar to the basis of Bituminous Baseline Study, to facilitate comparison. Owner’s costs are included in the present estimates as they were in the Baseline study estimates.

Specifically, the costs and performance data from Case 14 - NGCC with CO_2 capture, in the Bituminous Baseline Study were employed as the references for cost scaling. The costs and performance data for this study’s reference case with CO_2 capture, Ref2, were assumed to match the BB Case 14 values with the exception of an increase in the MEA auxiliary load and corresponding small decrease in the overall efficiency as described in section 3.1.2. Natural gas combined cycle plants have the following general plant subsystems in addition to site improvements, and buildings and structures:

- Natural gas supply system;
- Combustion turbine;
- Heat recovery steam generator;
- NO_x control system;
- Carbon dioxide recovery facility;
- Steam turbine;
- Water and steam systems;
- Accessory electric plant; and
- Instrumentation and control.

Most of the cases in this study include equipment items that aren't found in a conventional NGCC plant. Costs for the non-conventional equipment were estimated by WorleyParsons, scaled from integrated gasification combined cycle (IGCC) equipment items, or from other reference estimates as described below:

- Case 2, Account 4.4 (Low Temperature Heat Recovery): This account consists of four heat exchangers, three are gas-liquid and one is gas-gas. The costs were estimated by WorleyParsons using reference costs from other projects employing heat exchangers operating at similar conditions and scaling on heat duty.
- Case 2, Account 5.1 (Autothermal Reactor): This account was scaled by the project team using a steam-methane reformer cost from an ongoing coal-to-hydrogen study as the reference [16]. The cost was scaled on outlet volumetric flow rate. The resulting cost was multiplied by 0.8, assuming that an ATR is less expensive than an equivalent steam-methane reformer.
- Case 2, Account 5A.2 (Syngas Cooling): This gas-gas heat exchanger is similar to a syngas cooler in an IGCC plant and was scaled using the replacement cost of the Wabash River syngas cooler as a reference. It was assumed that two exchangers would be required, and they were scaled on heat duty.
- Case 2, Account 5A.3 (Gas-Gas Heat Exchanger): WorleyParsons provided the cost estimate for this exchanger using a gas-gas heat exchanger (used to cool extraction air from a combustion turbine) as the reference. Cost was scaled based on heat duty.
- Case 2, Account 5B.1 (MDEA CO_2 Removal System): The cost of the MDEA system was scaled by WorleyParsons from previous MDEA cost estimates that appear in the Bituminous Baseline report [1].
- Case 3, Account 4.4 (Low Temperature Heat Recovery): This account consists of three gas-liquid heat exchangers and an economizer/evaporator/superheater. The costs were estimated by WorleyParsons using reference costs from other projects employing heat exchangers operating at similar conditions and scaling on heat duty.
- Case 3, Account 5A.1 (Partial Oxidizers): The partial oxidizer costs were scaled by using the CoP gasifier cost in Bituminous Baseline study as a reference [1]. Per WorleyParsons, syngas cooling represents 41 percent of gasifier equipment costs and 44 percent of labor costs in the reference estimate. The POX scaling parameter, volumetric flow rate, was multiplied by 0.6 to account for the reduced residence time required to combust natural gas relative to coal. A correction factor was applied to account for the higher POX operating pressure.
- Case 3, Account 5A.2 (Syngas Cooling): WorleyParsons estimated the cost of this quench vessel using an FGD absorber for the reference cost with modifications for operating conditions.
- Case 3, Account 5A.3 (Syngas Cooling): WorleyParsons provided the cost estimate for this exchanger using a gas-gas heat exchanger (used to cool extraction air from a combustion turbine) as the reference. Cost was scaled based on heat duty.

- Case 3, Account 5B.1 (MDEA CO_2 Removal System): The cost of the MDEA system was scaled by WorleyParsons from previous MDEA cost estimates that appear in the Bituminous Baseline report [1].
- Case 3, Account 6.5 (Syngas Expander): The cost of the syngas expander was estimated by WorleyParsons based on quotes for similar equipment items with corrections for operating conditions.
- Case 4, Account 6.1 (Combustion Turbine Generator): A factor of 1.5 was applied to the equipment cost of a conventional CT to account for the higher than normal pressure ratio and oxy-combustion configuration. A process contingency was also applied as described in Section 4.3.
- Case 5, Account 5A.1 (Gas Generator): The CoP gasifier from the Bituminous Baseline study was used as the reference cost and volumetric flow rate was the process parameter used for scaling with corrections made for pressure differences.
- Case 5, Account 6.5 (HTHP Gas Expanders): WorleyParsons estimated the expander cost with a 50 percent margin included in the equipment cost for advanced metallurgy.

The technologies modeled in this study include some unconventional equipment items that operate at temperatures and pressures beyond that currently offered commercially. Because the estimates for these equipment items contain a large degree of uncertainty, cost sensitivities were performed to show the impact of higher or lower than estimated costs.

The rest of costing methodology is outlined in the following sections.

4.1.1 Scaling Methodology

The costs for all the subsystems in the reference case are available along with the heat and material balances. These costs were scaled on the basis of the unit attributes (such as capacity, heat duty, and inlet flow rate) following the generally accepted scaling equation shown below.

$$\frac{C_a}{C_b} = \left(\frac{A_a}{A_b}\right)^n \quad (1)$$

where C_a and C_b = the costs of plant a and plant b (or equipment a and equipment b)

A_a and A_b = attributes of plant (or equipment) a and b respectively.

The cost exponent, n, depends on the type of equipment in question and can range from 0.30 for a vertical carbon steel tank to 0.84 for a reciprocating compressor. The majority of the exponents for this scaling have already been established in previous studies and those exponents were used as is. For equipment without established exponents, an exponent of 0.7 was used. The attributes for scaling vary from equipment to equipment. For example, the HRSG was scaled against the heat duty of the HRSG (exponent = 0.70); steam turbine costs were scaled against the steam turbine capacity (exponent = 0.71); AGR costs were scaled by the actual gas flow rate into the AGR (exponent = 0.70); and ASU was scaled on the basis of air compressor power with an exponent of 0.57. Exponents employed for scaling various equipment items are detailed in the individual spreadsheet templates.

There were also new key pieces of equipment included in the simulations where costs were not available in the reference study. Those equipment items were independently estimated by WorleyParsons as discussed above, by employing available cost equations, or by scaling against costs of similar equipment. Since the cases evaluated had different cycle configurations and in some cases operated at vastly different temperatures and pressures from that of the reference case, simple scaling on the basis of Equation (1) alone was not sufficient to obtain reasonable estimated costs. Correction factors for pressure and temperature, in addition to the index correction to the June 2007 dollars, were evaluated separately and applied. They are discussed below.

4.1.2 <u>Purchased Equipment Costs</u>

For common equipment items whose costs were not available from previous studies or costs were not provided by WorleyParsons, the costs were calculated using the following equation. In some cases, the costs provided by WorleyParsons were for multiple equipment items lumped together and the breakdown of costs for individual pieces of equipment was difficult.

$$\log_{10} \mathrm{C_p^o} = K_1 + K_2 \log_{10} A + K_3 (\log_{10} A)^2 \qquad (2)$$

where C_p^o = equipment cost at ambient operating pressure using carbon steel

K_1, K_2, K_3 = correlation coefficients depending on the type of equipment

A = capacity or size parameters of the equipment

The correlation coefficients and the correct parameters to use for 'A' for different types of equipment are tabulated in Table A.1 of reference [17]. For centrifugal, axial, and reciprocating compressors, $K_1 = 2.2897$, $K_2 = 1.3604$, and $K_3 = -0.1027$ and the parameter A to use in Equation (2) is motor size in kW. Equation (2) was used to evaluate the costs of compressors with the pressure correction factors and material correction factors discussed below.

4.1.3 <u>Evaluation of Pressure Correction Factors</u>

The pressure correction factors for both the horizontal and vertical pressurized vessels were evaluated based on the following equation [17]. The applied pressure correction factor is the ratio of this factor calculated for the vessel of interest divided by the factor calculated for the reference vessel.

$$F_{p,vessel} = \frac{\left(\dfrac{(P+1)D}{2[350 - 0.6\,(P+1)]}\right) + 0.00315}{0.0063} \qquad for \;\; t_{vessel} > 6.3mm \qquad (3)$$

where D - diameter of the vessel in meter

$F_{p,vessel}$ - pressure correction factor for vessel

P - pressure in barg

t_{vessel} vessel thickness

Equation (3) was derived based on the American Society of Mechanical Engineers (ASME) code for pressure vessel design using a maximum allowable carbon steel stress of 944 bar (13,700 psi), a welding efficiency of 0.9, and a corrosion allowance of 3.15 mm (0.125 in).

For other process equipment, the pressure factors were calculated on the basis of Equation (4).

$$\log_{10} \mathrm{F_P} = C_1 + C_2 \log_{10} P + C_3 \left(\log_{10} P\right)^2 \qquad (4)$$

where F_p - pressure correction factor

C_1, C_2, C_3 - correlation constants depending on type of equipment

The correlation constants for different equipment items can be found in Table A.2 of reference [17]. For shell-and-tube heat exchangers, C_1 = 0.03881, C_2 = -0.11272, and C_3 = 0.08183; and for centrifugal pumps, C_1 = -0.3935, C_2 = 0.3957, and C_3 = -0.00226.

4.1.4 Evaluation of Material Correction Factors

The effect of temperature on equipment cost is usually manifested as changes in material requirements. The more exotic materials required, the higher the cost will be. The material correction factors depend also on the type of equipment under consideration. The material correction factors for process vessels, heat exchangers, and pumps are given in Figure A. 18 of reference [17]. In general, stainless steel will cost approximately 2 to 3 times that of carbon steel, Nickel (Ni) alloy 3 to 4 times, and Titanium (Ti) up to 10 times.

4.2 ENGINEERING, PROCUREMENT, AND CONSTRUCTION COSTS

The costs of services provided by the engineering, procurement and construction (EPC) contractor and project and process contingencies were estimated for each account based on the factors applied to similar equipment estimates in previous studies, most notably the referenced BB study natural gas cases [1]. More detailed information on estimating and applying contingencies is presented in Section 2.7.1 of the Bituminous Baseline study report [1]. EPC services include: detailed design, contractor permitting (i.e., those permits that individual contractors must obtain to perform their scopes of work, as opposed to project permitting, which is not included here), and project/construction management costs.

4.3 PROCESS AND PROJECT CONTINGENCIES

Process and project contingencies are included in estimates to account for unknown costs that are omitted or unforeseen due to a lack of complete project definition and engineering. Contingencies are added because experience has shown that such costs are likely, and expected, to be incurred even though they cannot be explicitly determined at the time the estimate is prepared. The contingencies are estimated for each account based on the factors applied to similar equipment estimates in previous studies, most notably the referenced BB study natural gas cases [1]. More detailed information on estimating and applying contingencies is presented in Section 2.7.1 of the Bituminous Baseline study report [1]. Project contingencies were added to cover project uncertainty and the cost of any additional equipment that would result from a detailed design. The contingencies represent costs that are expected to occur. Process contingency is intended to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of a technology. The cost accounts that received a process contingency include:

- Combustion Turbine Generator - five percent contingency on Cases 1a and 1c for modifications required to the inlet air system (material, static mixer) and the compressor. A 40 percent contingency is used in Case 1b since major design modifications would be required to accommodate the low combustor oxygen concentration, and 40 percent was also used for Case 4 because of the oxy-combustion configuration and the high pressure ratio.
- MEA-based carbon capture processes - 20 percent contingency on Cases Ref2, 1a, 1b, and 1c because it is considered unproven at commercial scale for power plant applications.
- Gas Generator and Reheat Combustor – 15 percent contingency for high-pressure oxy-combustion reactors.

4.4 OWNER'S COSTS AND TOTAL CAPITAL COSTS

The TPC comprises the BEC plus the cost of services provided by the engineering, procurement and construction (EPC) contractor and project and process contingencies. Owner's costs are added to the TPC estimate to generate the total overnight cost (TOC) values. With some exceptions, the estimation method follows guidelines in Sections 12.4.7 to 12.4.12 of AACE International Recommended Practice No. 16R-90 [18]. The Electric Power Research Institute's "Technical Assessment Guide (TAG®) – Power Generation and Storage Technology Options" also has guidelines for estimating owner's costs. The EPRI and AACE guidelines are very similar. In instances where they differ, this study has sometimes adopted the EPRI approach. The owner's costs included in the TOC cost estimate for this study are shown in Exhibit 4-1.

The TOC comprises the TPC plus owner's costs. TOC is an "overnight" cost, expressed in base-year (2007) dollars and as such does not include escalation during construction or interest during construction. TOC is an overnight cost expressed in base-year (2007) dollars.

The TASC is the sum of all capital expenditures as they are incurred during the capital expenditure period including their escalation. TASC also includes interest during construction. Accordingly, TASC is expressed in mixed, current-year dollars over the capital expenditure period. TASC is calculated by multiplying a factor times the value of TOC. This calculation is discussed in more detail in Section 4.7.

Again, more detailed information on estimating and capital cost calculations is presented in Section 2.7.1 of the Bituminous Baseline study report [1].

Exhibit 4-1 Owner's Costs Included in TOC

Owner's Cost	Comprised of
Preproduction Costs	• 6 months O&M, and administrative & support labor • 1 month maintenance materials @ 100% CF • 1 month non-fuel consumables @ 100% CF • 1 month of waste disposal costs @ 100% CF • 25% of one month's fuel cost @ 100% CF • 2% of TPC
Inventory Capital	• 60 day supply of fuel and consumables @100% CF • 0.5% of TPC (spare parts)
Land	• $3,000/acre (300 acres for greenfield IGCC and PC, and 100 acres for NGCC)
Financing Costs	• 2.7% of TPC
Other Owner's Costs	• 15% of TPC
Initial Cost for Catalyst and Chemicals	• All initial fills not included in BEC
Prepaid Royalties	• Not included in owner's costs (included with BEC)
AFUDC and Escalation	• Varies based on levelization period and financing scenario • 33-yr IOU high risk: TASC = TOC * 1.078 • 33-yr IOU low risk: TASC = TOC * 1.075

4.5 EVALUATION OF O&M COSTS

The operating and maintenance labor costs were evaluated on the basis of labor requirements outlined in Exhibit 4-2 at an average labor cost of $34.65 per hour with an operating labor burden of 30 percent and a labor overhead charge rate of 25 percent of labor cost. Property taxes and insurance is 2 percent of the total plant cost. For Cases 1a, 1b, 1c and 4, the labor requirements were assumed to be similar to that of the reference case. For Cases 2 and 3, the labor requirements were assumed to be intermediate between that required in referenced BB study IGCC cases and the reference NGCC cases [1]. One additional operator and ½ of an additional skilled operator were added to Case 5 because of exceptionally high temperature and high pressure operations.

The variable O&M costs were scaled primarily from the reference case. For operations not in the reference case, references were obtained from other cases in previous studies. For example, reference costs for WGS catalyst and MDEA solution were from other cases in the BB study [1].

Exhibit 4-2 Operating and Maintenance Labor

Labor Classification	Number of Operators per Shift		
	Cases 1a, 1b, 1c & Case 4	Case 2 & Case 3	Case 5
Skilled Operator	1.0	1.5	2.0
Operator	3.3	5.0	6.0
Forman	1.0	1.0	1.0
Lab Technician, etc	1.0	1.5	1.5
Total	6.3	9.0	10.5

4.6 CO_2 TRANSPORT STORAGE AND MONITORING

The capital and operating costs for CO_2 transport, storage, and monitoring (TS&M) were independently estimated by NETL. Those costs were adjusted for the specific product CO_2 flow rates of each case in this study and used to estimate the additional component required for incorporating into the overall cost of electricity results. The cost metrics utilized in this study provide a best estimate of TS&M costs for a "typical" sequestration project, and may vary significantly based on variables such as terrain to be crossed by the pipeline, reservoir characteristics, and number of land owners from which sub-surface rights must be acquired. Raw capital and operating costs are derived from detailed cost metrics found in the literature, escalated to June 2007-year dollars using appropriate price indices. These costs were then verified against values quoted by industrial sources where possible. Where regulatory uncertainty exists or costs are undefined, such as liability costs and the acquisition of underground pore volume, analogous existing policies were used for representative cost scenarios. More detailed information on the TS&M cost calculations is presented in Section 3.7 of the Bituminous Baseline study report [1].

CO_2 is supplied to the pipeline at the plant fence line at a pressure of 15.3 MPa (2,215 psia). The CO_2 is transported 80 km (50 miles) via pipeline to a geologic sequestration field for injection into a saline formation.

The product gas composition varies in the cases presented. Cases 1a, 1b, 1c, 2, and 3 are expected to meet the specification described in Exhibit 4-3 [19]. The compositions of the CO_2 streams in the oxy-combustion cases (cases 4 and 5) do not meet the pipeline specification. While the recovery of the CO_2 is high for these cases, purifying this CO_2 stream presents some difficulty, with the presence of light gases making further separation costly. No attempt was made to further purify the CO_2 in this study.

Exhibit 4-3 CO_2 Pipeline Specification

Parameter	Units	Parameter Value
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	35 (95)
N_2 Concentration	ppmv	< 300
O_2 Concentration	ppmv	< 40
Ar Concentration	ppmv	< 10

4.7 FINANCE STRUCTURE, DISCOUNTED CASH FLOW ANALYSIS, AND COE

The global economic assumptions used in this study are listed in Exhibit 4-4. Finance structures were chosen based on the assumed type of developer/owner and the assumed risk profile of the plant being assessed (low-risk or high-risk). For this study the owner/developer was assumed to be an investor-owned utility (IOU). All NGCC cases with CO_2 capture were considered high risk. The non-capture NGCC reference case was considered low risk.

Exhibit 4-5 describes the low-risk IOU and high-risk IOU finance structures that were assumed for this study. These finance structures were recommended in a 2008 NETL report based on interviews with project developers/owners, financial organizations and law firms [20].

Exhibit 4-4 Global Economic Assumptions

Parameter	Value
TAXES	
Income Tax Rate	38% (Effective 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 years
CONTRACTING AND FINANCING TERMS	
Contracting Strategy	Engineering Procurement Construction Management (owner assumes project risks for performance, schedule and cost)
Type of Debt Financing	Non-Recourse (collateral that secures debt is limited to the real assets of the project)
Repayment Term of Debt	15 years
Grace Period on Debt Repayment	0 years
Debt Reserve Fund	None
ANALYSIS TIME PERIODS	
Capital Expenditure Period	Natural Gas Plants: 3 Years
Operational Period	30 years
Economic Analysis Period (used for IRROE)	33 Years (capital expenditure period plus operational period)
TREATMENT OF CAPITAL COSTS	
Capital Cost Escalation During Capital Expenditure Period (nominal annual rate)	3.6%[2]
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	3-Year Period: 10%, 60%, 30%
Working Capital	zero for all parameters
% of Total Overnight Capital that is Depreciated	100% *(this assumption introduces a very small error even if a substantial amount of TOC is actually non-depreciable)*
ESCALATION OF OPERATING REVENUES AND COSTS	
Escalation of COE (revenue), O&M Costs, and Fuel Costs (nominal annual rate)	3.0%[3]

[2] A nominal average annual rate of 3.6 percent is assumed for escalation of capital costs during construction. This rate is equivalent to the nominal average annual escalation rate for process plant construction costs between 1947 and 2008 according to the *Chemical Engineering* Plant Cost Index.

[3] An average annual inflation rate of 3.0 percent is assumed. This rate is equivalent to the average annual escalation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods, the so-called "headline" index of the various Producer Price Indices. (The Producer Price Index for the Electric Power Generation Industry may be more applicable, but that data does not provide a long-term historical perspective since it only dates back to December 2003.)

Exhibit 4-5 Financial Structure for Investor Owned Utility High and Low Risk Projects

Type of Security	% of Total	Current (Nominal) Dollar Cost	Weighted Current (Nominal) Cost	After Tax Weighted Cost of Capital
Low Risk				
Debt	50	4.5%	2.25%	
Equity	50	12%	6%	
Total			8.25%	7.39%
High Risk				
Debt	45	5.5%	2.475%	
Equity	55	12%	6.6%	
Total			9.075%	8.13%

For scenarios that adhere to the global economic assumptions listed in Exhibit 4-4 and utilize one of the finance structures listed in

Exhibit 4-5, the multipliers shown in Exhibit 4-6 can be used to translate TOC to TASC to account for the impact of both escalation and interest during construction. TOC is expressed in base-year dollars and the resulting TASC is expressed in mixed-year, current-year dollars over the entire capital expenditure period

Exhibit 4-6 TASC/TOC Factor

Finance Structure	High Risk IOU	Low Risk IOU
Capital Expenditure Period	**Three Years**	**Three Years**
TASC/TOC Factor	1.078	1.075

4.7.1 Estimating COE with Capital Charge Factors

The **COE** is the revenue received by the generator per net megawatt-hour during the power plant's first year of operation, *assuming that the COE escalates thereafter at a nominal annual rate equal to the general inflation rate, i.e., that it remains constant in real terms over the operational period of the power plant*.

For scenarios that adhere to the global economic assumptions listed in Exhibit 4-4 and utilize one of the finance structures listed in

Exhibit 4-5, the following simplified equation can be used to estimate COE as a function of TOC[4], fixed O&M, variable O&M (including fuel), capacity factor and net output. The equation requires the application of one of the capital charge factors (CCF) listed in Exhibit 4-7. These CCFs are valid only for the global economic assumptions listed in Exhibit 4-4, the stated finance structure, and the stated capital expenditure period.

[4] Although TOC is used in the simplified COE equation, the CCF that multiplies it accounts for escalation during construction and interest during construction (along with other factors related to the recovery of capital costs).

Exhibit 4-7 Capital Charge Factors for COE Equation

Finance Structure	High Risk IOU	Low Risk IOU
Capital Expenditure Period	Three Years	Three Years
Capital Charge Factor (CCF)	0.111	0.105

All factors in the COE equation are expressed in base-year dollars. The base year is the first year of capital expenditure, which for this study is assumed to be 2007. As shown in Exhibit 4-4, all factors (COE, O&M and fuel) are assumed to escalate at a nominal annual general inflation rate of 3.0 percent. Accordingly, all first-year costs (COE and O&M) are equivalent to base-year costs when expressed in base-year (2007) dollars.

$$COE = \frac{\begin{array}{c}\textit{first year}\\ \textit{capital charge}\end{array} + \begin{array}{c}\textit{first year}\\ \textit{fixed operating}\\ \textit{costs}\end{array} + \begin{array}{c}\textit{first year}\\ \textit{variable operating}\\ \textit{costs}\end{array}}{\begin{array}{c}\textit{annual net megawatt hours}\\ \textit{of power generated}\end{array}}$$

$$COE = \frac{(CCF)(TOC) + OC_{FIX} + (CF)(OC_{VAR})}{(CF)(MWH)}$$

where:

COE =	revenue received by the generator (\$/MWh, equivalent to mills/kWh) during the power plant's first year of operation (*but expressed in base-year dollars*), assuming that the COE escalates thereafter at a nominal annual rate equal to the general inflation rate, i.e., that it remains constant in real terms over the operational period of the power plant.
CCF =	capital charge factor taken from Exhibit 4-7 that matches the applicable finance structure and capital expenditure period
TOC =	total overnight capital, *expressed in base-year dollars*
OC_{FIX} =	the sum of all fixed annual operating costs, *expressed in base-year dollars*
OC_{VAR} =	the sum of all variable annual operating costs, including fuel at 100 percent capacity factor, *expressed in base-year dollars*
CF =	plant capacity factor, assumed to be constant over the operational period
MWH =	annual net megawatt-hours of power generated at 100 percent capacity factor

4.7.2 Estimating LCOE from COE

The **LCOE** was presented in Exhibit ES-2 for reference, and is the revenue received by the generator per net megawatt-hour during the power plant's first year of operation, *assuming that the COE escalates thereafter at a nominal annual rate of 0 percent, i.e., that it remains constant in nominal terms over the operational period of the power plant.* To calculate the LCOE, a levelization factor was applied to the COE values. The levelization factor and LCOE were calculated using the following equation.

$$LF = \frac{\frac{(1-(1+E)^{T}(1+DR)^{-T})}{(DR-E)}}{\frac{(1+DR)^{T}-1}{DR(1+DR)^{T}}}$$

$$LCOE_T = LF * COE$$

where

LF =	Levelization factor based on end of year values
E =	Escalation rate (3%)
DR =	Discount Rate (assumed equal to ROE = 12%)
T =	Levelization Time (30 years)
$LCOE_T$ =	levelized cost of electricity over T years, \$/MWh
COE =	cost of electricity estimated from CCFs using simplified equation, \$/MWh
ROE =	required rate of return on equity

4.7.3 Estimating Cost of CO_2 Avoided from COE

The CO_2-avoided costs are calculated as follows:

$$Avoided\ Cost = \frac{\{COE_{with\ removal} - COE_{w/o\ removal}\}\ \$/MWh}{\{CO_2 Emissions - CO_2 Emissions_{with\ removal}\}\ tons/MWh}$$

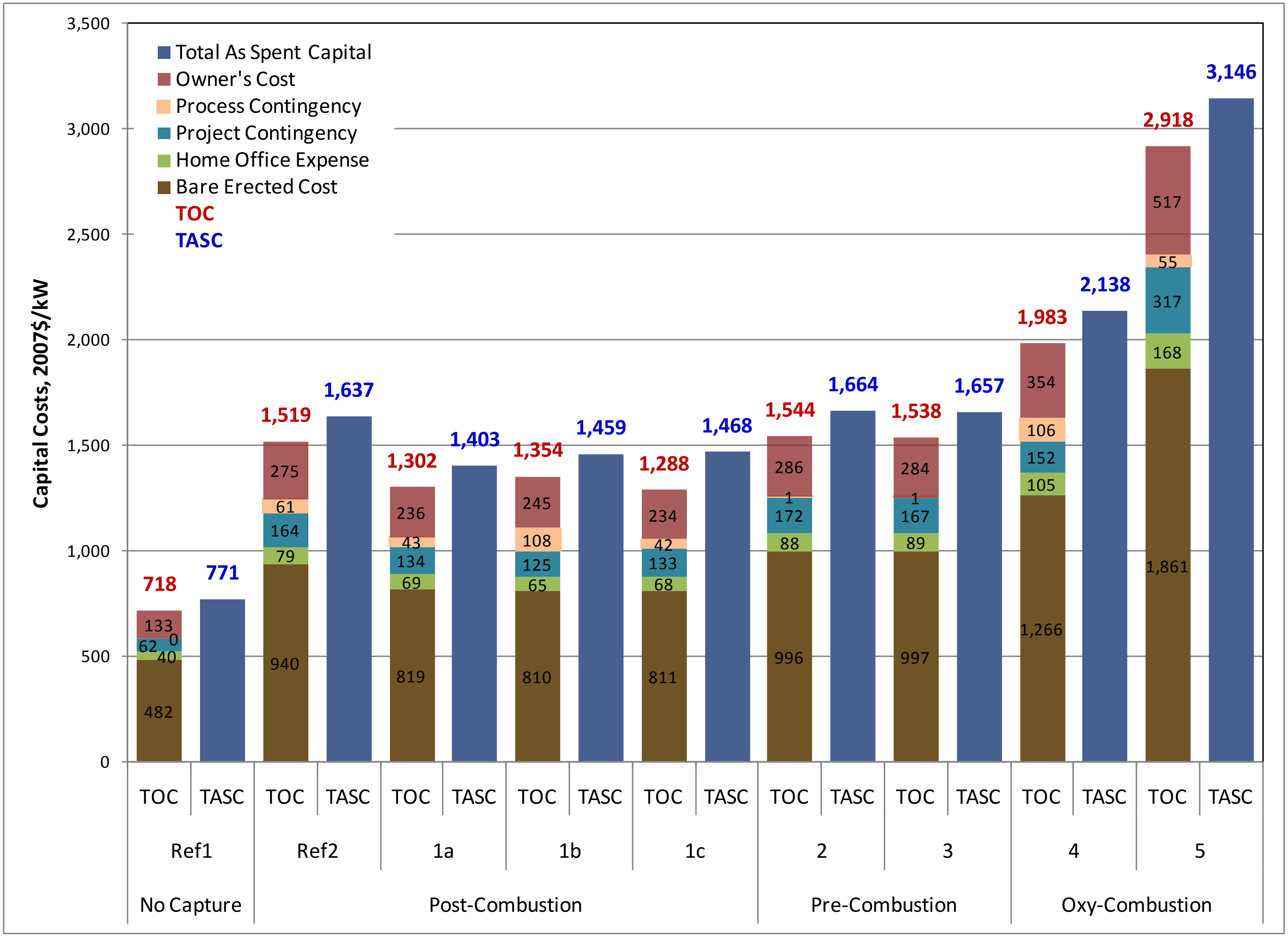
The COE with CO_2 removal includes the costs of capture and compression as well as TS&M costs.

4.8 FINAL COST ESTIMATING RESULTS

The normalized capital cost components for each technology are shown in Exhibit 4-8.

The cost estimates carry an accuracy of -15%/+30%, consistent with a “feasibility study” level of design engineering applied to the various cases in this study. The value of the study lies not in the absolute accuracy of the individual case results but in the fact that all cases were evaluated under the same set of technical and economic assumptions. This consistency of approach allows meaningful comparisons among the cases evaluated.

Exhibit 4-8 Plant Capital Cost



The COE results are listed in Exhibit 4-9 and shown graphically in Exhibit 4-10. The capital cost, fixed operating cost, variable operating cost, and fuel cost shown separately. In the capture cases, the CO_2 transport, storage, and monitoring (TS&M) costs are also shown as a separate bar segment.

Exhibit 4-9 COE Component Details (mills/kWh or $/MWh) for All Cases

Case	Ref1	Ref2	1a	1b	1c	2	3	4	5
Capital	10.10	22.66	19.41	20.19	19.21	23.02	22.93	29.58	43.52
Fixed O&M	2.96	5.74	5.04	5.31	4.99	6.00	5.89	7.34	11.17
Variable O&M	1.32	2.60	1.89	2.04	1.89	2.74	2.41	2.74	3.90
Fuel	44.51	52.93	51.79	51.46	50.89	52.87	54.24	53.78	50.05
CO_2 TS&M total	0.0	3.25	3.08	3.03	3.03	2.80	2.61	3.25	3.60
Transport	*0.0*	*1.95*	*1.92*	*1.88*	*1.89*	*1.73*	*1.56*	*2.03*	*2.25*
Storage	*0.0*	*0.90*	*0.83*	*0.82*	*0.81*	*0.74*	*0.71*	*0.87*	*0.97*
Monitoring	*0.0*	*0.40*	*0.33*	*0.33*	*0.33*	*0.33*	*0.34*	*0.35*	*0.39*
COE Total	**58.90**	**87.17**	**81.22**	**82.02**	**80.01**	**87.44**	**88.08**	**96.69**	**112.24**
LCOE, total (including TS&M)	74.66	110.50	102.96	103.97	101.42	110.84	111.66	122.57	142.28

[1] CF is 85% for NGCC cases

It can be seen that Cases 1a, 1b, and 1c all have lower COE than that of the reference case. The efficiencies of Cases 1a, 1b, and 1c increase from 48.0%, 48.3%, to 48.8% (HHV) respectively. This reduces fuel consumption accordingly. The flue gas recycle also reduces the size of the amine CO_2 capture unit. This tends to reduce both COE due to fixed and variable O&M costs because the maintenance labor cost, the maintenance material cost, and the property taxes & insurance cost all are related to the overall plant cost.

The efficiencies of Cases 2 and 3 are lower compared to the reference case, only 47.0 percent and 45.8 percent respectively. This slightly increases the fuel related COE. In addition, the processes are more complicated with more pieces of key equipment which not only increase the capital costs but also increase the variable O&M costs. Besides the MDEA solution for CO_2 capture and removal, additional chemicals are required, such as expensive ZnO to remove sulfur from natural gas, water-gas-shift catalysts for WGS reactors, and reformer catalysts for syngas reformation.

Case 4 is very similar to Cases 1a, 1b, and 1c except that the ASU has replaced the amine CO_2 capture unit. Case 4 has a lower efficiency of 46.2 percent (LHV) and the ASU is also more expensive than the amine unit. Thus though Case 4 is competitive, its COE is higher.

Case 5 is the most expensive configuration with the capital cost almost twice that of the reference case. In addition to the expensive ASU, Case 5 also has a high temperature, high pressure gas generator, combustor, and a series of unique high temperature, high pressure turboexpanders. Because these pieces of equipment operate at conditions beyond current commercial offerings, large process contingencies were included during evaluation and this configuration has the greatest uncertainty in cost. Though the process is the most efficient, 49.6 percent (LHV), its COE is the highest compared to all the cases evaluated. Case 5 also has a proportionally larger condenser, water circulation pumps, and water treatment facility.

The COE with CO_2 removal includes the costs of capture and compression as well as TS&M costs. The resulting avoided costs are shown in Exhibit 4-11 for each of the cases in this study. The avoided costs for each capture case are calculated using the NGCC non-capture case (Ref1) as the reference and again with Bituminous Baseline Supercritical Pulverized Coal without CO_2 capture as the reference. [1]

Exhibit 4-10 COE Components for All Cases

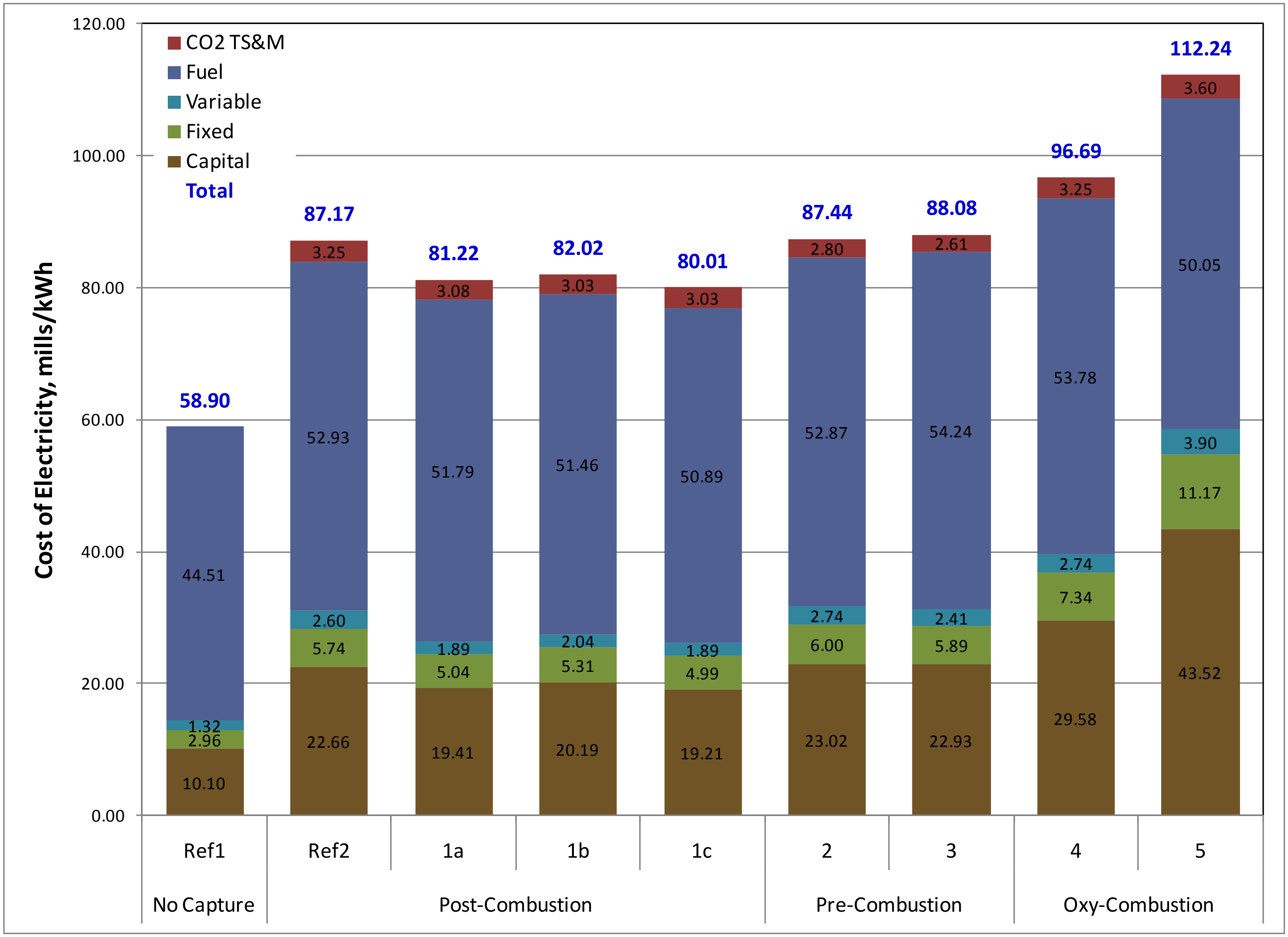
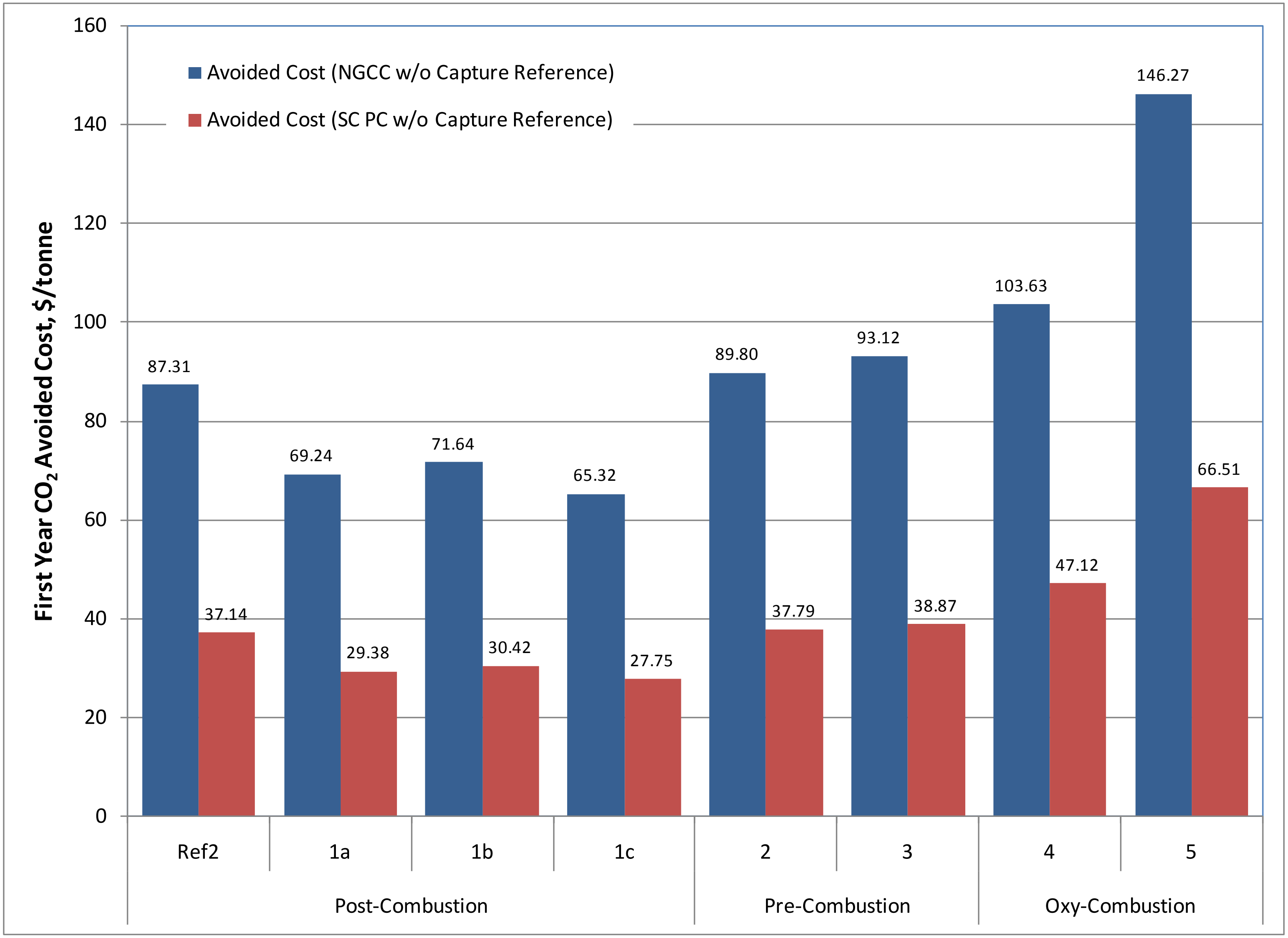


Exhibit 4-11 Summary and Comparison of CO_2 Avoided Costs for All Cases



4.9 EVALUATION OF KEY EQUIPMENT COSTS IN CASES 1 THROUGH 5 AND SENSITIVITY ANALYSES

Key equipment costs were evaluated on the basis of methodology discussed in previous sections. The costs for most of the equipment were obtained by scaling and applying factors from equations outlined in previous sections. Some of the unique equipment items with operating pressures and temperatures pushing or beyond limits offered commercially were estimated by WorleyParsons as discussed earlier.

4.9.1 Case 1a, Case 1b, and Case 1c

For Case 1a with 35 percent recycle, the gas turbine design was expected to change little. However, a 5 percent process contingency was added to Account 6.1 – Combustion Turbine Generator for unanticipated design modifications to the inlet air system and compressor. For Case 1b with 50 percent recycle, the exhaust gas oxygen content is less than the 4 percent recommended by the gas turbine manufacturer. The gas turbine combustor will have to be redesigned to improve combustion efficiency and minimize CO production. The gas turbine costs were escalated by 20 percent and a 40 percent process contingency was also added.

The cost of the amine-based CO_2 removal system was scaled on the basis of actual gas flow rate into the AGR with an exponent of 0.7.

The HRSG was assumed to require redesign to accommodate the change in composition and heat transfer properties of the recycle gas flow. For the 35 percent recycle case, Case 1a, the HRSG was scaled based on heat duty then both the HRSG equipment cost and direct labor were increased by 15 percent. Similarly, after scaling on heat duty, both the equipment and direct labor costs were escalated by 30 percent for the case with 50 percent recycle, Case 1b, because substantially more redesign was anticipated. Case 1c is similar to Case 1a except that the amine AGR unit was revised to improve operating efficiency.

The processes for Cases 1a, 1b, and 1c are relatively simple and require minimal modifications from the base case. The substantially reduced gas flow rate into the amine unit for Cases 1a, 1b, and 1c made the amine unit smaller and cheaper, which along with increased efficiency results in a lower COE for these cases relative to the reference case. If the costs of the HRSG and amine unit are doubled, the COE is slightly greater than the reference case in Case 1a and nearly equal to the reference case in Case 1b as shown in Exhibit 4-12 and Exhibit 4-13.

The operation of Case 1b with 50 percent flue gas recycle may be marginal because of low concentration of oxygen in the combustor. Further study is needed to verify the combustor efficiency and the CO emissions under the proposed operating conditions.

Exhibit 4-12 Sensitivity Analysis for Case 1a

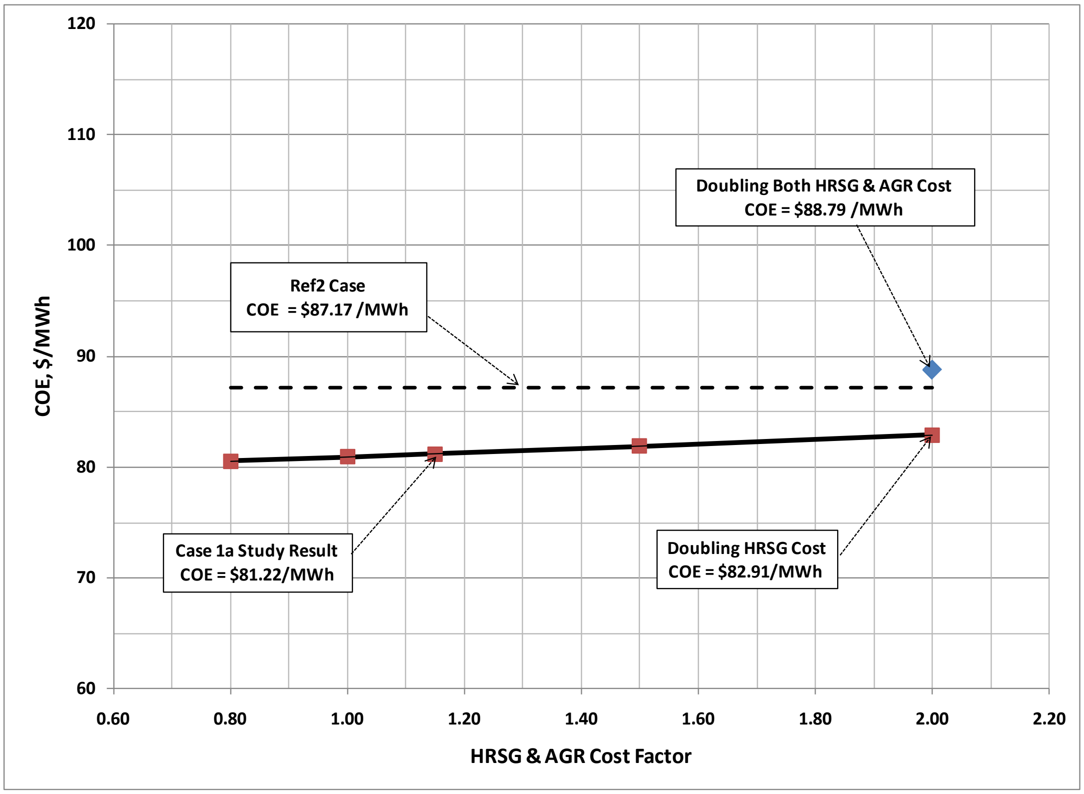
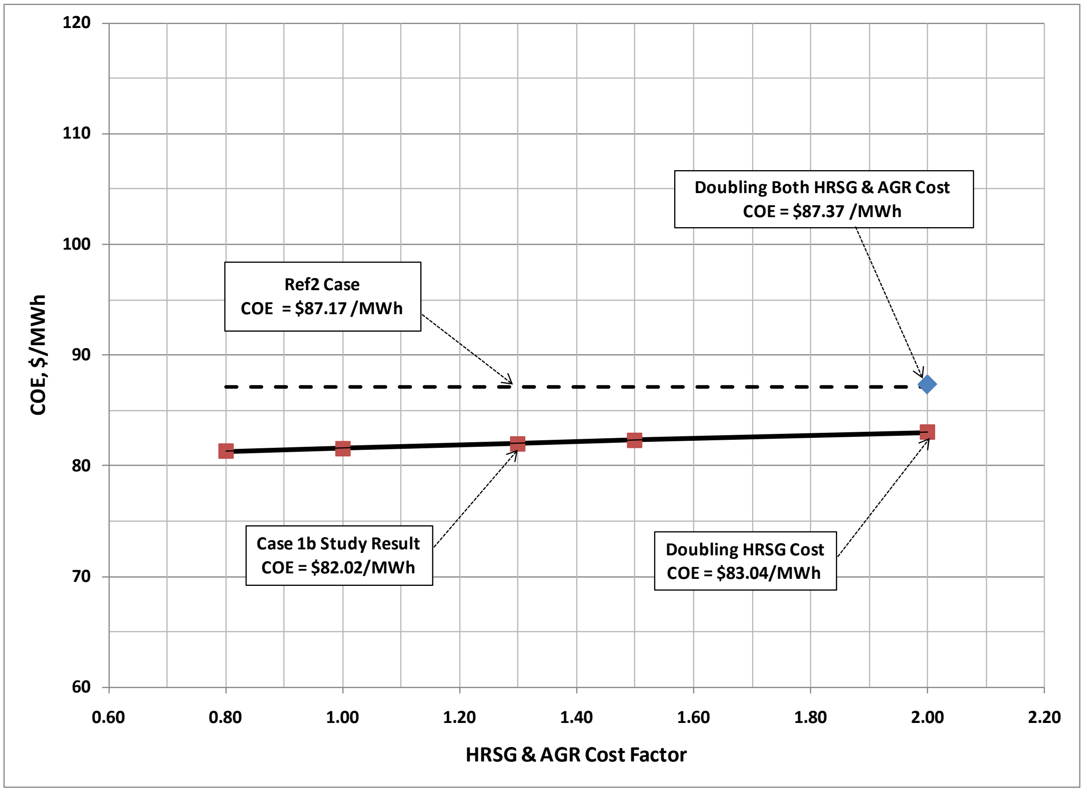


Exhibit 4-13 Sensitivity Analysis of Case 1b



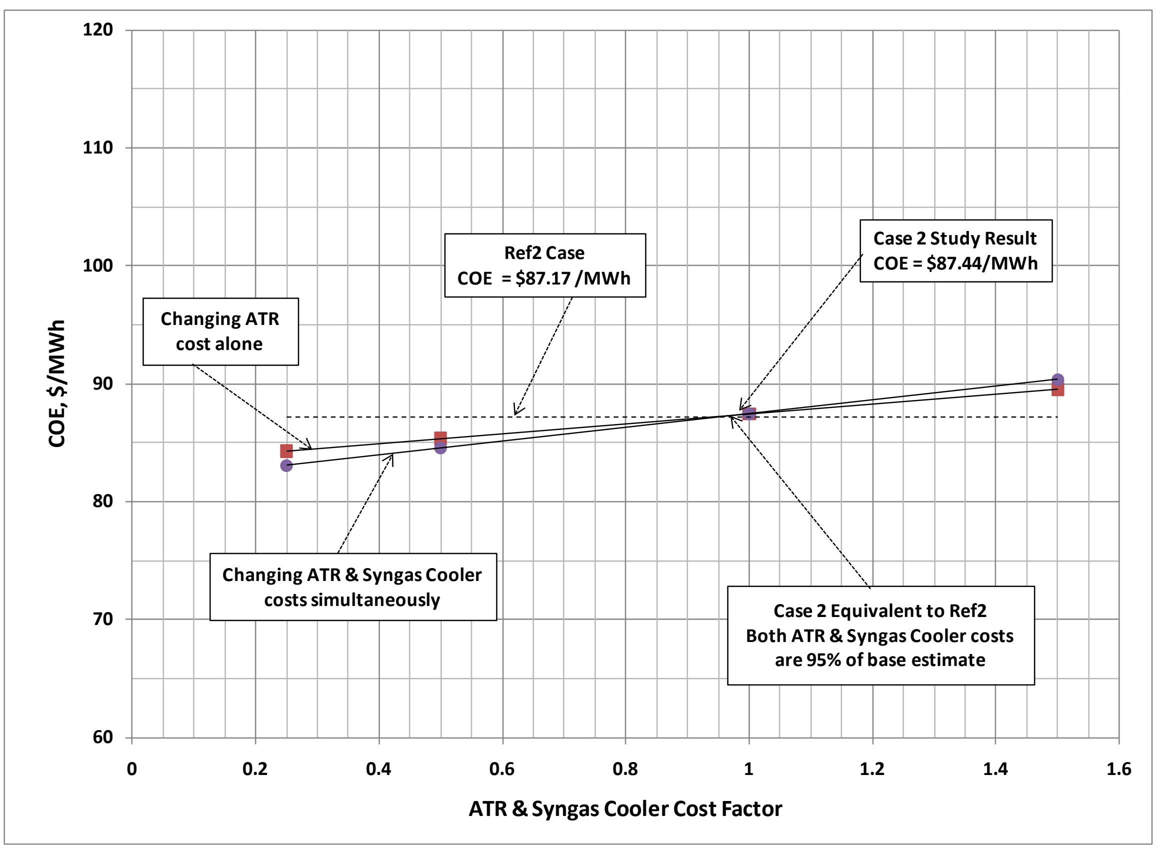
4.9.2 Case 2

The cost basis for unique equipment items was described in Section 4.1.

The fuel compressor cost was evaluated based on Equation (2) with the coefficients: $K_1 = 2.2897$; $K_2 = 1.3604$; and $K_3 = -0.1027$. The parameter in Equation (2), A, is the compressor capacity in kW. The resulting costs are for compressors constructed with carbon steel and operate at ambient pressure and temperature in 2001 dollars. The resulting costs were corrected from 2001 to 2007 $ using the CE plant cost index and multiplied by a material correction factor of 1.5 derived from that discussed in Section 4.1.

Sensitivity analysis for Case 2 was carried out by changing the cost of the ATR alone or by changing the cost of the ATR and Syngas Cooler simultaneously as shown in Exhibit 4-14. It can be seen that, under the current cost assumptions, the COE of Case 2 is essentially equal to that of the reference case. Reducing the cost of both the ATR and syngas cooler to 25 percent of estimated cost reduces the COE to $83/MWh and increasing the cost of the two equipment items by 50 percent of the estimated cost increases the COE to $90.5/MWh. Thus significant adjustment to the cost estimates of these equipment items results in a COE delta of only about 5 percent.

Exhibit 4-14 Sensitivity Analysis of Case 2



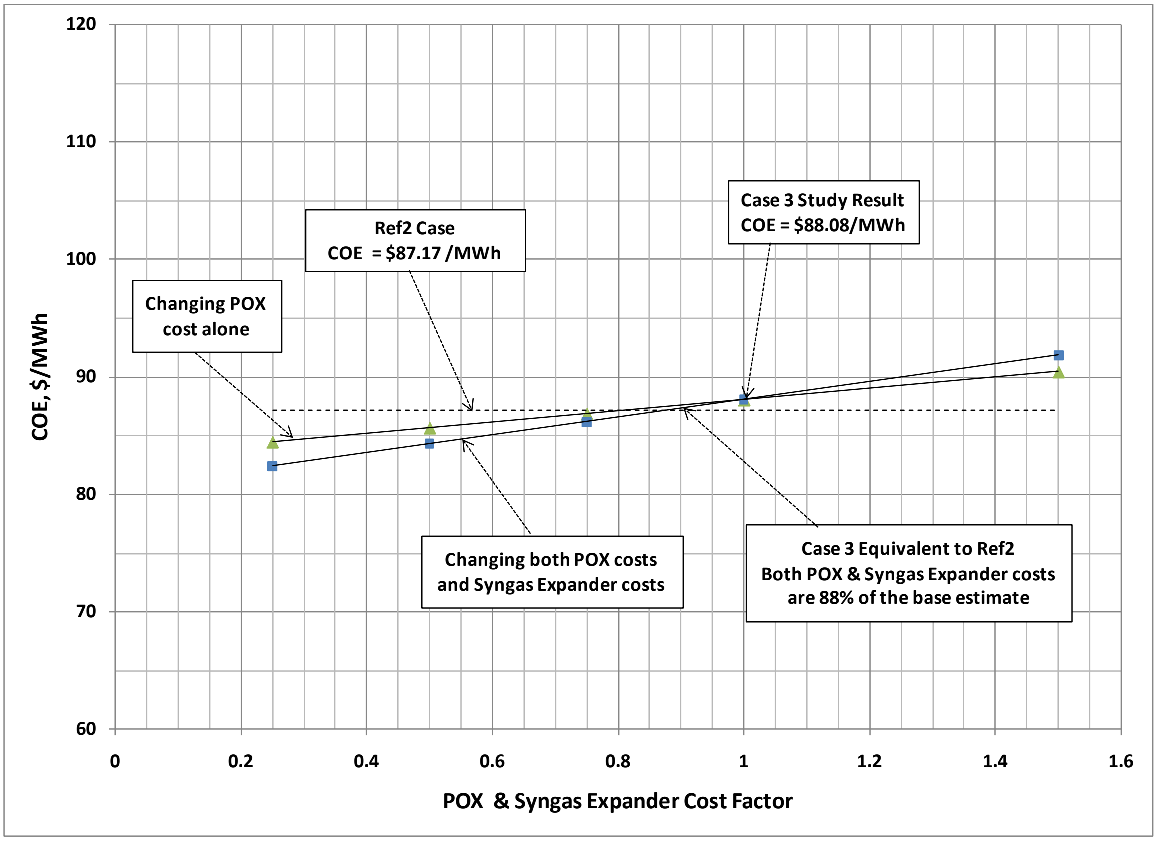
4.9.3 Case 3

The cost basis for unique equipment items was described in Section 4.1.

The air compressor costs were similarly calculated as the fuel compressor in Case 2.

Case 3 is very similar to Case 2. The major differences are that Case 3 has a partial oxidizer (POX) rather than an ATR and operates at even higher pressures. Case 3 also features a high temperature, high pressure syngas expander. The results of cost estimates indicate that the COE of Case 3 is slightly higher than that of Case 2. The benefits of operating at higher pressures did not overcome the costs incurred by increasing process complexity and increasing the number of major equipment items. The feasibility of operating the partial oxidizer, shift reactors, and AGR at the proposed higher pressure needs to be verified. The feasibility of operating the gas-gas heat exchanger under the proposed high temperature and high pressure conditions also needs to be studied. From Exhibit 4-15, it can be seen that with a reduction of costs of the POX reactor and Syngas Expander by ten percent, the COE of Case 3 drops slightly below that of the reference case. Reducing the cost of the two equipment items to 25 percent of the base estimate results in a COE 7 percent less than the reference cost while increasing the cost to 150 percent of the base estimate results in a COE 6 percent greater than the reference cost.

Exhibit 4-15 Sensitivity Analysis of Case 3



4.9.4 Case 4

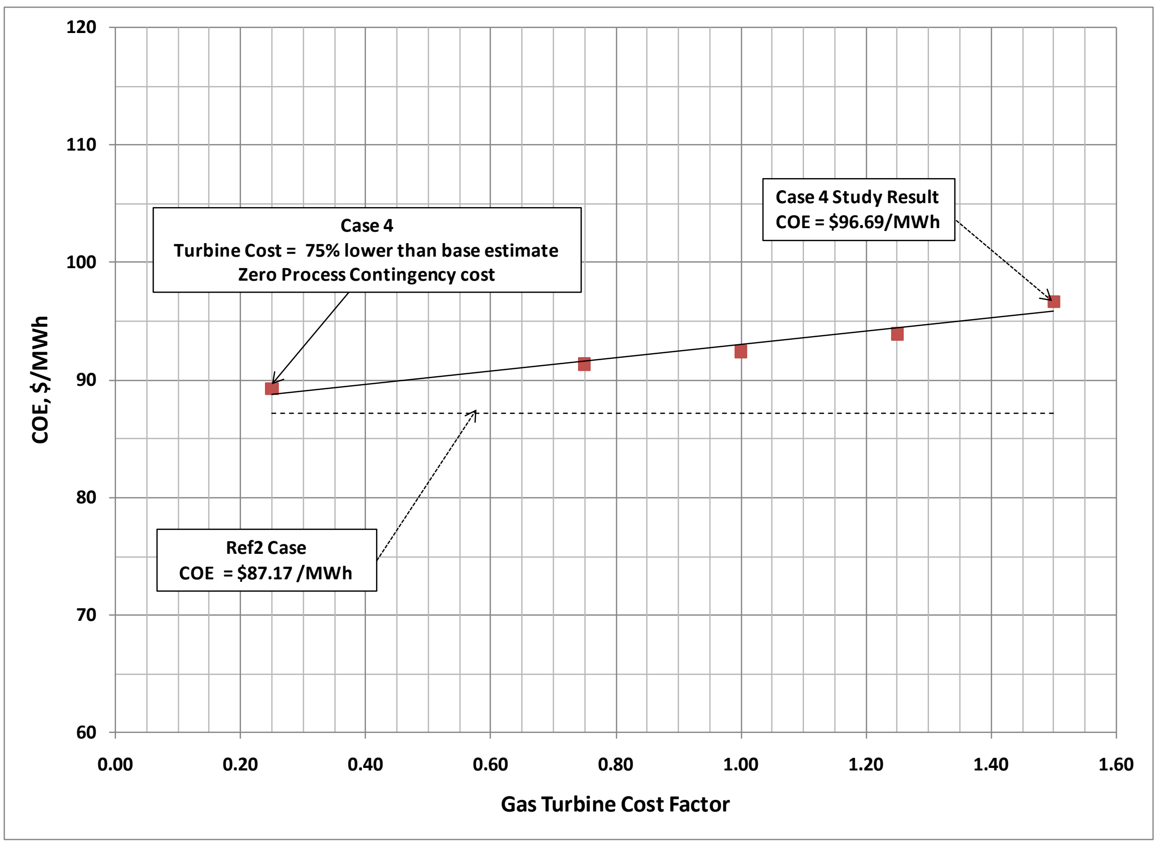
No cost correction was applied to the HRSG in this case. The flow rate through the HRSG is actually 20 percent less than the reference case.

Case S1A from the Low Rank Coal Study [21] was employed as a reference case for cost estimates for the ASU. The costs were scaled on the basis of air compressor power with an exponent of 0.57 derived from a correlation established from analyzing the costs of Cases S1A, S1B, L1A, and L1B of the Low Rank Coal Study [21].

The fuel compressor cost was calculated as discussed in Section 4.1.1 with a cost index correction factor of 1.34 (to 2007 dollars) and a material correction factor of 2.5.

The high pressure ratio combustion turbine was assumed to be 50 percent more expensive than that of the advanced 7FB and a process contingency of 40 percent was also added. The results gave a COE of Case 4 slightly higher than that of the reference case (see Exhibit 4-16) but the sensitivity analysis shows that even with a significant reduction (75 percent) in the turbine costs and elimination of the process contingency, the COE for Case 4 is still higher than the reference case.

Exhibit 4-16 Sensitivity Analysis of Case 4



4.9.5 Case 5

The ASU was scaled similarly as that in Case 4 employing Case S1A from the Low Rank Coal Study as a reference case with an exponent of 0.57 [21].

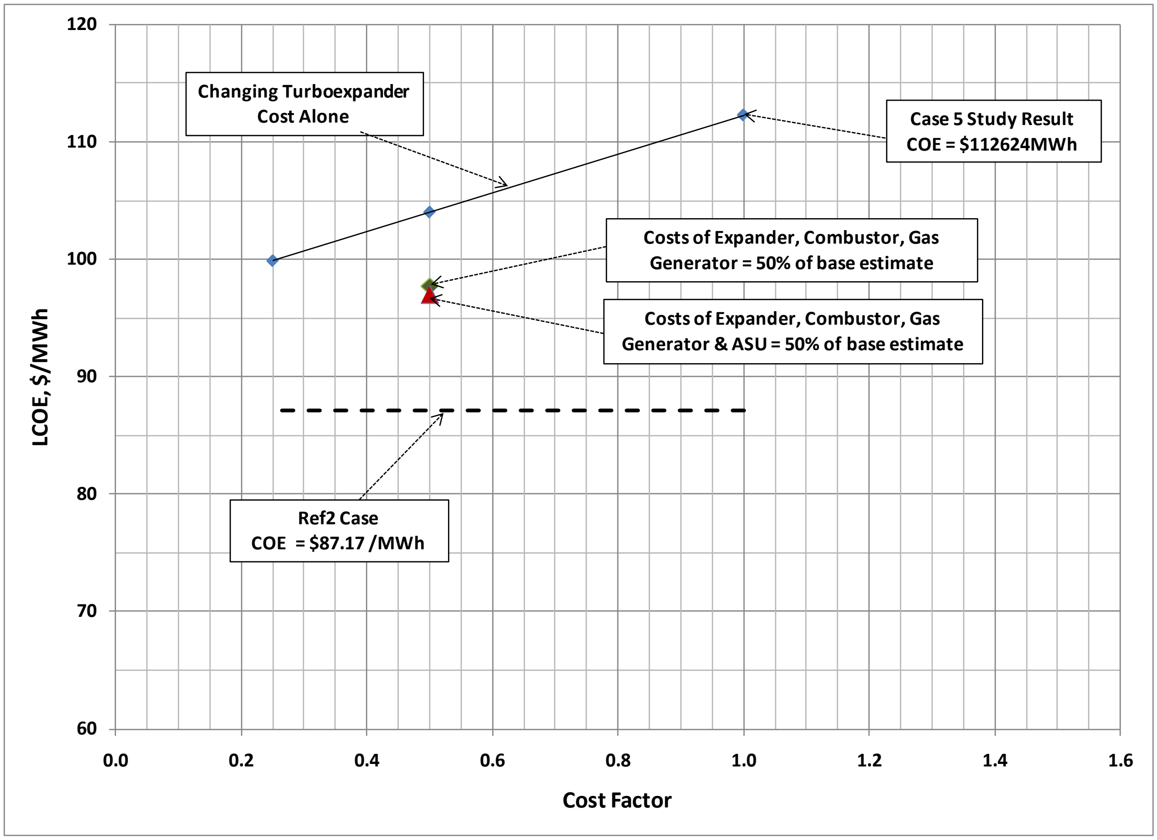
The cost of the gas generator was evaluated using reference cost for the CoP gasifier from Case 3 of the Bituminous Baseline Study [1]. Syngas cooling costs were removed from the lumped gasifier costs before scaling.

The costs of the high temperature turboexpander were provided by WorleyParsons as discussed earlier. Both the costs of the combustion turbine generator and steam turbine generator were removed and replaced with the costs of the turboexpander. The cost of the high temperature combustor was assumed to be 15 percent of the cost of combustion turbine. The costs of the cooling tower and HRSG were also removed while the condenser and auxiliary costs remained. Condenser costs were scaled on the basis of heat duty with an exponent of 0.7.

The fuel compressor costs were evaluated following the same methodology employed in Case 4 with a cost index correction factor of 1.34 and a material correction factor of 2.5.

Case 5 has the greatest cost uncertainty and is substantially more expensive than all other cases. It also introduces some one-of-a-kind equipment items, such as the high temperature, high pressure turboexpander, gas generator, and reheat combustor. The sensitivity analysis shown in Exhibit 4-17 indicates that even with the turboexpander cost reduced to 25 percent of assumed cost; or reducing the costs of the turboexpander, reheat combustor, and gas generator to 50 percent of assumed costs; or reducing the costs of the turboexpander, reheat combustor, gas generator, and ASU to 50 percent of assumed costs; the COE of Case 5 is still much higher than that of the reference case.

Exhibit 4-17 Sensitivity Analysis of Case 5



5. CONCLUSIONS

A summary table showing quantitative (performance estimates) and qualitative (maturity, level of complexity, and technology status) items for all cases is shown in Exhibit 5-1.

The following conclusions were drawn from the performance:

- The post-combustion CO_2 capture processes using EGR and the CES-based system have higher net efficiencies than the reference capture case, which consists of conventional NGCC and post-combustion amine capture. The pre-combustion capture cases and the conventional oxy-combustion case are approximately equal to or less than the reference capture case in net efficiency.
- The efficiency spread is only 3.4 absolute percentage point between the highest and lowest efficiency capture technology (HHV basis).
- The normalized water demand varies from 6.5 to 9.6 gpm/MWnet for the capture cases. At the low end of the range is case 3, high pressure partial oxidation with amine CO_2 separation, primarily because the additional power recovered in the syngas expander results in the highest net output of all the cases as opposed to a significant reduction in water demand. At the high end of the range is the post-combustion capture reference case (Ref2), which has a relatively high water demand because of the Econamine process and relatively low net output.
- Raw water consumption for the capture cases varies over a narrower range than demand, 4.8 to 6.4 gpm/MWnet, primarily because the technologies with high demand also had high process water discharge which serves to reduce consumption.

A summary table showing the cost estimation results for all cases is shown in Exhibit 5-2.

The cost results and sensitivity analyses lead to the following conclusions:

- Case 1c with 35 percent EGR and reduced amine reboiler duty has the lowest COE of all CO_2 capture cases, including the reference case.
- All three cases employing EGR have comparable and relatively low COEs. The simple modification of providing various amounts of flue gas recycle performs much better compared to other cases evaluated. Cases 1a, 1b, and 1c employ conventional technologies which can be readily implemented commercially with minimum modifications. The optimal amount of recycled flue gas and feasibility based on flue gas oxygen content still needs to be determined.
- The pre-combustion capture cases (Case 2 and Case 3) are quite similar. Case 3 features a high temperature, high pressure syngas expander that operates outside the range of the currently available commercial machines. The process cycles are quite complicated with gas-gas heat exchangers, syngas cooling, shift reactors, air compressors, and fuel gas expanders, in addition to the AGR or POX. The number of equipment items and process complexity result in higher COEs than the EGR cases, but is still competitive with the reference case COE. Remaining to be verified are the feasibility and availability of reactor operations including the ATR, POX, shift reactors, and AGR at high temperature and high pressure operating conditions.
- Substituting oxygen for air results in the highest COEs of the cases examined in this study. Case 4, a more conventional oxy-combustion concept, has a COE that is

11 percent higher than the reference capture case. The ASU is more costly than the amine CO_2 removal system it replaces and thus is also more costly than Cases 1a, 1b, and 1c.

- The CES-based system was the most expensive in the current study. The case contains several expensive pieces of equipment, including the ASU, the very high temperature, high pressure gas generator, and reheat combustor, and the high temperature, high pressure turboexpander. This process also had the highest degree of cost uncertainty. Even when reducing the cost of the ASU, the gas generator, the reheat combustor and the expander by 50 percent, the COE for Case 5 was still greater than the reference case. In addition, Case 5 generates the lowest net power compared to other cases, which also negatively impacts the COE.

The cases examined in this study largely show promise for reducing CO_2 emissions at costs that are lower than simply applying a post-combustion amine system to a conventional NGCC system. The EGR technologies are relatively simple and result in reduced capital cost as well as increased efficiency, which in turn translate to lower COE compared to the reference CO_2 capture case. The pre-combustion cases operate at a different pressure profile than the post-combustion cases using significantly different equipment items, but still result in a COE that is nearly equivalent to the reference case. The difference in the COE capital cost component between the pre-combustion cases and the reference case is small. The oxy-combustion based systems are the most expensive of the technologies examined in this study. Uncertainty in cost for specific equipment items contributed to the increased COE. However, the cost sensitivity analysis showed that even when greatly discounting the cost of the uncertain equipment items, the COE still remained greater than the reference case.

Exhibit 5-1 Summary Performance Table

Case	Ref1	Ref2	Ref2	1a	1b	1c	2	3	4	5
Description	Non-capture	Capture	Modified	Post-combustion with 35% exhaust gas recycle	Post-combustion with 50% exhaust gas recycle	Post-combustion with 35% exhaust gas recycle & lower amine system steam	Pre-combustion ATR	Pre-combustion high pressure POX-GE configuration	Oxy-combustion with CO_2 recycle	CES-Based Oxy-combustion with water/steam recycle
	Plant Output									
Gas Turbine Power (kWe)	362,200	362,200	362,200	361,600	359,800	361,600	384,227	456,776	362,100	
Steam Turbine Power (kWe)	202,500	148,800	148,800	151,400	155,700	160,200	260,973	172,422	223,800	
Syngas Expander (kWe)								97,950		
Total (kWe)	**564,700**	**511,000**	**511,000**	**513,000**	**515,500**	**521,800**	**645,200**	**727,148**	**585,900**	**528,700**
	Auxiliary Load									
ASU Compression (kWe)									70,740	50,920
Fuel Gas Compression (kWe)							25,680			
Oxygen Compression (kWe)									28,520	26,740
Condensate Pumps (kWe)	170	80	80	80	90	100	250	160	160	70
Boiler Feedwater Pumps (kWe)	2,720	2,710	2,710	2,750	2,840	2,750	4,220	2,280	2,900	2,660
Amine System Auxiliaries (kWe)	0	9,600	16,364	10,637	8,182	10,637	1,000	1,200	NA	NA
CO2 Compression (kWe)	0	15,200	15,200	14,910	14,950	14,910	19,110	21,820	20,880	31,940
Circulating Water Pump (kWe)	2,300	4,360	4,360	4,110	4,150	4,310	3,430	2,900	5,260	3,086
Ground Water Pumps (kWe)	210	360	360	340	340	360	400	280	430	255
Cooling Tower Fans (kWe)	1,190	2,250	2,250	2,120	2,140	2,230	1,770	1,500	2,720	1,592
SCR (kWe)	10	10	10	10	10	10	10	10		
Turbine Auxiliaries (kWe)										829
Gas Turbine Auxiliaries (kWe)	700	700	700	1,000	1,000	1,000	700	700	1,000	
Steam Turbine Auxiliaries (kWe)	100	100	100	100	100	100	100	100	100	
Miscellaneous Balance of Plant (kWe)	500	500	500	500	500	500	500	500	500	500
Transformer Losses (kWe)	1,720	1,560	1,580	1,640	1,640	1,670	2,030	2,390	1,850	1,840
Air Compression (kWe)								37,300		
Natural Gas Compression (kWe)								6,310	1,410	2,340
Total	9,620	37,430	44,214	38,197	35,942	38,577	59,200	77,450	136,470	122,772
	Plant Performance									
Net Auxiliary Load (kWe)	9,620	37,430	44,214	38,197	35,942	38,577	59,200	77,450	136,470	122,772
Net Plant Power (kWe)	555,080	473,570	466,786	474,803	479,558	483,223	586,000	649,698	449,430	405,928
Net Plant Efficiency (HHV)	50.2%	42.8%	42.2%	43.2%	43.5%	44.0%	42.4%	41.3%	41.6%	44.7%
Net Plant Efficiency (LHV)	55.7%	47.5%	46.8%	48.0%	48.3%	48.8%	47.0% 47.6%[1] 45.9%[2] 46.9%[3] 49.7%[4]	45.8% 47.5%[5]	46.2% 47.0%[3] 46.2%[4]	49.6%
Net Plant Heat Rate (HHV) [kJ/kWhr (Btu/kWhr)]	7,172 (6,798)	8,406 (7,968)	8,528 (8,083)	8,326 (7,892)	8,274 (7,842)	8,181 (7,754)	8,500 (8,057)	8,720 (8,265)	8,646 (8,194)	8,046 (7,626)
Net Plant Heat Rate (LHV) [kJ/kWhr (Btu/kWhr)]	6,466 (6,129)	7,579 (7,184)	7,689 (7,288)	7,507 (7,115)	7,460 (7,070)	7,376 (6,991)	7,664 (7,264)	7,863 (7,452)	7,795 (7,388)	7,255 (6,876)
Natural Gas Feed Flow [kg/hr (lb/hr)]	75,901 (167,333)	75,901 (167,333)	75,901 (167,333)	75,374 (166,172)	75,648 (166,774)	75,374 (166,172)	94,971 (209,375)	108,022 (238,148)	74,083 (163,325)	62,272 (137,285)
Thermal Input (HHV) [kWth]	1,105,812	1,105,812	1,105,812	1,098,140	1,102,121	1,098,140	1,383,644	1,573,791	1,079,327	907,255
Thermal Input (LHV) [kWth]	997,032	997,032	997,032	990,114	993,704	990,114	1,247,533	1,418,975	973,152	818,007
(MMBtu/hr)	1,139 (1,080)	528 (500)	528 (500)	549 (520)	570 (540)	654 (620)	1,340 (1,270)	1,087 (1,030)	1,139 (1,080)	1,245 (1,180)
Raw Water Withdrawal [m3/min (gpm)]	8.9 (2,362)	15.1 (3,980)	15.1 (3,980)	14.2 (3,741)	14.1 (3,729)	14.9 (3,944)	16.8 (4,430)	11.8 (3,104)	18.2 (4,801)	10.6 (2,801)
Raw Water Consumption [m3/min (gpm)]	6.9 (1,831)	11.3 (2,985)	11.3 (2,985)	10.6 (2,802)	10.5 (2,781)	11.2 (2,959)	13.8 (3,638)	9.2 (2,434)	13.6 (3,584)	7.8 (2,056)
Removal Efficiency	0.0%	90.7%	90.7%	90.5%	90.4%	90.5%	89.6%	88.9%	>99%	>99%
Maturity	Mature	Developmental	Developmental	Conceptual	Conceptual	Conceptual	Conceptual	Conceptual	Conceptual	Conceptual
Level of Complexity	Low	Low	Low	Low	Low	Low	High	High	Low	Medium
Technology Status	Commercial	Demonstrated (on coal fired plants)	Demonstrated (on coal fired plants)	Future development: 6-10 years	Future development: 6-10 years	Future development: 6-10 years	Future development: 6-10 years	Future development: 6-10 years	Future development: 6-10 years	Future development: 6-10 years

[1]Analysis of Gas-Steam CC w/NG reforming & CO_2 Capture: Corradetti & Desideri

[2]NG fired Power plants w/CO_2 Capture: Kvamsdal , Anderson, & Bolland

[3]A quantitative comparison of gas turbine cycles w/CO_2 capture: Kvamsdal, Jordal, & Bolland

[4]CO_2 Capture from NGCC: Lozza, Chiesa, Romanco, & Valenti

[5]Performance & Cost Analysis of Advanced Gas Turbine Cycles w/pre-combustion CO_2 capture: Hoffmann, Bartlett, Finkenrath, Evulet, & Ursin

Exhibit 5-2 Summary Cost Results

Case	Ref1	Ref2	1a	1b	1c	2	3	4	5
Capital Costs ($/kW)									
Bare Erected Cost	482	940	819	810	811	996	997	1,266	1,861
Home Office Expense	40	79	69	65	68	88	89	105	168
Process Contingency	0	61	43	108	42	1	1	106	55
Project Contingency	62	164	134	125	133	172	167	152	317
Owner's Cost	133	275	236	245	234	286	284	354	517
TOC	718	1,519	1,302	1,354	1,288	1,544	1,538	1,983	2,918
Total As Spent Capital	771	1,637	1,403	1,459	1,468	1,664	1,657	2,138	3,146
COE, (mills/kWh)									
Capital	10.10	22.66	19.41	20.19	19.21	23.02	22.93	29.58	43.52
Fixed O&M	2.96	5.74	5.04	5.31	4.99	6.00	5.89	7.34	11.17
Variable O&M	1.32	2.60	1.89	2.04	1.89	2.74	2.41	2.74	3.90
Fuel	44.51	52.93	51.79	51.46	50.89	52.87	54.24	53.78	50.05
CO_2 TS&M total	0.0	3.25	3.08	3.03	3.03	2.80	2.61	3.25	3.60
Transport	*0.0*	*1.95*	*1.92*	*1.88*	*1.89*	*1.73*	*1.56*	*2.03*	*2.25*
Storage	*0.0*	*0.90*	*0.83*	*0.82*	*0.81*	*0.74*	*0.71*	*0.87*	*0.97*
Monitoring	*0.0*	*0.40*	*0.33*	*0.33*	*0.33*	*0.33*	*0.34*	*0.35*	*0.39*
COE Total	**58.90**	**87.17**	**81.22**	**82.02**	**80.01**	**87.44**	**88.08**	**96.69**	**112.24**
LCOE, total (including TS&M)	74.66	110.50	102.96	103.97	101.42	110.84	111.66	122.57	142.28
Cost of CO_2 Avoided ($/tonne) NGCC w/o Capture as Reference		87.31	69.24	71.64	65.32	89.80	93.12	103.63	146.27
Cost of CO_2 Avoided ($/tonne) SC PC w/o Capture as Reference		37.14	29.38	30.42	27.75	37.79	38.87	47.12	66.51

APPENDIX A - ACCOUNT BY ACCOUNT COMPARISON OF COSTS FOR ALL CASES

Exhibit A-1 Comparison of Total Plant Costs

		Ref2 Case	Case 1a	Case 1b	Case 1c	Case 2	Case 3	Case 4	Case 5
Cost Base (June, 2007) = $x1,000		Total Plant Cost							
Acct No.	Item/Description								
3	**FEEDWATER & MISC. BOP SYSTEMS**								
3.1	Feedwater System	8,690	8,803	9,011	8,803	12,169	7,612	9,208	9,479
3.2	Water Makeup & Pretreating	4,497	4,305	4,295	4,469	4,853	3,770	5,140	6,186
3.3	Other Feedwater Subsystems	2,229	2,258	2,312	2,258	3,122	1,953	2,362	2,432
3.4	Service Water Systems	3,188	3,051	3,044	3,167	3,440	2,672	3,643	4,384
3.5	Other Boiler Plant Systems	4,798	4,592	4,581	4,767	5,177	4,022	5,483	6,598
3.6	Natural Gas, incl. pipeline	19,389	19,372	19,388	19,372	19,987	20,587	19,295	18,202
3.7	Waste Treatment Equipment	1,569	1,502	1,499	1,560	1,694	1,316	1,794	2,159
3.8	Misc. Equip.(Cranes,AirComp.,Comm.)	1,951	1,948	1,950	1,948	2,010	2,071	1,941	1,831
	SUBTOTAL 3.	**46,312**	**45,831**	**46,080**	**46,346**	**52,451**	**44,003**	**48,867**	**51,270**
4	**GASIFIER & ACCESSORIES**								
4.1	Gasifier, Syngas Cooler & Auxiliaries	0	0	0	0	0	0	0	0
4.2	Syngas Cooling	0	0	0	0	0	0	0	0
4.3	ASU/Oxidant Compression	0	0	0	0	0	0	269,594	154,838
4.4	LT Heat Recovery & FG Saturation	0	0	0	0	56,392	31,413	0	0
4.5	Misc. Gasification Equipment	0	0	0	0	0	0	0	0
4.6	Other Gasification Equipment	0	0	0	0	0	0	0	0
4.8	Major Component Rigging	0	0	0	0	0	0	0	0
4.9	Gasification Foundations	0	0	0	0	0	0	0	0
	SUBTOTAL 4.	**0**	**0**	**0**	**0**	**56,392**	**31,413**	**269,594**	**154,838**
5A	**GAS CLEANUP & PIPING**								
5A.1	ATR (Case 2), POX (Case 3), Gas Generator (Case 5)	0	0	0	0	105,682	132,834	0	182,463
5A.2	Syngas Cooling	0	0	0	0	40,347	42,899	0	0
5A.3	Gas-Gas Heat Exchangers	0	0	0	0	33,901	73,321	0	4,136

		Ref2 Case	Case 1a	Case 1b	Case 1c	Case 2	Case 3	Case 4	Case 5
5A.4	WGS Reactors	0	0	0	0	26,338	20,023	0	0
5A.5	Zinc Oxide Guard Bed	0	0	0	0	1,138	1,246	0	0
5A.7	Fuel Gas Compressors	0	0	0	0	0	0	2,953	0
5A.9	HGCU Foundations	0	0	0	0	0	0	0	0
	SUBTOTAL 5A.	**0**	**0**	**0**	**0**	**207,406**	**270,324**	**2,953**	**186,599**
5B	**CO_2 REMOVAL & COMPRESSION**								
5B.1	CO_2 Removal System	215,943	121,957	100,444	121,957	64,087	41,818	0	0
5B.2	CO_2 Compression & Drying	24,390	34,993	35,056	34,993	28,377	29,256	28,960	31,936
	SUBTOTAL 5B.	**240,334**	**156,950**	**135,500**	**156,950**	**92,464**	**71,074**	**28,960**	**31,936**
6	**COMBUSTION TURBINE/ACCESSORIES**								
6.1	Combustion Turbine Generator	95,589	99,593	148,710	99,593	95,588	95,588	180,333	0
6.2	Combustion Turbine Accessories	0	0	0	0	0	0	0	0
6.3	Compressed Air Piping	0	0	0	0	0	0	0	0
6.4	Air or Fuel Compressors (Cases 2, 3, & 5)	0	0	0	0	4,931	14,401	0	2,854
6.5	Syngas Expanders (Case 3) or Turboexpander (Case 5)	0	0	0	0	0	77,533	0	354,384
6.6	HT Combustor (Case 5)	0	0	0	0	0	0	0	39,491
6.9	Combustion Turbine Foundations	1,901	1,902	1,902	1,902	1,902	1,902	1,902	1,902
	SUBTOTAL 6.	**97,490**	**101,495**	**150,612**	**101,495**	**102,421**	**189,424**	**182,235**	**398,631**
7	**HRSG, DUCTING & STACK**								
7.1	Heat Recovery Steam Generator	44,448	45,874	43,740	45,874	46,543	46,197	8,957	0
7.2-7.9	SCR System, Ductwork, Stack, & Foundations	3,120	2,800	2,362	2,800	3,267	3,243	629	0
7.3	Ductwork	0	0	0	0	0	0	0	0
7.4	Stack	0	0	0	0	0	0	0	0
7.9	HRSG,Duct & Stack Foundations	1,057	798	657	798	1,318	1,403	122	0
	SUBTOTAL 7.	**48,624**	**49,472**	**46,759**	**49,472**	**51,128**	**50,843**	**9,708**	**0**
8	**STEAM TURBINE GENERATOR**								
8.1	Steam TG & Accessories	27,110	27,445	27,997	28,569	40,398	30,100	36,223	0
8.2	Turbine Plant Auxiliaries	548	555	566	578	826	610	738	0
8.3	Condenser & Auxiliaries	3,711	3,658	3,756	4,114	6,849	5,894	6,085	6,480

		Ref2 Case	Case 1a	Case 1b	Case 1c	Case 2	Case 3	Case 4	Case 5
8.4	Steam Piping	8,281	8,373	8,529	8,436	11,489	7,282	8,762	9,012
8.9	TG Foundations	2,140	2,167	2,210	2,258	3,225	2,383	2,883	5,400
	SUBTOTAL 8.	**41,791**	**42,199**	**43,057**	**43,955**	**62,787**	**46,269**	**54,691**	**20,892**
9	**COOLING WATER SYSTEM**								
9.1	Cooling Towers	8,182	7,857	7,907	8,107	1,816	1,730	9,338	0
9.2	Circulating Water Pumps	2,232	2,257	2,272	2,330	1,998	1,786	2,665	2,929
9.3	Circ.Water System Auxiliaries	184	178	179	183	158	142	208	228
9.4	Circ.Water Piping	6,128	10,170	10,230	10,464	9,106	8,226	11,821	12,879
9.5	Make-up Water System	912	880	878	908	974	786	1,022	1,195
9.6	Component Cooling Water Sys	983	948	954	977	844	758	1,110	1,215
9.9	Circ.Water System Foundations	6,783	6,558	6,596	6,742	5,892	5,338	7,588	8,245
	SUBTOTAL 9.	**25,403**	**28,848**	**29,016**	**29,711**	**20,787**	**18,767**	**33,751**	**26,692**
11	**ACCESSORY ELECTRIC PLANT**								
11.1	Generator Equipment	10,028	10,051	10,080	10,153	11,507	11,338	10,871	10,232
11.2	Station Service Equipment	2,498	2,520	2,455	2,531	3,042	3,415	4,357	4,163
11.3	Switchgear & Motor Control	3,402	3,431	3,342	3,445	4,142	4,649	5,932	5,668
11.4	Conduit & Cable Tray	5,949	6,001	5,846	6,027	7,245	8,133	10,376	9,915
11.5	Wire & Cable	7,337	7,400	7,209	7,432	8,935	10,029	12,795	12,226
11.6	Protective Equipment	2,966	2,966	2,966	2,966	2,966	2,966	2,966	2,966
11.7	Standby Equipment	224	224	225	226	251	248	239	228
11.8	Main Power Transformers	12,937	13,184	13,184	13,430	16,473	15,100	15,100	20,137
11.9	Electrical Foundations	546	547	549	554	643	632	601	559
	SUBTOTAL 11.	**45,888**	**46,325**	**45,857**	**46,763**	**55,204**	**56,509**	**63,237**	**66,093**
12	**INSTRUMENTATION & CONTROL**								
12.1	IGCC Control Equipment	0	0	0	0	0	0	0	0
12.2	Combustion Turbine Control	0	0	0	0	0	0	0	0
12.3	Steam Turbine Control	0	0	0	0	0	0	0	0
12.4	Other Major Component Control	1,690	1,694	1,681	1,697	1,794	1,858	2,000	1,972
12.5	Signal Processing Equipment	0	0	0	0	0	0	0	0
12.6	Control Boards,Panels & Racks	497	498	494	499	528	546	588	580
12.7	Computer & Accessories	4,892	4,905	4,866	4,911	5,192	5,377	5,788	5,709
12.8	Instrument Wiring & Tubing	2,426	2,431	2,412	2,435	2,574	2,665	2,869	2,830

		Ref2 Case	Case 1a	Case 1b	Case 1c	Case 2	Case 3	Case 4	Case 5
12.9	Other I & C Equipment	5,813	5,827	5,781	5,835	6,169	6,388	6,876	6,783
	SUBTOTAL 12.	**15,318**	**15,356**	**15,235**	**15,376**	**16,257**	**16,834**	**18,121**	**17,874**
13	**IMPROVEMENTS TO SITE**								
13.1	Site Preparation	2,524	2,499	2,499	2,501	2,511	2,501	2,569	2,515
13.2	Site Improvements	2,472	2,447	2,447	2,449	2,459	2,449	2,516	2,463
13.3	Site Facilities	4,471	4,427	4,426	4,429	4,448	4,430	4,551	4,454
	SUBTOTAL 13.	**9,467**	**9,373**	**9,372**	**9,379**	**9,418**	**9,380**	**9,636**	**9,431**
14	**BUILDINGS & STRUCTURES**								
14.1	Combustion Turbine Area	447	447	447	447	447	447	447	447
14.2	Steam Turbine Building	4,756	4,656	4,655	4,662	4,704	4,664	4,939	4,718
14.3	Administration Building	962	950	950	951	956	951	984	957
14.4	Circulation Water Pumphouse	272	267	268	272	248	231	295	313
14.5	Water Treatment Buildings	1,089	1,042	1,040	1,082	1,175	913	1,244	1,498
14.6	Machine Shop	815	813	813	813	814	813	819	814
14.7	Warehouse	515	514	514	514	514	514	517	515
14.8	Other Buildings & Structures	166	166	166	166	165	166	167	166
14.9	Waste Treating Building & Str.	1,052	1,045	1,045	1,050	1,061	1,028	1,069	1,094
	SUBTOTAL 14.	**10,075**	**9,900**	**9,897**	**9,957**	**10,084**	**9,726**	**10,480**	**10,522**
	TOTAL PLANT COST	**580,701**	**505,749**	**531,385**	**509,404**	**736,801**	**814,567**	**732,234**	**974,777**
	OWNERS								
	Preproduction Costs								
	6 Months All Labor	4,163	3,858	4,1163	3,876	5,712	6,108	4,960	7,136
	1 Month Maintenance Materials	612	540	612	545	822	915	800	1,070
	1 Month Non-fuel Consumables	273	232	226	242	559	429	258	289
	1 Month Waste Disposal	0	0	0	0	0	0	0	0
	25% of 1 Months Fuel Cost at 100% CF	4,509	4,489	4,505	4,489	5,656	6,434	4,412	3,709
	2% of TPC	11,614	10,115	10,628	10,188	14,736	16,291	14,645	19,496
	SUBTOTAL Production Costs	**21,170**	**19,235**	**20,134**	**19,339**	**27,485**	**30,176**	**25,075**	**31,699**
	Inventory Capital								
	60 day supply of consumables at 100% CF	357	287	276	296	907	710	289	282

		Ref2 Case	Case 1a	Case 1b	Case 1c	Case 2	Case 3	Case 4	Case 5
	0.5% of TPC (spare parts)	2,904	2,529	2,657	2,547	3,684	4,073	3,661	4,874
	SUBTOTAL Inventory Capital	**3,260**	**2,816**	**2,933**	**2,843**	**4,591**	**4,783**	**3,951**	**5,156**
	Initial Cost for Catalyst and Chemicals	**823**	**390**	**307**	**390**	**4,930**	**4,930**	**0**	**47**
	Land	**300**	**300**	**300**	**300**	**300**	**300**	**300**	**300**
	Other Owner's Costs	**87,105**	**75,862**	**79,708**	**76,411**	**110,520**	**122,185**	**109,835**	**146,217**
	Financing Costs	**15,679**	**13,665**	**14,347**	**13,754**	**19,894**	**21,993**	**19,770**	**26,319**
	TOTAL OVERNIGHT COST	**709,039**	**618,008**	**649,113**	**622,441**	**904,522**	**998,934**	**891,165**	**1,184,515**
	TOTAL AS-SPENT COST	**764,344**	**704,529**	**739,989**	**709,583**	**1,031,155**	**1,138,785**	**1,015,928**	**1,350,5347**

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6. REFERENCES

1 Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2, DOE/NETL-2010/1397, November 2010.

2 NETL Power Systems Financial Model Version 5.0, December 2008
User Guide available at:
http://www.netl.doe.gov/business/solicitations/ssc2008/references/PSFM%20User%20Guide.pdf

3 Quality Guidelines for Energy System Studies, prepared by DOE NETL Office of Systems and Policy Support, June 4, 2003

4 Gross Domestic Product: Chain-type Price Index, Seasonally Adjusted, http://research.stlouisfed.org/fred2/series/GDPCTPI

5 Improvement in Power Generation with Post-Combustion Capture of CO_2, IEA Greenhouse Gas R&D Programme, Report PH4/33, November 2004

6 Reddy, Satish, Johnson, Dennis, and Gilmartin, John, "Fluor's Econamine FG Plus[SM] Technology for CO_2 Capture at Coal-Fired Power Plants," Power Plant Air Pollutant Control "Mega" Symposium, Baltimore, MD, August 2008

7 Updated Coal Power Systems Study, Pulverized Coal Boiler, Fluor Subcontract Report for Carbon Dioxide Capture Unit, Fluor, December 2005

8 Reddy, Satish, "Econamine FG PlusSM Technology for Post-combustion CO_2 Capture," 11th Meeting of the International Post-Combustion CO_2 Capture Network, May 20 – 21, 2008, Vienna, Austria

9 ElKady, A, Evulret, A, Brand, A Ursin, T, Lynghjem, A: Exhaust Gas Recirculation in LN F-Class Gas Turbines for Post-Combustion CO_2 Capture, GT2008-51152, ASME Turbo Expo 2008: Power for Land, Sea, and Air, June 9-13, 2008. Berlin, Germany.

10 Corradetti, A. and Desideri, U., "Analysis if Gas-Stream Combined Cycles with natural Gas Reforming and CO_2 Capture," Journal of Engineering for Gas Turbines and Power 127 (July 2005, pages 545-552)

11 Kvamsdal, H, Thormad, A, Bolland, O., " Natural Gas Fired Power Plants with CO_2 Capture-Process Integration for High Fuel-to-Electricity Conversion Efficiency"

12 IEA Greenhouse Gas R&D Program, "CO_2 Capture in Low Rank Coal Power Plants," Technical Study, Report Number 2006/1, January 2006.

13 Hoffman, S., Bartlett, M., Finkenrath, M., Evulet, A., Ursin, T., "Performance and Cost Analysis of Advanced Gas Turbine Cycles with Pre-combustion CO_2 Capture," Journal of Engineering for Gas Turbines and Power 131 (March 2009, pages 021701-1 to 021701-7)

14 Kvamsdal, H., Jordal, K., Bolland, O., "A Quantitative Comparison of Gas Turbine Cycles with CO_2 Capture."

15 Lozza, G., Chiesa, P., Romano, M., Valenti, G., "CO_2 Capture from Natural Gas Combined Cycles."

16 Assessment of Hydrogen Production with CO_2 Capture, Volume 1: Baseline State-of-the-Art Plants, DOE/NETL-401/082410, August 24, 2010

17 Turton, R., Bailie, R. C., Whiting, W. B., and Shaeiwitz, A. in Analysis, Synthesis, and Design of Chemical Processes, 3rd Ed., Prentice Hall: Upper Saddle River, New Jersey, 2009, 923-953.

18 "Conducting Technical and Economic Evaluations – As Applied for the Process and Utility Industries; TCM Framework: 3.2 – Asset Planning, 3.3 Investment Decision Making, AACE International Recommended Practice 16R-90, 2003

19 Tennessee Valley Authority, NETL, and EPRI. "Economic Evaluation of CO_2 Storage and Sink Enhancement Options," December 2002.

20 "Recommended Project Finance Structures for the Economic Analysis of Fossil-Based Energy Projects," DOE/NETL-401/090808, September 2008

21 Cost and Performance Baseline for Low-rank coal Fossil Energy IGCC Power Plants with and without CO_2 Capture, 2010, Final Report, DOE/NETL-2010/1399.