



U.S. DEPARTMENT OF
ENERGY

National Energy
Technology Laboratory

OFFICE OF FOSSIL ENERGY



Cost and Performance Baseline for
Fossil Energy Plants
Volume 1b: Bituminous Coal (IGCC) to
Electricity
Revision 2b – Year Dollar Update

July 31, 2015

DOE/NETL-2015/1727

Disclaimer

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference therein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed therein do not necessarily state or reflect those of the United States Government or any agency thereof.

Author List:

National Energy Technology Laboratory (NETL)

Tim Fout

General Engineer

Office of Program Performance & Benefits

Energy Sector Planning and Analysis (ESPA)

*Alexander Zoelle, Dale Keairns, Marc Turner, Mark Woods, Norma Kuehn,
Vasant Shah, Vincent Chou*

Booz Allen Hamilton

Lora Pinkerton

WorleyParsons Group, Inc.

This report was prepared by Energy Sector Planning and Analysis (ESPA) for the United States Department of Energy (DOE), National Energy Technology Laboratory (NETL). This work was completed under DOE NETL Contract Number DE-FE0004001. This work was performed under ESPA Task 341.03, Activity 17.

Acknowledgments

The authors wish to acknowledge the excellent guidance, contributions, and cooperation of the NETL staff, particularly:

Tim Fout, NETL Technical Monitor

Kristin Gerdes
Director of OPPB

Travis Shultz
Director, Performance Division of OPPB

John Wimer
Director of OCPR&D

James Black

Daniel Cicero
formerly of NETL

Jared Ciferno

Eric Grol

Jeffrey Hoffmann

Julianne Klara

Patrick Le

Michael Matuszewski
formerly of NETL

Sean Plasynski

Larry Rath
formerly of NETL

Walter Shelton

Gary Stiegel
formerly of NETL

William Summers

Thomas J. Tarka, P.E.

Maria Vargas

John L. Haslbeck
formerly of Booz Allen Hamilton

Eric G. Lewis
formerly of Booz Allen Hamilton

Pamela Capicotto
formerly of Parsons Corporation

Michael Rutkowski
formerly of Parsons Corporation

Ronald Schoff
formerly of Parsons Corporation

Thomas Buchanan
formerly of WorleyParsons Group, Inc.

James Simpson
WorleyParsons Group, Inc.

Elsy Varghese
formerly of WorleyParsons Group, Inc.

Vladimir Vaysman
WorleyParsons Group, Inc.

Richard Weinstein
WorleyParsons Group, Inc.

DOE Contract Number DE-FE0004001

Table of Contents

Executive Summary	1
Results Analysis.....	3
1 Introduction.....	10
2 General Evaluation Basis	13
2.1 Site Characteristics.....	13
2.2 Coal Characteristics	15
2.3 Environmental Targets.....	16
2.3.1 Study Cases	18
2.3.2 Carbon Dioxide.....	19
2.4 Capacity Factor	19
2.4.1 Capacity Factor Assumptions	19
2.4.2 Capacity Factor for IGCC Plants	19
2.4.3 Capacity Factor for Plants with Carbon Capture	20
2.4.4 Perspective	20
2.5 Raw Water Withdrawal and Consumption	20
2.6 Cost Estimating Methodology	21
2.6.1 Capital Costs	22
2.6.2 Operation and Maintenance Costs	25
2.6.3 CO ₂ Transport and Storage	27
2.6.4 Cost of CO ₂ Captured and Avoided.....	29
2.7 Differences in IGCC Study Cost Estimates Compared to Industry Estimates	30
2.7.1 Project Scope	30
2.7.2 Labor	31
2.7.3 Contracting Methodology	31
3 IGCC Power Plants.....	32
3.1 IGCC Common Process Areas.....	32
3.1.1 Coal Receiving and Storage.....	32
3.1.2 Air Separation Unit (ASU) Choice and Integration.....	33
3.1.3 Water Gas Shift Reactors.....	37
3.1.4 Mercury Removal	38
3.1.5 Acid Gas Removal (AGR) Process Selection.....	39
3.1.6 Sulfur Recovery/Tail Gas Cleanup Process Selection	42
3.1.7 Slag Handling.....	44
3.1.8 Power Island.....	45
3.1.9 Steam Generation Island.....	48
3.1.10 Accessory Electric Plant	51
3.1.11 Waste Treatment/Miscellaneous Systems.....	52
3.1.12 Instrumentation and Control	52
3.1.13 Performance Summary Metrics	52
3.2 Shell Global Solutions IGCC Cases	54
3.2.1 Gasifier Background	54
3.2.2 Process Description.....	55
3.2.3 Key System Assumptions	61
3.2.4 Sparing Philosophy	63
3.2.5 Case B1A - Performance Results.....	64

3.2.6 Case B1A - Major Equipment List	75
3.2.7 Case B1A - Cost Estimating	80
3.2.8 Case B1B - Shell IGCC Power Plant with CO ₂ Capture	86
3.2.9 Case B1B - Performance Results	91
3.2.10 Case B1B - Major Equipment List	100
3.2.11 Case B1B - Cost Estimating	105
3.3 CB&I E-Gas TM IGCC Cases	111
3.3.1 Gasifier Background	111
3.3.2 Process Description	112
3.3.3 Key System Assumptions	118
3.3.4 Sparing Philosophy	118
3.3.5 Case B4A - Performance Results	119
3.3.6 Case B4A - Major Equipment List	129
3.3.7 Case B4A - Costs Estimating Results	134
3.3.8 Case B4B - E-Gas TM IGCC Power Plant with CO ₂ Capture	140
3.3.9 Case B4B - Performance Results	145
3.3.10 Case B4B - Major Equipment List	154
3.3.11 Case B4B - Cost Estimating Results	159
3.4 General Electric Energy IGCC Cases	165
3.4.1 Gasifier Background	165
3.4.2 Process Description	166
3.4.3 Key System Assumptions	173
3.4.4 Sparing Philosophy	175
3.4.5 Case B5A - Performance Results	175
3.4.6 Case B5A - Major Equipment List	186
3.4.7 Case B5A - Cost Estimating	191
3.4.8 Case B5B - GEE Radiant IGCC with CO ₂ Capture	197
3.4.9 Case B5B - Performance Results	201
3.4.10 Case B5B - Major Equipment List	211
3.4.11 Case B5B - Cost Estimating	216
3.4.12 Case B5B-Q - GEE Quench IGCC with CO ₂ Capture	222
3.4.13 Case B5B-Q - Performance Results	226
3.4.14 Case B5B-Q - Major Equipment List	236
3.4.15 Case B5B-Q - Cost Estimating	241
3.5 IGCC Case Summary	247
3.5.1 Sensitivities	253
4 Revision Control	257
5 References	261

Exhibits

Exhibit ES-1 Case configuration summary.....	1
Exhibit ES-2 Performance summary and environmental profile for all cases.....	4
Exhibit ES-3 Net plant efficiency (HHV basis).....	5
Exhibit ES-4 Cost summary for all cases	6
Exhibit ES-5 COE by cost component.....	8
Exhibit 1-1 Case descriptions	12
Exhibit 2-1 Site characteristics	13
Exhibit 2-2 Site ambient conditions.....	14
Exhibit 2-3 Design coal	15
Exhibit 2-4 Probability distribution of mercury concentration in the Illinois No. 6 coal	16
Exhibit 2-5 MATS and NSPS emission limits for PM, HCl, SO ₂ , NO _x , and Hg.....	16
Exhibit 2-6 Environmental targets	18
Exhibit 2-7 Capital cost levels and their elements.....	23
Exhibit 2-8 CO ₂ transport and storage costs	29
Exhibit 3-1 Air extracted from the combustion turbine and supplied to the ASU in non-carbon capture cases	35
Exhibit 3-2 Typical ASU process schematic	36
Exhibit 3-3 Flow diagram for a conventional AGR unit	40
Exhibit 3-4 Typical three-stage Claus sulfur plant	43
Exhibit 3-5 State-of-the-art 2008 F-class combustion turbine performance characteristics using natural gas (42)	45
Exhibit 3-6 Typical fuel specification for state-of-the-art 2008 F-class machines.....	46
Exhibit 3-7 Allowable gas fuel contaminant level for state-of-the-art 2008 F-class machines....	47
Exhibit 3-8 Case B1A block flow diagram, Shell IGCC without CO ₂ capture	56
Exhibit 3-9 Case B1A stream table, Shell IGCC without capture.....	57
Exhibit 3-10 Shell IGCC plant study configuration matrix	62
Exhibit 3-11 Balance of plant assumptions	63
Exhibit 3-12 Case B1A plant performance summary	65
Exhibit 3-13 Case B1A plant power summary	66
Exhibit 3-14 Case B1A air emissions	67
Exhibit 3-15 Case B1A carbon balance	68
Exhibit 3-16 Case B1A sulfur balance.....	68
Exhibit 3-17 Case B1A water balance	69
Exhibit 3-18 Case B1A coal gasification and ASU heat and mass balance schematic	71
Exhibit 3-19 Case B1A syngas cleanup heat and mass balance schematic	72
Exhibit 3-20 Case B1A combined cycle power generation heat and mass balance schematic	73
Exhibit 3-21 Case B1A overall energy balance (0°C [32°F] reference).....	75
Exhibit 3-22 Case B1A total plant cost details	81
Exhibit 3-23 Case B1A owner's costs	84
Exhibit 3-24 Case B1A initial and annual operating and maintenance costs	85
Exhibit 3-25 Case B1A COE breakdown	85
Exhibit 3-26 Case B1B block flow diagram, Shell IGCC with CO ₂ capture	88
Exhibit 3-27 Case B1B stream table, Shell IGCC with capture	89
Exhibit 3-28 Case B1B plant performance summary	91
Exhibit 3-29 Case B1B plant power summary	92

Exhibit 3-30 Case B1B air emissions	93
Exhibit 3-31 Case B1B carbon balance	94
Exhibit 3-32 Case B1B sulfur balance.....	94
Exhibit 3-33 Case B1B water balance	95
Exhibit 3-34 Case B1B coal gasification and ASU heat and mass balance schematic.....	96
Exhibit 3-35 Case B1B syngas cleanup heat and mass balance schematic	97
Exhibit 3-36 Case B1B combined cycle power generation heat and mass balance schematic....	98
Exhibit 3-37 Case B1B overall energy balance (0°C [32°F] reference).....	100
Exhibit 3-38 Case B1B total plant cost details	106
Exhibit 3-39 Case B1B owner's costs	109
Exhibit 3-40 Case B1B initial and annual operating and maintenance costs	110
Exhibit 3-41 Case B1B COE breakdown.....	110
Exhibit 3-42 Case B4A block flow diagram, E-Gas™ IGCC without CO ₂ capture	113
Exhibit 3-43 Case B4A stream table, E-Gas™ IGCC without capture	114
Exhibit 3-44 E-Gas™ IGCC plant study configuration matrix	119
Exhibit 3-45 Case B4A plant performance summary	120
Exhibit 3-46 Case B4A plant power summary	121
Exhibit 3-47 Case B4A air emissions	122
Exhibit 3-48 Case B4A carbon balance	123
Exhibit 3-49 Case B4A sulfur balance.....	123
Exhibit 3-50 Case B4A water balance	124
Exhibit 3-51 Case B4A coal gasification and ASU heat and mass balance schematic	125
Exhibit 3-52 Case B4A syngas cleanup heat and mass balance schematic	126
Exhibit 3-53 Case B4A combined cycle power generation heat and mass balance schematic ..	127
Exhibit 3-54 Case B4A overall energy balance (0°C [32°F] reference).....	129
Exhibit 3-55 Case B4A total plant cost details	135
Exhibit 3-56 Case B4A owner's costs	138
Exhibit 3-57 Case B4A initial and annual operating and maintenance costs	139
Exhibit 3-58 Case B4A COE breakdown	139
Exhibit 3-59 Case B4B block flow diagram, E-Gas™ IGCC with CO ₂ capture.....	142
Exhibit 3-60 Case B4B stream table, E-Gas™ IGCC with capture.....	143
Exhibit 3-61 Case B4B plant performance summary	145
Exhibit 3-62 Case B4B plant power summary	146
Exhibit 3-63 Case B4B air emissions	147
Exhibit 3-64 Case B4B carbon balance	148
Exhibit 3-65 Case B4B sulfur balance.....	148
Exhibit 3-66 Case B4B water balance	149
Exhibit 3-67 Case B4B coal gasification and ASU heat and mass balance schematic.....	150
Exhibit 3-68 Case B4B syngas cleanup heat and mass balance schematic	151
Exhibit 3-69 Case B4B combined cycle power generation heat and mass balance schematic..	152
Exhibit 3-70 Case B4B overall energy balance (0°C [32°F] reference).....	154
Exhibit 3-71 Case B4B total plant cost details	160
Exhibit 3-72 Case B4B owner's costs	163
Exhibit 3-73 Case B4B initial and annual operating and maintenance costs	164
Exhibit 3-74 Case B4B COE breakdown.....	164
Exhibit 3-75 Case B5A block flow diagram, GEE IGCC without CO ₂ capture.....	169

Exhibit 3-76 Case B5A stream table, GEE IGCC without capture	170
Exhibit 3-77 GEE IGCC plant study configuration matrix	174
Exhibit 3-78 Case B5A plant performance summary	176
Exhibit 3-79 Case B5A plant power summary	177
Exhibit 3-80 Case B5A air emissions	178
Exhibit 3-81 Case B5A carbon balance	179
Exhibit 3-82 Case B5A sulfur balance.....	179
Exhibit 3-83 Case B5A water balance	180
Exhibit 3-84 Case B5A coal gasification and ASU heat and mass balance schematic	182
Exhibit 3-85 Case B5A syngas cleanup heat and mass balance schematic	183
Exhibit 3-86 Case B5A combined-cycle power generation heat and mass balance schematic ..	184
Exhibit 3-87 Case B5A overall energy balance (0°C [32°F] reference).....	186
Exhibit 3-88 Case B5A total plant cost details	192
Exhibit 3-89 Case B5A owner's costs	195
Exhibit 3-90 Case B5A initial and annual operating and maintenance costs	196
Exhibit 3-91 Case B5A COE breakdown	196
Exhibit 3-92 Case B5B block flow diagram, GEE IGCC with CO ₂ capture.....	198
Exhibit 3-93 Case B5B stream table, GEE IGCC with capture.....	199
Exhibit 3-94 Case B5B plant performance summary	202
Exhibit 3-95 Case B5B plant power summary	203
Exhibit 3-96 Case B5B air emissions	204
Exhibit 3-97 Case B5B carbon balance	205
Exhibit 3-98 Case B5B sulfur balance.....	205
Exhibit 3-99 Case B5B water balance	206
Exhibit 3-100 Case B5B coal gasification and ASU heat and mass balance schematic.....	207
Exhibit 3-101 Case B5B syngas cleanup heat and mass balance schematic	208
Exhibit 3-102 Case B5B combined-cycle power generation heat and mass balance schematic	209
Exhibit 3-103 Case B5B overall energy balance (0°C [32°F] reference).....	211
Exhibit 3-104 Case B5B total plant cost details	217
Exhibit 3-105 Case B5B owner's costs	220
Exhibit 3-106 Case B5B initial and annual operating and maintenance costs	221
Exhibit 3-107 Case B5B COE breakdown.....	221
Exhibit 3-108 Case B5B-Q block flow diagram, GEE quench-only IGCC with CO ₂ capture..	223
Exhibit 3-109 Case B5B-Q stream table, GEE quench-only IGCC with capture	224
Exhibit 3-110 Case B5B-Q plant performance summary	227
Exhibit 3-111 Case B5B-Q plant power summary	228
Exhibit 3-112 Case B5B-Q air emissions	229
Exhibit 3-113 Case B5B-Q carbon balance	230
Exhibit 3-114 Case B5B-Q sulfur balance.....	230
Exhibit 3-115 Case B5B-Q water balance	231
Exhibit 3-116 Case B5B-Q coal gasification and ASU heat and mass balance schematic	232
Exhibit 3-117 Case B5B-Q syngas cleanup heat and mass balance schematic	233
Exhibit 3-118 Case B5B-Q combined cycle power generation heat and mass balance schematic	234
Exhibit 3-119 Case B5B-Q overall energy balance.....	236
Exhibit 3-120 Case B5B-Q total plant cost details	242

Exhibit 3-121 Case B5B-Q owner's costs	245
Exhibit 3-122 Case B5B-Q initial and annual operating and maintenance costs	246
Exhibit 3-123 Case B5B-Q COE breakdown	246
Exhibit 3-124 Estimated performance and cost results for all cases	247
Exhibit 3-125 Plant capital cost for all cases	249
Exhibit 3-126 COE for all cases	250
Exhibit 3-127 Cost of CO ₂ captured and avoided in all cases	251
Exhibit 3-128 Raw water withdrawal and consumption in all cases	252
Exhibit 3-129 Capacity factor sensitivity of all cases.....	253
Exhibit 3-130 Sensitivity of COE to coal price for all cases	254
Exhibit 3-131 Impact of carbon emissions price on study technologies	255
Exhibit 3-132 Impact of carbon sales price on study technologies	256
Exhibit 4-1 Record of revisions	257

Acronyms and Abbreviations

AACE	Association for the Advancement of Cost Engineering	CO ₂	Carbon dioxide
acfm	Actual cubic feet per minute	CO ₂ E	Carbon dioxide equivalent
ACI	Activated carbon injection	COE	Cost of electricity
ADIP	Aqueous di-isoproponal	COH	Cost of hydrogen
AEO	Annual Energy Outlook	COR	Contracting Officer's Representative
AF	Available factor	COS	Carbonyl sulfide
AGR	Acid gas removal	CRT	Cathode ray tube
ANSI	American National Standards Institute	CS	Carbon steel
Ar	Argon	CT	Combustion turbine
AR	As received	CTG	Combustion turbine-generator
Aspen	Aspen Plus®	CWP	Circulating water pump
ASU	Air separation unit	CWR	Cooling water return
atm	Atmosphere (14.696 psi)	CWS	Circulating water system
BACT	Best available control technology	CWT	Cold water temperature
BAH	Booz Allen Hamilton	dB	Decibel
BDL	Blowdown losses	DB	Dry basis
BEC	Bare erected cost	DCA	Direct contact absorber
BFD	Block flow diagram	DCS	Distributed control system
BFW	Boiler feedwater	DEA	Diethanolamine
Btu	British thermal unit	DI	De-ionized
Btu/hr	British thermal units per hour	Dia.	Diameter
Btu/kWh	British thermal units per kilowatt hour	DIPA	Diisopropanolamine
Btu/lb	British thermal units per pound	DLN	Dry low NO _x
Btu/scf	British thermal units per standard cubic foot	DME	Di-methyl ether
CAAA	Clean Air Act Amendments of 1990	DOE	Department of Energy
CAIR	Clean Air Interstate Rule	DSCR	Debt service coverage ratio
CAMR	Clean Air Mercury Rule	DSI	Dry sorbent injection
CB&I	Chicago Bridge & Iron Company	DSRP	Direct sulfur recovery process
CCF	Capital charge factor	eGRID	Emissions & Generation Resource Integrated Database
CCS	Carbon capture and sequestration	E-Gas™	CB&I gasifier technology
CCS	Carbon capture and storage	EAF	Equivalent availability factor
CDR	Carbon dioxide recovery	EGU	Electric utility steam generating units
cf	Cubic feet	EIA	Energy Information Administration
CF	Capacity factor	EM	Electromagnetic
cfm	Cubic feet per minute	EMF	Emission modification factors
CFR	Code of Federal Regulations	EOR	Enhanced oil recovery
CGE	Cold gas efficiency	EPA	Environmental Protection Agency
CGCU	Cold gas cleanup	EPC	Engineer/Procure/Construct
CH ₄	Methane	EPCC	Engineering, Procurement and Construction Cost
CL	Closed-loop	EPRI	Electric Power Research Institute
cm	Centimeter	EPCM	Engineering/Procurement/Construction Management
CO	Carbon monoxide	ERC	Emission reduction credits

ESP	Electrostatic precipitator	HP	High pressure
ESPA	Energy Sector Planning and Analysis	HRSG	Heat recovery steam generator
ETE	Effective thermal efficiency	HSS	Heat stable salts
FD	Forced draft	HV	High voltage
FE	Fossil energy	HVAC	Heating, ventilating, and air conditioning
FERC	Federal Energy Regulatory Commission	HWT	Hot water temperature
FG	Flue gas	Hz	Hertz
FGD	Flue gas desulfurization	ICR	Information Collection Request
FO&M	Fixed operations and maintenance	ID	Induced draft
FOAK	First-of-a-kind	IEA	International Energy Agency
FRP	Fiberglass-reinforced plastic	IEEE	Institute of Electrical and Electronics Engineers
FSQ	Full-slurry quench	IGCC	Integrated gasification combined cycle
ft	Foot, Feet	IGV	Inlet guide vane
ft, w.g.	Feet of water gauge	in. H ₂ O	Inch water
FW	Feedwater	in. Hg	Inch mercury (absolute pressure)
GADS	Generating Availability Data System	in. W.C.	Inch water column
gal	Gallon	IOU	Investor-owned utility
gal/MWh	Gallons per megawatt hour	IP	Intermediate pressure
GCV	Gross calorific value	IPCC	Intergovernmental Panel on Climate Change
GDP	Gross domestic product	IPM	Integrated Planning Model
GEE	General Electric Energy	IPP	Independent power producer
GJ	Gigajoule	IRR	Internal rate of return
GJ/hr	Gigajoules per hour	IRROE	Internal rate of return on equity
GHG	Greenhouse gas	ISBL	Inside battery limits
GHGEV	Greenhouse gas emission value	ISO	International Organization for Standardization
gpd	Gallons per day	kg/GJ	Kilograms per gigajoule
gpm	Gallons per minute	kg/hr	Kilograms per hour
gr/100 scf	grains per one hundred standard cubic feet	kg/s	Kilograms per second
GRI	Gas Research Institute	kgmol	Kilogram mole
GSU	Generator step-up transformers	kgmol/hr	Kilogram moles per hour
Gt	Gigatonnes	kJ	Kilojoule
GWh	Gigawatt-hour	kJ/hr	Kilojoules per hour
h, hr	Hour	kJ/kg	Kilojoules per kilogram
H ₂	Hydrogen	km	Kilometer
H ₂ A	DOE's Hydrogen Advisory Group	KO	Knockout
H ₂ O	Water	kPa	Kilopascal
H ₂ S	Hydrogen sulfide	kV	Kilovolt
H ₂ SO ₄	Sulfuric acid	kW, kWe	Kilowatt electric
HAP	Hazardous air pollutant	kWh	Kilowatt-hour
HCl	Hydrochloric acid	kWt	Kilowatt thermal
HCPV	Hydrocarbon pore volume	LAER	Lowest Achievable Emission Rate
HDPE	High density polyethylene	lb	Pound
Hg	Mercury	lb/gal	Pound per gallon
HHV	Higher heating value		
Horiz.	Horizontal		
hp	Horsepower		

lb/hr	Pounds per hour	MMscfy	Million standard cubic feet per year
lb/ft ²	Pounds per square foot	mol%	Mole percent (percent by mole)
lb/MMBtu	Pounds per million British thermal units	MNQC	Multi Nozzle Quiet Combustor
lbmol	Pound mole	MPa	Megapascal
lbmol/hr	Pound moles per hour	mpg	Miles per gallon
lb/MWh	Pounds per megawatt hour	MTG	Methanol to gasoline
lb/TBtu	Pounds per trillion British thermal units	MVA	Mega volt-amps
LC	Life cycle	MW	Megawatt
LCA	Life cycle analysis	MWe	Megawatt electric
LCC	Life cycle cost	MWh	Megawatt-hour
LCI	Life cycle inventory	MWt	Megawatt thermal
LCIA	Life cycle impact assessment	N ₂	Nitrogen
LGTI	Louisiana Gasification Technology, Inc.	N ₂ O	Nitrous oxide
LHV	Lower heating value	N/A	Not applicable
LIBOR	London Interbank Offered Rate	NAAQS	National Ambient Air Quality Standards
LMP	Locational Marginal Price	NaOH	Sodium hydroxide
LNG	Liquified natural gas	NCF	Net capacity factor
LNB	Low NO _x burner	Neg.	Negligible
LP	Low pressure	NEMA	National Electrical Manufacturers Association
LPG	Liquefied petroleum gas	NEMS	National Energy Modeling System
lpm	Liters per minute	NERC	North American Electric Reliability Council
LSE	Load-serving entities	NETL	National Energy Technology Laboratory
LV	Low voltage	NFPA	National Fire Protection Association
M	Thousand	NG	Natural gas
m	Meter	NGCC	Natural gas combined cycle
m/min	Meters per minute	NGL	Natural gas liquids
m ³ /min	Cubic meters per minute	NH ₃	Ammonia
MAC	Main air compressor	Nm ³	Normal cubic meter
MAF	Moisture and ash free	Nm ³ /hr	Normal cubic meter per hour
MATS	Mercury and Air Toxics Standards	NOAK	N th -of-a-kind
Mbbl	Thousand barrels	NO _x	Oxides of nitrogen
Mcf	Thousand cubic feet	NSPS	New Source Performance Standards
MCR	Maximum continuous rate	NSR	New Source Review
md	Millidarcy (a measure of permeability)	O ₂	Oxygen
MDEA	Methyldiethanolamine	O&GJ	Oil & Gas Journal
MeOH	Methanol	O&M	Operation and maintenance
MHz	Megahertz	OC _{Fn}	Category n fixed operating cost for the initial year of operation
mi	Mile	OC _{Vnq}	Category n variable operating cost for the initial year of operation
MJ/Nm ³	Megajoules per normal cubic meter	OD	Outside diameter
MM	Million	OEM	Original equipment manufacturers
MMBtu	Million British thermal units	OFA	Overfire air
MMBtu/hr	Million British thermal units per hour	OPPA	Office of Program Planning & Analysis
MMscf	Million standard cubic feet		
MMscfd	Million standard cubic feet per day		

OP/VWO	Over pressure/valve wide open	scfd	Standard cubic feet per day
OSHA	Occupational Safety and Health Administration	scfh	Standard cubic feet per hour
		scfm	Standard cubic feet per minute
PA	Primary air	Sch.	Schedule
PAC	Powdered activated carbon	scmh	Standard cubic meter per hour
PC	Pulverized coal	SCOT	Shell Claus Off-gas Treating
PAS	PAS, Inc.	SCR	Selective catalytic reduction process or equipment
PC	Pulverized coal		
p.f.	Power factor	SDA	Spray dryer absorber
PFD	Process flow diagram	SEAP	Office of Strategic Energy Analysis & Planning
PM	Particulate matter		
PM ₁₀	Particulate matter measuring 10 µm (micrometers) or less	SFC	Synthetic Fuels Corporation
		SG	Specific gravity
PO	Purchase order	SGC	Synthesis gas cooler
POTW	Publicly owned treatment works	SGS	Sour gas shift
ppbw	Parts per billion weight	Shell	Shell Global Solutions
ppm	Parts per million	SO ₂	Sulfur dioxide
ppmv	Parts per million volume	SO _x	Oxides of sulfur
ppmvd	Parts per million volume, dry	SNCR	Selective non-catalytic reduction process or equipment
ppmw	Parts per million weight		
ppmwd	Parts per million weight, dry	SNG	Synthetic natural gas
PRB	Powder River Basin coal region	SRU	Sulfur recovery unit
PSA	Pressure swing adsorption	SS	Stainless steel
PSD	Prevention of significant deterioration	SS Amine	SS Specialty Amine
psi	Pounds per square inch	st	Short ton
psia	Pound per square inch absolute	STG	Steam turbine generator
psid	Pound per square inch differential	SWS	Sour water scrubber
psig	Pound per square inch gage	Syngas	Synthetic gas
PSFM	Power systems financial model	T&D	Transmission and distribution
PTFE	Teflon (Polytetrafluoroethylene)	T&S	Transport and storage
PV	Present value	TASC	Total as-spent cost
QGESS	Quality Guidelines for Energy System Studies	TCR	Total capital requirement
		TEWAC	Totally Enclosed Water-to-Air Cooled
Qty	Quantity		
R&D	Research and development	TGTU	Tail gas treating unit
RCRA	Resource Conservation and Recovery Act	tonne	Metric ton (1,000 kg)
		TOC	Total overnight cost
RD&D	Research, development, and demonstration	TPC	Total plant cost
		tpd	Ton per day
RDS	Research and Development Solutions, LLC	tph	Tons per hour
		TPI	Total plant investment
RH	Reheater	U.S.	United States
RSC	Radiant syngas cooler	USC	Ultra-supercritical
RSP	Required selling price	V	Volt
RTO	Regional transmission operations/operator	VOC	Volatile organic compound
		VO&M	Variable operations and maintenance
SC	Supercritical	V-L	Vapor liquid portion of stream (excluding solids)
SC PC	Supercritical Pulverized Coal		
scf	Standard cubic feet		

vol%	Volume percent	\$/kW	Dollars per kilowatt
WB	Wet bulb	\$/MMBtu	Dollars per million British thermal units
wg	Water gauge		
WGPU	Warm gas cleanup	\$M	Millions of dollars
WGS	Water gas shift	μS/cm	micro Siemens per cm
wt%	Weight percent	°C	Degrees Celsius
yr	Year	°F	Degrees Fahrenheit
ZnO	Zinc oxide	5-10s	Fifty hour work weeks
ZnS	Zinc sulfide		
\$/GJ	Dollars per gigajoule		

This page intentionally left blank.

Executive Summary

This report presents the cost and performance results of an assessment of seven integrated gasification combined cycle (IGCC) plants. The primary value of this report lies not in the absolute accuracy of the capital cost estimates for the individual cases (estimated to be -15 percent/+30), but in the application of a consistent approach to allow meaningful comparisons of relative costs among the cases evaluated.

This report is part of an update to Volume 1 of a four-volume series, which consists of:

- Volume 1: Bituminous Coal and Natural Gas to Electricity
- Volume 2: Coal to Synthetic Natural Gas and Ammonia (Various Coal Ranks)
- Volume 3: Low Rank Coal and Natural Gas to Electricity
- Volume 4: Bituminous Coal to Liquid Fuels

This report, Volume 1b, retains the IGCC design and performance updates from the November 2010 release of Volume 1 (Revision 2)¹, but updates the IGCC costs for all cases to 2011 year dollars. Volume 1a covers the cost and performance update of all pulverized coal (PC) and natural gas combined cycle (NGCC) cases. Section 4 has a revision control table listing the updates applied to this report.

Seven power plant configurations were analyzed in this report, including two Shell Global Solutions (Shell) gasifiers (with and without carbon dioxide [CO₂] capture), two Chicago Bridge and Iron (CB&I) E-GasTM full-slurry quench (FSQ) gasifiers (with and without CO₂ capture), and three General Electric Energy (GEE) gasifiers (one without and two with CO₂ capture [one radiant and one quench]). A summary of the case configurations in the report is shown in Exhibit ES-1.

Exhibit ES-1 Case configuration summary

Case	Steam Cycle, psig/°F/°F	Combustion Turbine	Gasifier Technology	H ₂ S Separation	PM Control	CO ₂ Separation ^B
B1A	1800/1050/1050	2 x State-of-the-art 2008 F-Class	Shell	Sulfinol-M	Cyclone, candle filter, and water scrubber	N/A
B1B	1800/1000/1000	2 x State-of-the-art 2008 F-Class	Shell	Selexol	Cyclone, candle filter, and water scrubber	Selexol 2 nd stage
B4A	1800/1050/1050	2 x State-of-the-art 2008 F-Class	CB&I E-Gas TM	Refrigerated MDEA	Cyclone, candle filter, and water scrubber	N/A
B4B	1800/1000/1000	2 x State-of-the-art 2008 F-Class	CB&I E-Gas TM	Selexol	Cyclone, candle filter, and water scrubber	Selexol 2 nd stage
B5A	1800/1050/1050	2 x State-of-the-art 2008 F-Class	GEE Radiant	Selexol	Quench, water scrubber, and AGR adsorber	N/A
B5B	1800/1000/1000	2 x State-of-the-art 2008 F-Class	GEE Radiant	Selexol	Quench, water scrubber, and AGR adsorber	Selexol 2 nd stage
B5B-Q	1800/1000/1000	2 x State-of-the-art 2008 F-Class	GEE Quench	Selexol	Quench, water scrubber, and AGR adsorber	Selexol 2 nd stage

¹ In order to meet the new mercury limits, discussed in Section 2.3, the design of the carbon bed was updated to be a dual sequential carbon bed.

All plant configurations were evaluated based on installation at a greenfield site. Capacity factors (CF) are assumed to equal approximate availability at 80 percent for all configurations. Achieving such capacity factors would require that these plants be near the top of the dispatch list.

The combustion turbines (CT) used in each of the configurations are manufactured in discrete sizes; each of the configurations utilizes an identical CT design with the output maximized. While the output of the combustion turbines is consistent between configurations, the cases have net outputs ranging from 494 (Case B5B-Q) to 629 MW (Case B1A). The range in net output is caused by the significant auxiliary load imposed in the CO₂ capture cases – primarily due to CO₂ compression – and the need for extraction steam for CO₂ capture and the sour gas shift (SGS) reactions, which reduces steam turbine output.

Environmental emission requirements are based on the mercury (Hg) and hydrochloric acid (HCl) limits set by the March 2013 update to the Utility Mercury and Air Toxics Standards (MATS), and particulate matter (PM), sulfur dioxide (SO₂), and nitrogen oxide (NO_x) limits set by the February 2013 update to the New Source Performance Standards (NSPS).

Mercury is controlled with dual sulfur-impregnated carbon beds; H₂S is separated with an acid gas removal (AGR) process and controlled with a Claus plant to limit SO₂ stack emissions; NO_x formation is minimized with low NO_x burners (LNB) and nitrogen dilution; the particulate matter is controlled with a syngas scrubber, cyclone, and candle filter; the HCl is primarily removed in the syngas scrubber with the remainder removed with the low temperature heat recovery condensate, and the combined effluent is treated with calcium hydroxide.

All of the power plant configurations with carbon capture in this report are designed to achieve 90 percent capture, resulting in atmospheric CO₂ emissions at levels far below proposed Environmental Protection Agency (EPA) regulation.²

The performance results presented in this report are unchanged from Revision 2. The methodology used in their development included performing steady-state simulations of the seven power plant configurations using the Aspen Plus® (Aspen) modeling program. The major plant equipment performance and process limits were based on published reports, information obtained from vendors and users of the technology, performance data from design/build utility projects, and/or best engineering judgment.

In Revision 2, mass and energy balance data from the Aspen models were used to size major pieces of equipment, which formed the basis for developing the cost estimates presented in that report. The capital and operating costs for the major equipment and plant sub-systems had been estimated by WorleyParsons based on the simulation results using an in-house database and conceptual estimating models. The cost results were further calibrated using a combination of adjusted vendor-furnished data and scaled estimates from previous design/build projects. All Revision 2 costs were reported in 2007 dollars.

It was the purpose of this report (Revision 2b) to update the costs to 2011 dollars. This effort was accomplished through scaled escalation and corrections. The only exception was to address

² The Environmental Protection Agency (EPA) proposed a new source performance standard on April 13, 2012, for emissions of carbon dioxide for new fossil fuel-fired electric utility generating units. As of the publication of this report, the proposed regulation has been published in the Federal Register. (54) The limit set by the proposed regulation is 1,100 lb-CO₂/MWh-gross.

the use of a dual carbon system for Hg removal, which was re-costed from a single carbon system to address environmental targets discussed in Section 2.3.

The baseline fuel cost for this analysis is specified in the November 2012 revision of the Quality Guidelines for Energy System Studies (QGESS) document on “Fuel Prices for Selected Feedstocks in NETL Studies.” (1) The first year of capital expenditure price used is \$2.78/GJ (\$2.94/MMBtu) for Illinois No. 6 coal delivered to the Midwest, on a higher heating value (HHV) basis and in 2011 United States (U.S.) dollars.

The cost metric used in this report is the cost of electricity (COE), which is the revenue that must be received by the generator per net megawatt-hour produced to meet the desired internal rate of return on equity. The COE is assumed to escalate at a nominal annual rate equal to the general inflation rate, i.e., it remains constant or levelized in real terms over the operational period of the power plant. The cost of CO₂ transport and storage (T&S) of \$11 per tonne of CO₂ is added to the COE and represents a 62 km (100 mile) CO₂ pipeline and storage in a deep saline formation in the Midwest.³

The cost and performance of the various fossil fuel-based technologies will be important in determining which combination of technologies will be utilized to meet the demands of the power market in the future.

Selection of new generation technologies will depend on many factors, including:

- Capital and operating costs
- Overall energy efficiency
- Fuel prices
- COE
- Availability, reliability, and environmental performance
- Current and potential regulations governing air, water, and solid waste discharges from fossil-fueled power plants
- Market penetration of clean coal technologies that have matured and improved as a result of recent commercial-scale demonstrations under the Department of Energy’s (DOE) Clean Coal and Carbon Management Program

Results Analysis

Exhibit ES-2 shows the performance and environmental profile summary for all cases. A graph of the net plant efficiency (HHV basis) is provided in Exhibit ES-3.

³ Estimated using the FE/NETL CO₂ Transport Cost Model and the FE/NETL CO₂ Saline Storage Cost Model. Additional detail on development of these costs is available in the May 2014 revision of the QGESS document “Carbon Dioxide Transport and Storage Costs in NETL Studies.” (26)

Exhibit ES-2 Performance summary and environmental profile for all cases

	Integrated Gasification Combined Cycle						
	Shell		E-Gas™ FSQ		GEE R+Q		
Case Name (Old Case Name) ^A	B1A (5)	B1B (6)	B4A (3)	B4B (4)	B5A (1)	B5B (2)	B5B-Q (2a)
PERFORMANCE							
Gross Power Output (MWe)	737	673	738	704	748	734	684
Auxiliary Power Requirement (MWe)	108	177	113	190	126	191	190
Net Power Output (MWe)	629	497	625	513	622	543	494
Coal Flow rate (lb/hr)	436,646	465,264	459,956	484,212	466,898	487,005	486,978
Natural Gas Flow rate (lb/hr)	N/A	N/A	N/A	N/A	N/A	N/A	N/A
HHV Thermal Input (kW _t)	1,492,878	1,590,722	1,572,575	1,655,503	1,596,309	1,665,056	1,664,962
Net Plant HHV Efficiency (%)	42.1%	31.2%	39.7%	31.0%	39.0%	32.6%	29.7%
Net Plant HHV Heat Rate (Btu/kWh)	8,100	10,927	8,585	11,002	8,758	10,459	11,505
Raw Water Withdrawal, gpm	4,150	5,652	4,382	5,757	4,755	5,834	6,137
Process Water Discharge, gpm	788	1,021	905	1,095	984	1,080	1,129
Raw Water Consumption, gpm	3,362	4,631	3,477	4,662	3,771	4,754	5,008
CO ₂ Capture Rate (%)	0	90	0	90	0	90	90
CO ₂ Emissions (lb/MMBtu)	197	20	199	20	197	20	20
CO ₂ Emissions (lb/MWh-gross)	1,361	161	1,448	158	1,434	152	165
CO ₂ Emissions (lb/MWh-net)	1,595	218	1,711	217	1,724	206	228
SO ₂ Emissions (lb/MMBtu)	0.004	0.002	0.012	0.002	0.001	0.002	0.002
SO ₂ Emissions (lb/MWh-gross)	0.029	0.017	0.085	0.017	0.009	0.017	0.018
NO _x Emissions (lb/MMBtu)	0.059	0.049	0.060	0.049	0.059	0.049	0.049
NO _x Emissions (lb/MWh-gross)	0.409	0.396	0.434	0.396	0.430	0.376	0.404
PM Emissions (lb/MMBtu)	0.007	0.007	0.007	0.007	0.007	0.007	0.007
PM Emissions (lb/MWh-gross)	0.049	0.057	0.052	0.057	0.052	0.055	0.059
Hg Emissions (lb/TBtu)	0.434	0.372	0.413	0.374	0.412	0.388	0.361
Hg Emissions (lb/MWh-gross)	3.00E-06	3.00E-06	3.00E-06	3.00E-06	3.00E-06	3.00E-06	3.00E-06

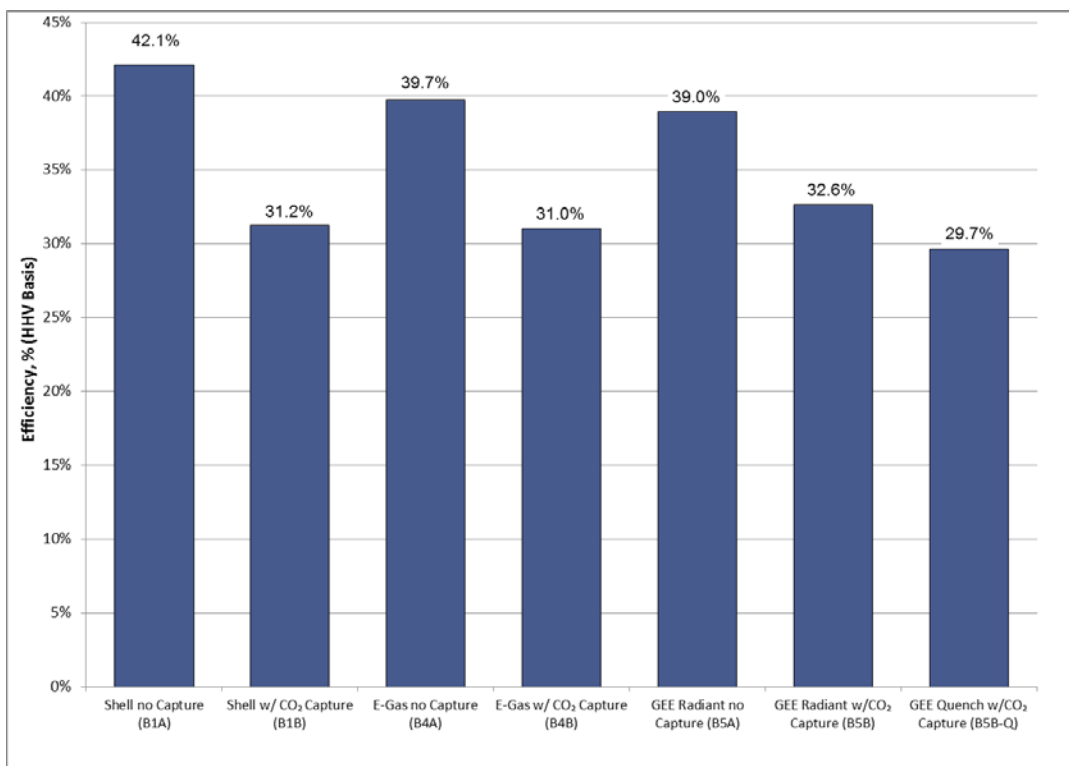
^A Previous versions of this report used a different naming convention. The old case numbers are provided here, paired with the new case numbers for reference.

The primary conclusions that can be drawn are:

- In the non-carbon capture cases the Shell gasifier has the highest net plant efficiency (42.1 percent), followed by the two-stage E-Gas™ slurry fed gasifier (39.7 percent).
- The energy penalty associated with adding CO₂ capture is due to steam extraction for use in the SGS reaction, the auxiliary load for the CO₂ separation and compression equipment, and a slight derate of the gas turbine due to the higher moisture content of the flue gas. The reduction in net plant efficiency ranges from 6 to 11 percentage points with the variability being due to the optimum gasifier designs (e.g., slurry vs. dry feed, syngas quench vs. syngas heat recovery). The capture and non-capture plant configurations may have different optimum gasifier designs.
- The lowest energy penalty (6 percentage points) corresponds to the GEE Radiant gasifier cases primarily due to the non-capture plant design (slurry feed, water quench), which results in a high moisture content in the syngas and thus a low addition of shift steam for SGS for the capture plant design.

- The highest energy penalty (11 percentage points) corresponds to the Shell gasifier cases. The design uses a dry feed system and, in the non-capture configuration, has relatively high heat recovery in the syngas cooler with no water quench, resulting in very low moisture content in the syngas. For the capture configuration, a water quench is added, which increases the moisture content of the syngas for the SGS reaction but decreases the heat recovery in the syngas cooler.
- CB&I E-GasTM has the highest SO₂ emissions (0.085 lb/MWh-gross) of the seven cases because refrigerated MDEA has the lowest H₂S removal efficiency of the AGR technologies.
- Emissions of Hg, PM, NO_x, HCl, and SO₂ are all below the regulatory limits currently in effect and applicable to the technologies within this report.

Exhibit ES-3 Net plant efficiency (HHV basis)



Source: NETL

The cost results for all cases are provided in Exhibit ES-4. A graph of the COE is provided in Exhibit ES-5.

Exhibit ES-4 Cost summary for all cases

	Integrated Gasification Combined Cycle ^A						
	Shell		E-Gas TM FSQ		GEE R+Q		
Case Name	B1A	B1B	B4A	B4B	B5A	B5B	B5B-Q
COST							
Total Plant Cost (2011\$/kW)	2,725	3,981	2,372	3,540	2,449	3,387	3,405
Bare Erected Cost	2,067	2,963	1,809	2,638	1,870	2,525	2,548
Home Office Expenses	207	296	181	264	187	253	255
Project Contingency	376	562	322	492	330	467	467
Process Contingency	75	160	61	147	61	143	135
Total Overnight Cost (2011\$M)	2,107	2,442	1,830	2,252	1,888	2,279	2,086
Total Overnight Cost (2011\$/kW)	3,351	4,917	2,929	4,386	3,036	4,195	4,224
Owner's Costs	626	936	557	846	587	807	819
Total As-Spent Cost (2011\$/kW)	3,820	5,605	3,339	5,000	3,461	4,782	4,815
COE (\$/MWh) (excluding T&S)	107.0	152.6	99.8	141.9	102.6	135.4	138.7
Capital Costs	59.3	87.0	51.8	77.6	53.7	74.2	74.7
Fixed Costs	14.7	20.5	13.5	19.1	13.7	18.2	18.0
Variable Costs	9.3	13.0	9.2	12.8	9.4	12.2	12.2
Fuel Costs	23.8	32.1	25.2	32.3	25.7	30.7	33.8
COE (\$/MWh) (including T&S)	107.0	162.4	99.8	151.8	102.6	144.7	148.9
CO ₂ T&S Costs	0.0	9.8	0.0	9.9	0.0	9.2	10.2
CO ₂ Captured Cost (excluding T&S), \$/tonne ^B	N/A	78.9	N/A	66.5	N/A	63.2	61.0
CO ₂ Avoided Cost (including T&S), \$/tonne ^B	N/A	118.7	N/A	102.9	N/A	91.7	99.3

^AAll cases use high-risk financial assumptions consistent with NETL's "QGESS: Cost Estimation Methodology for NETL Assessments of Power Plant Performance." (2)

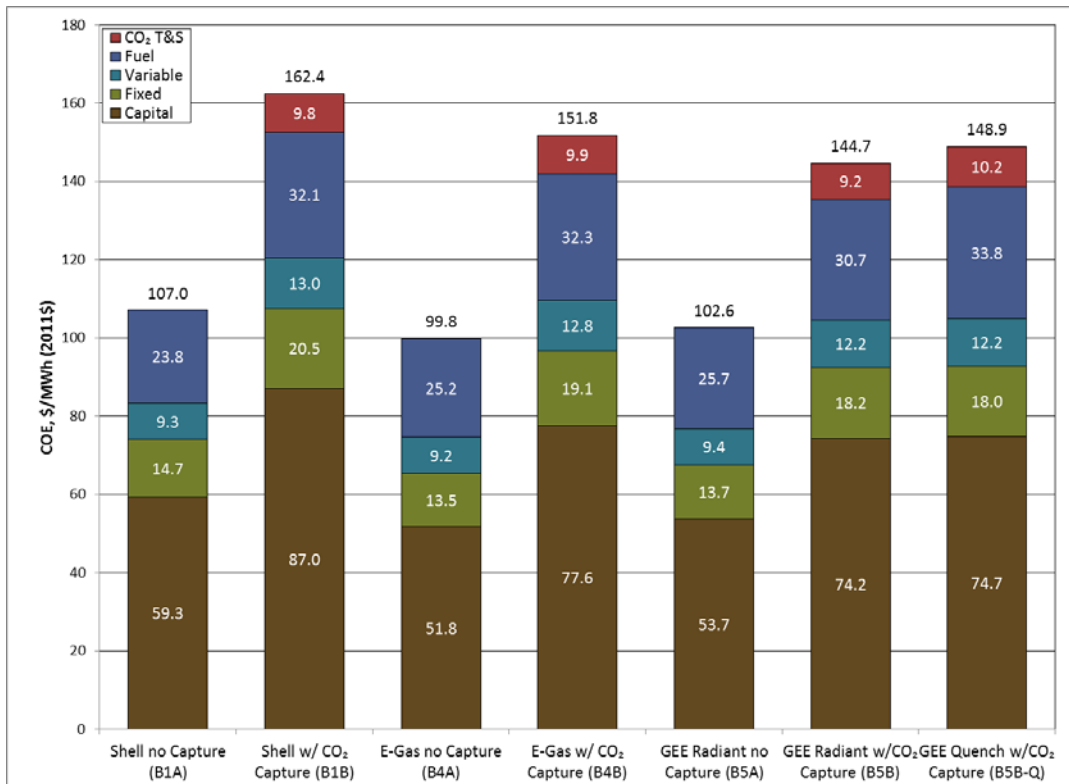
^BBoth the costs of CO₂ captured and avoided were calculated based on the non-capture supercritical pulverized coal (Case B12A) data provided in the "Cost and Performance Baseline for Fossil Energy Plants, Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity, Revision 3" report. (3) Case B12A has a COE of \$82.3/MWh, a CO₂ emission rate of 1,618 lb/MWh-gross, a gross plant output of 579,700 kW, and a net plant output of 550,012 kW.

The primary conclusions that can be drawn are:

- E-GasTM has the lowest total overnight cost (TOC) cost among the non-capture cases. The E-GasTM technology has several features that lend it to being lower cost, such as:
 - The firetube syngas cooler is much smaller and less expensive than a radiant section. E-GasTM can use a firetube boiler because the two-stage design reduces the syngas temperature (slurry quench) into a range where a radiant cooler is not needed.
 - The firetube syngas cooler sits next to the gasifier instead of above or below it, which reduces the height of the main gasifier structure. The E-GasTM proprietary slag removal system – used instead of lock hoppers below the gasifier – also contributes to the lower structure height.
- The normalized TOC of the GEE Radiant gasifier case is approximately 4 percent greater than E-GasTM, and Shell is approximately 14 percent higher.

- The GEE radiant gasifier (GEE quench is less than 1 percent greater than GEE radiant) is the low-cost technology in the CO₂ capture cases, with E-GasTM normalized TOC approximately 5 percent higher and Shell approximately 17 percent higher.
- The ASU cost represents, on average, 12 percent of the TOC (range from 10-13 percent). The ASU cost includes oxygen and nitrogen compression, and in the non-capture cases, it also includes the cost of the CT extraction air heat exchanger. With nitrogen dilution used to the maximum extent possible, nitrogen compression costs are significant.
- The normalized TOC premium for adding CO₂ capture averages 43 percent (\$4,430/kW versus \$3,105/kW).
- The COE is dominated by capital costs and is at least 50 percent of the total in all cases.
- In the non-capture cases the E-GasTM gasifier has the lowest COE, but the differential with Shell is reduced (relative to the normalized TOC comparison) primarily because of the higher efficiency of the Shell gasifier. The Shell COE is 7 percent higher than E-GasTM (compared to 14 percent higher normalized TOC). The GEE gasifier COE is about 3 percent higher than E-GasTM.
- In the capture cases the variation in COE is small; however, the order of the GEE and E-GasTM gasifiers is reversed. The range is from \$144.7/MWh for GEE Radiant to \$162.4/MWh for Shell with GEE Quench and E-GasTM intermediate at \$148.9/MWh and \$151.8/MWh, respectively. The COE CO₂ capture premium for the cases averages 47 percent (range of 41 to 52 percent).
- The CO₂ T&S COE component composes less than 7 percent of the total COE in all capture cases.
- In the event that future legislation assigns a cost to carbon emissions, all of the technologies examined in this report will become more expensive. The technologies without carbon capture will be impacted to a larger extent than those with carbon capture.

Exhibit ES-5 COE by cost component



Source: NETL

Special Considerations on Reported Costs

Capital Costs:

The capital cost estimates documented in this report reflect an uncertainty range of -15 percent/+30 percent, consistent with Association for the Advancement of Cost Engineering (AACE) Class 4 cost estimates (i.e., feasibility study) (4) (5) (6), based on the level of engineering design performed. In all cases, the report intends to represent the next commercial offering and relies on vendor cost estimates for component technologies. It also applies process contingencies at the appropriate subsystem levels in an attempt to account for expected but undefined costs, which can be a challenge for emerging technologies.

Costs of Emerging Technologies and Designs:

The cost estimates for plant designs that include technologies that are not yet fully mature (e.g., IGCC plants and any plant with CO₂ capture) use the same cost estimating methodology as for mature plant designs (e.g., PC and NGCC power plants without CO₂ capture), which does not fully account for the unique cost premiums associated with the initial, complex integrations of emerging technologies in a commercial application. Thus, it is anticipated that initial deployments of IGCC plants – both with and without CO₂ capture – may incur costs higher than those reflected within this report.

Other Factors:

Actual reported project costs for all of the plant types are also expected to deviate from the cost estimates in this report due to project- and site-specific considerations (e.g., contracting strategy, local labor costs, seismic conditions, water quality, financing parameters, local environmental concerns, weather delays, etc.) that may make construction more costly. Such variations are not captured by the reported cost uncertainty.

Future Cost Trends:

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

1 Introduction

The objective of this report is to present an accurate, independent assessment of the cost and performance of fossil energy power systems, specifically integrated gasification combined cycle (IGCC) plants, using a consistent technical and economic approach that reasonably reflects current market conditions.

This report is part of an update to Volume 1 of the four volume series, which consists of:

- Volume 1: Bituminous Coal and Natural Gas to Electricity
- Volume 2: Coal to Synthetic Natural Gas and Ammonia (Various Coal Ranks)
- Volume 3: Low Rank Coal and Natural Gas to Electricity
- Volume 4: Bituminous Coal to Liquid Fuels

This report, Volume 1b, retains the IGCC design and performance updates from the November 2010 release of Volume 1 (Revision 2)⁴, but updates the IGCC costs for all cases to 2011 year dollars. Volume 1a covers the cost and performance update of all pulverized coal (PC) and natural gas combined cycle (NGCC) cases (3). Section 4 has a revision control table listing the updates applied to this report.

Selection of new generation technologies will depend on many factors, including:

- Capital and operating costs
- Overall energy efficiency
- Fuel prices
- Cost of electricity (COE)
- Availability, reliability, and environmental performance
- Current and potential regulation of air, water, and solid waste discharges from fossil-fueled power plants
- Market penetration of clean coal technologies that have matured and improved as a result of recent commercial-scale demonstrations under the Department of Energy's (DOE) Clean Coal and Carbon Management Program

Seven power plant configurations were analyzed, as listed in Exhibit 1-1. The list includes two Shell Global Solutions (Shell) gasifiers (with and without carbon dioxide [CO₂] capture), two CB&I E-GasTM full-slurry quench (FSQ) gasifiers (with and without CO₂ capture), and three General Electric Energy (GEE) gasifiers (one without and two with CO₂ capture [one radiant and one quench]).

While input had previously been sought from various technology vendors during the development of the previous revision (Revision 2) of this report, the final assessment of performance was determined independently and has not been reviewed by individual vendors.

⁴ In order to meet the new mercury limits, discussed in Section 2.3, the design of the carbon bed was updated to be a dual sequential carbon bed.

Generating Unit Configurations

A summary of plant configurations considered in this report is presented in Exhibit 1-1. Components for each plant configuration are described in more detail in the corresponding report sections for each case.

The cases have different gross and net power outputs because of the combustion turbine (CT) size constraint. The state-of-the-art 2008 F-class CT used to model the cases comes in a standard size of 232 MW when operated on syngas at conditions set by the International Standards Organization (ISO). Each case uses two CTs for a combined gross output of 464 MW. In the combined cycle a heat recovery steam generator (HRSG) extracts heat from the CT exhaust to power a steam turbine.

The CO₂ capture cases consume more extraction steam than the non-capture cases, thus reducing the steam turbine output. In addition, the capture cases have a higher auxiliary load requirement than non-capture cases, which serves to further reduce net plant output.

While the two CTs provide 464 MW gross output in all seven cases, the overall combined cycle gross output ranges from 673 to 748 MW, which results in a range of net output from 494 (Case B5B-Q) to 630 MW (Case B1A). The coal feed rate required to achieve the gross power output is also different between the six cases, ranging from 198,220 to 220,899 kg/hr (437,000 to 487,000 lb/hr).

The balance of this report is organized as follows:

- Chapter 2 provides the basis for technical, environmental, and cost evaluations.
- Chapter 3 describes the IGCC technologies modeled and presents the results for the seven cases.
- Chapter 4 includes a record of report revisions.

Exhibit 1-1 Case descriptions

Case (Old Case Name ^A)	Unit Cycle	Steam Cycle, psig/°F/°F	Combustion Turbine	Gasifier Technology	Oxidant	H ₂ S Separation	Sulfur Removal	PM Control	NO _x Control	CO ₂ Separation ^B
B1A (5)	IGCC	1800/1050/1050	2 x State-of-the-art 2008 F-Class	Shell	95 mol% O ₂	Sulfinol-M	Claus Plant	Cyclone, candle filter, and water scrubber	LNB and N ₂ dilution	
B1B (6)	IGCC	1800/1000/1000	2 x State-of-the-art 2008 F-Class	Shell	95 mol% O ₂	Selexol	Claus Plant	Cyclone, candle filter, and water scrubber	LNB and N ₂ dilution	Selexol 2 nd stage
B4A (3)	IGCC	1800/1050/1050	2 x State-of-the-art 2008 F-Class	CB&I E-Gas™	95 mol% O ₂	Refrigerated MDEA	Claus Plant	Cyclone, candle filter, and water scrubber	LNB and N ₂ dilution	
B4B (4)	IGCC	1800/1000/1000	2 x State-of-the-art 2008 F-Class	CB&I E-Gas™	95 mol% O ₂	Selexol	Claus Plant	Cyclone, candle filter, and water scrubber	LNB and N ₂ dilution	Selexol 2 nd stage
B5A (1)	IGCC	1800/1050/1050	2 x State-of-the-art 2008 F-Class	GEE Radiant	95 mol% O ₂	Selexol	Claus Plant	Quench, water scrubber, and AGR adsorber	LNB and N ₂ dilution	
B5B (2)	IGCC	1800/1000/1000	2 x State-of-the-art 2008 F-Class	GEE Radiant	95 mol% O ₂	Selexol	Claus Plant	Quench, water scrubber, and AGR adsorber	LNB and N ₂ dilution	Selexol 2 nd stage
B5B-Q (2a)	IGCC	1800/1000/1000	2 x State-of-the-art 2008 F-Class	GEE Quench	95 mol% O ₂	Selexol	Claus Plant	Quench, water scrubber, and AGR adsorber	LNB and N ₂ dilution	Selexol 2 nd stage

^APrevious versions of this report used a different naming convention. The change was made to provide consistent naming conventions across various baseline reports published by NETL. The old case numbers are provided in parenthesis in this table and are paired with the new case numbers for reference, and will henceforth be referred to exclusively by their current nomenclature.

^BAll cases have a nominal 90 percent removal rate based on the total feedstock minus unburned carbon in slag. The rate of CO₂ capture from the syngas in the Selexol systems varies. An explanation for the difference is provided in Section 2.3.2. All capture cases sequester the CO₂ offsite.

2 General Evaluation Basis

For each of the plant configurations analyzed in this report, an Aspen Plus[®] (Aspen) model was developed and used to generate material and energy balances, which were, in turn, used to provide a design basis for items in the major equipment list. The equipment list and material balances were used as the basis for generating the capital and operating cost estimates.

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. Capital and operating costs were estimated by WorleyParsons based on simulation results using an in-house database and conceptual estimating models. Costs were further calibrated using a combination of adjusted vendor-furnished data and scaled estimates from previous design/build projects. These costs were established in 2007 dollars in prior reports and updated to 2011 dollars in this report. No changes were made to the performance or base cost methods in this update. Ultimately, a COE was calculated for each of the cases and is reported as the revenue requirement figure-of-merit.

The balance of this section discusses the design basis common to all technologies, as well as environmental targets and cost assumptions used in this report. Technology specific design criteria are covered in subsequent chapters.

2.1 Site Characteristics

All plants in this report are assumed to be located at a generic plant site in Midwestern United States (U.S.), with site characteristics and ambient conditions as presented in Exhibit 2-1 and Exhibit 2-2. The ambient conditions are the same as ISO conditions.

Exhibit 2-1 Site characteristics

Parameter	Value
Location	Greenfield, Midwestern U.S.
Topography	Level
Size, acres	300
Transportation	Rail or Highway
Slag Disposal	Off-Site
Water	50% Municipal and 50% Ground Water

Exhibit 2-2 Site ambient conditions

Parameter	Value
Elevation, (ft)	0
Barometric Pressure, MPa (psia)	0.101 (14.696)
Average Ambient Dry Bulb Temperature, °C (°F)	15 (59)
Average Ambient Wet Bulb Temperature, °C (°F)	10.8 (51.5)
Design Ambient Relative Humidity, %	60
Cooling Water Temperature, °C (°F) ^A	15.6 (60)
Air composition based on published psychrometric data, mass %	
N ₂	75.055
O ₂	22.998
Ar	1.280
H ₂ O	0.616
CO ₂	0.050
Total	100.00

^AThe cooling water temperature is the cooling tower cooling water exit temperature. This is set to 8.5°F above ambient wet bulb conditions in ISO cases.

The land area assumes that 30 acres are required for the plant proper, and the balance provides a buffer of approximately 0.25 mi to the fence line. The extra land could also provide for a rail loop if required (rail loop is not included in this report).

In all cases it was assumed that the steam turbine is enclosed in a turbine building. The gasifier is not enclosed.

The following design parameters are considered site-specific and are not quantified for this report. Allowances for normal conditions and construction are included in the cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Local code height requirements
- Noise regulations – Impact on site and surrounding area
- Other localized environmental concerns
- Weather delays

2.2 Coal Characteristics

The design coal is Illinois No. 6 with characteristics presented in Exhibit 2-3. The coal properties are from the January 2012 revision of the Quality Guidelines for Energy System Studies (QGESS) document “Detailed Coal Specifications.” (7)

Fuel costs used in this report are specified according to the November 2012 QGESS document “Fuel Prices for Selected Feedstocks in NETL Studies.” (8) The current coal price is \$2.94/MMBtu (HHV basis) for Illinois No. 6 bituminous coal delivered to the Midwest in June 2011 dollars.

Exhibit 2-3 Design coal

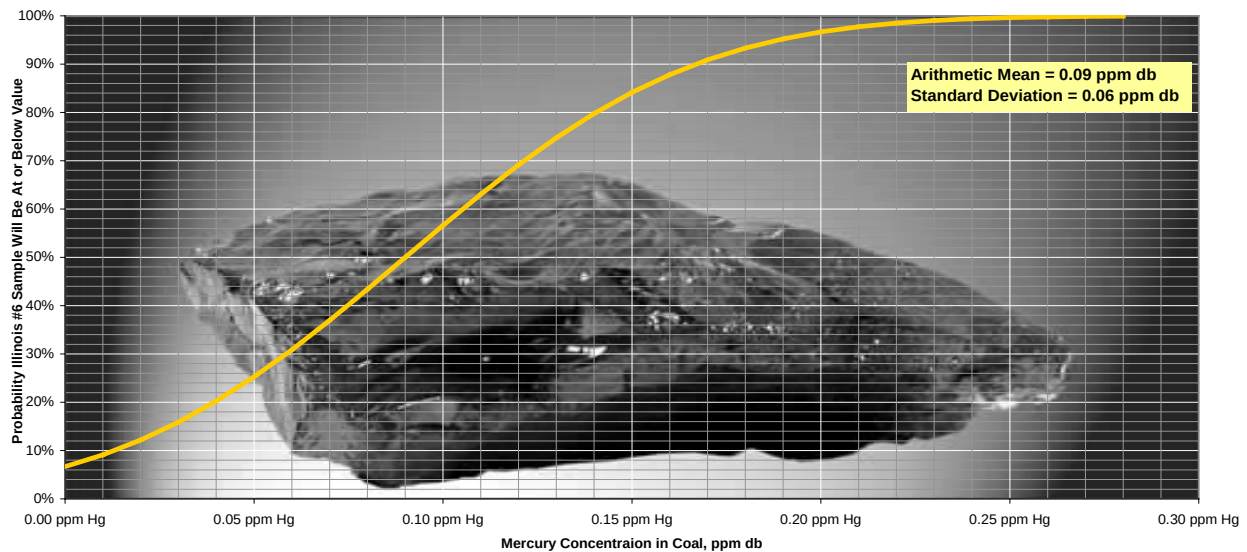
Rank	Bituminous	
Seam	Illinois No. 6 (Herrin)	
Source	Old Ben Mine	
Proximate Analysis (weight %) ^A		
	As Received	Dry
Moisture	11.12	0.00
Ash	9.70	10.91
Volatile Matter	34.99	39.37
Fixed Carbon	44.19	49.72
Total	100.00	100.00
Sulfur	2.51	2.82
HHV, kJ/kg (Btu/lb)	27,113 (11,666)	30,506 (13,126)
LHV, Btu/lb (Btu/lb)	26,151 (11,252)	29,544 (12,712)
Ultimate Analysis (weight %)		
	As Received	Dry
Moisture	11.12	0.00
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen ^B	6.88	7.75
Total	100.00	100.00

^AThe proximate analysis assumes sulfur as volatile matter

^BBy difference

The mercury content of 34 samples of Illinois No. 6 coal has an arithmetic mean value of 0.09 ppmwd with standard deviation of 0.06 based on coal samples shipped by Illinois mines. (9) Hence, as illustrated in Exhibit 2-4, there is a 50 percent probability that the mercury content in the Illinois No. 6 coal would not exceed 0.09 ppmwd. The coal mercury content for this report was assumed to be 0.15 ppmwd for all cases, which corresponds to the mean plus one standard deviation and encompasses about 84 percent of the samples. It was further assumed that all of the coal Hg enters the gas phase and none leaves with the slag. (10)

Exhibit 2-4 Probability distribution of mercury concentration in the Illinois No. 6 coal



Source: NETL

2.3 Environmental Targets

The environmental performance of the IGCC configurations described in this report were determined based upon the IGCC performance capabilities at the time of the plant model development in prior revisions. Since that time, specific environmental targets have been identified that require minimal adjustment to the previously designed systems. The only addition was the use of a dual bed carbon system for mercury (Hg) removal. The current environmental targets for this study are as follows:

- Mercury (Hg) and hydrochloric acid (HCl) limits were set by the March 2013 update to the Utility Mercury and Air Toxics Standards (MATS). (11), (12), (13)
- Particulate matter (PM), sulfur dioxide (SO₂), and nitrogen oxide (NO_x) limits were set by the February 2013 update to the New Source Performance Standards (NSPS). (13), (14)

The regulations divide the coal types into low rank and non-low rank based on their heating value. Coals with a higher heating value (HHV) of greater than 8,300 Btu per pound (Btu/lb) on a moist, mineral-matter free basis are considered non-low rank. Therefore, Illinois No. 6 coal, with an HHV (moist, mineral-matter free) of 12,900 Btu/lb, is considered a non-low rank coal.

The applicable emission limits imposed by MATS and NSPS are provided in Exhibit 2-5.

Exhibit 2-5 MATS and NSPS emission limits for PM, HCl, SO₂, NO_x, and Hg

Pollutant ^A	(lb/MWh-gross)
SO ₂	0.40
NO _x	0.70
PM (Filterable)	0.07
Hg	3x10 ⁻⁶
HCl	0.002

^ACO emissions may be considered in later revisions of this report, if necessary.

These new regulations apply to IGCC technologies that begin construction after May 3, 2011. Furthermore, these regulations state that (13), (15):

Fossil fuel is defined as natural gas, oil, coal, and any form of solid, liquid, or gaseous fuel derived from such material.

Electric utility steam generating units (EGU) are defined as a fossil fuel-fired combustion unit of more than 25 megawatts electric (MWe) that serves a generator that produces electricity for sale. A fossil fuel-fired unit that cogenerates steam and electricity and supplies more than one-third of its potential electric output capacity and more than 25 MWe output to any utility power distribution system for sale is considered an electric utility steam generating unit.

Fossil fuel-fired means an EGU that is capable of combusting more than 25 MW of fossil fuels. To be capable of combusting fossil fuels, an EGU would need to have these fuels allowed in its operating permit and have the appropriate fuel handling facilities on-site or otherwise available (e.g., coal handling equipment, including coal storage area, belts and conveyers, pulverizers, etc.; oil storage facilities).

Coal-fired electric utility steam generating units are defined as an EGU and meet the definition of “fossil fuel-fired,” which is that it burns coal for more than 10 percent of the average annual heat input during any three consecutive calendar years or for more than 15 percent of the annual heat input during any one calendar year.

Integrated gasification combined cycle electric utility steam generating units, or IGCCs, are defined as an EGU and meet the definition of “fossil fuel-fired,” which is that it burns a synthetic gas derived from coal and/or solid oil-derived fuel for more than 10 percent of the average annual heat input during any three consecutive calendar years or for more than 15 percent of the annual heat input during any one calendar year in a combined-cycle combustion turbine. No solid coal or solid oil-derived fuel is directly burned in the unit during operation.

Unit designed for low-rank virgin coal subcategory is defined as any coal-fired EGU that is designed to burn, and that is burning, non-agglomerating virgin coal having a calorific value (moist, mineral matter-free basis) of less than 19,305 kJ/kg (8,300 Btu/lb) that is constructed and operates at or near the mine that produces such coal.

Unit designed for coal $\geq 8,300$ Btu/lb subcategory is defined as any coal-fired EGU that is not a coal-fired EGU in the “unit designed for low rank virgin coal” subcategory.

Other regulations that could affect emissions limits from a new plant include the New Source Review (NSR) permitting process and Prevention of Significant Deterioration (PSD). The NSR process requires installation of emission control technology, meeting either the Best Available Control Technology (BACT) determinations for new sources being located in areas meeting ambient air quality standards (attainment areas), or Lowest Achievable Emission Rate (LAER) technology for sources being located in areas not meeting ambient air quality standards (non-attainment areas). Environmental area designation varies by county and can be established only for a specific site location. Based on the Environmental Protection Agency’s (EPA) Green Book Non-attainment Area Map, relatively few areas in the Midwestern U.S. are classified as “non-attainment,” so the plant site for this report was assumed to be in an attainment area. (14)

In addition to federal regulations, state and local jurisdictions can impose even more stringent regulations on a new facility. However, since each new plant has unique environmental requirements, it was necessary to apply some judgment in setting the environmental targets for this report.

2.3.1 Study Cases

Exhibit 2-6 provides the emissions limits for IGCC plants as well as a brief summary of the control technology utilized to satisfy the limits.

Exhibit 2-6 Environmental targets

Pollutant	(lb/MWh-gross)	Control Technology
SO ₂	0.40	Selexol, MDEA, or Sulfinol (depending on gasifier technology) ^A
NO _x	0.70	Low NO _x burners and syngas nitrogen dilution
PM (Filterable)	0.07	Quench, water scrubber, and/or cyclones and candle filters (depending on gasifier technology)
Hg	3x10 ⁻⁶	Dual carbon bed
HCl	0.002	Quench, water scrubber, calcium hydroxide treatment, sour water stripper

^AThe sulfur control technologies are used to remove H₂S formed in the gasifier to ultimately limit SO₂ emissions after the syngas is combusted in the CT.

Based on published vendor literature, it was assumed that low NO_x burners (LNB) and nitrogen dilution can achieve 15 ppmvd at 15 percent O₂, and that value was used for all cases. (16), (17)

The SO₂ limit is met via an acid gas removal (AGR) process. As the emissions limit is based on the gross power production of the plant, the actual removal efficiency is dependent on the net plant efficiency. Therefore, the required AGR H₂S removal efficiency varies from 75 to 98 percent in the cases without CO₂ capture. Vendor data on the AGR processes used in the non-capture cases indicate that the required level of sulfur removal in each technology is possible. The sulfur removal efficiency of the CO₂ capture cases is approximately 99.9 percent. The high rate of H₂S removal is a function of the CO₂ capture rate requirement (described in Section 2.3.2) of the two-stage Selexol process.

Most of the coal ash is removed from the gasifier as slag. The ash that remains entrained in the syngas is captured in the downstream equipment, including the syngas scrubber, cyclone, and either ceramic or metallic candle filters (E-GasTM and Shell). Each combination of particulate control devices can achieve the environmental target.

The Eastman Chemical plant, where syngas from a GEE gasifier is treated, achieved a mercury removal efficiency of 95 percent. Sulfur-impregnated activated carbon is used by Eastman as the adsorbent in the packed beds operated at 30°C (86°F) and 6.2 MPa (900 psig). Mercury removal between 90 and 95 percent has been reported with a bed life of 18 to 24 months. Removal efficiencies may be even higher, but, at 95 percent, the measurement precision limit was reached. Eastman has yet to experience any mercury contamination in its product. (18) As a mercury removal efficiency of up to 97 percent is required to meet the mercury emissions limit, a dual sulfur-impregnated carbon bed system (i.e., two beds in series) is required, which is capable of achieving greater than 99 percent mercury removal. It was assumed that the mercury removal

efficiency is linearly related to bed depth and, therefore, the bed depths were chosen to meet the emission limit.

The hydrochloric acid is removed primarily (approximately 95 percent) in the syngas scrubber. The remaining chloride in the syngas eventually drops out with condensed water downstream. The HCl removed in the syngas scrubber is sent to a treatment plant where calcium hydroxide is used to precipitate calcium chloride as a byproduct. The treated stream is mixed with the condensed water streams and sent to the sour water stripper, where the HCl is separated out with the vapor stream and sent to the Claus plant.

2.3.2 Carbon Dioxide

The EPA proposed a new source performance standard on April 13, 2012 for emissions of carbon dioxide for new fossil fuel-fired electric utility generating units. As of the publication of this report, the EPA is finalizing a rule that may set CO₂ emissions limits below that of fossil fuel-fired electric utility generation units without CO₂ capture.

For the cases that have CO₂ capture, the basis is a nominal 90 percent carbon removal rate based on carbon input from the coal and excluding carbon that exits the gasifier with the slag. In the GEE and Shell cases, this was accomplished by using two sour gas shift (SGS) reactors, to convert CO to CO₂, and a two-stage Selexol process with a second stage CO₂ removal efficiency of 92 percent, a number that was supported by vendor quotes. All gasifiers, except E-GasTM, achieve a nominal carbon removal rate of 90 percent. In the E-GasTM case, in an effort to achieve the capture target of 90 percent, a third SGS reactor was added, and the Selexol efficiency was increased to 95 percent (the maximum removal efficiency supported by vendor quotes). This was done because of the high syngas methane content (1.5 vol% compared to 0.10 vol% in the GEE radiant gasifier and 0.06 vol% in the Shell gasifier). The E-GasTM capture case resulted in 89.9 percent carbon capture.

2.4 Capacity Factor

2.4.1 Capacity Factor Assumptions

Availability is the percent of time during a specific period that a generating unit is capable of producing electricity. This report assumes that each new plant would be dispatched any time it is available and would be capable of generating the nameplate capacity when online. Therefore, the capacity factor (CF) and availability are equal. The operating period selected is also important. The calculations assume that the capacity factor and availability are constant over the life of the plant, but in actual operation may require that a plant have a higher peak availability to counter lower availability in the first several years of operation.

2.4.2 Capacity Factor for IGCC Plants

The Electric Power Research Institute (EPRI) has reported an availability goal for IGCC plants of 85 percent. (19) Plants built before 2000 have achieved availability of 80 percent for limited periods of time. Common projections from technology suppliers and EPRI are that IGCC plants are capable of 80-85 percent availability without a spare gasifier and could achieve greater than 90 percent availability with a spare gasifier. (20), (21), (22), (23), (24), (25) While an availability of 85 percent is the goal, given the IGCC technology experience and the commercialization status compared with the more conventional plants (e.g., PC), a CF of 80 percent was selected for IGCC plants with no spare gasifier.

2.4.3 Capacity Factor for Plants with Carbon Capture

The addition of carbon capture adds extra equipment to the power plant. Preliminary reliability analyses show small reductions in reliability if the reliability of the base plant components is kept constant. A solvent-based carbon capture technology is used in this report. The capture and CO₂ compression technologies have commercial operating experience with demonstrated ability for high reliability. Given the use of commercial technology, the assumption is made that the capacity factors for a given plant with and without carbon capture are the same. Thus, the capacity factor for IGCC plants with capture is 80 percent.

2.4.4 Perspective

Important factors required to achieve the projected IGCC plant availability include a quality plant design that utilizes lessons learned from similar plant designs, a focus on life cycle costs, a smart predictive maintenance, a trained plant staff, and an economic demand for unit power.

Plant availability is determined by the plant technology, the capital cost invested in the plant (e.g., what is the design approach with respect to minimizing scheduled and unplanned maintenance), the maintenance requirements, and the customer requirements for the electricity (e.g., customer costs due to a unit not being available). Since the unavailability cost will decrease with increasing unit availability and the maintenance and capital costs increase with increasing unit availability, there will be optimum economic unit availability for a given application. This report assumes that the plant design, plant maintenance, and electricity demand are consistent with the selected availability.

The existing plant data have not been analyzed with regard to the performance of individual plant availability over the life of the plant. As stated, this report assumes a constant availability of 80 percent for each year over the life of the plant. It is recognized that the availability of a given plant will vary over the life of the plant. As demonstrated by existing plant data, coal plants can be designed and operated with yearly availability ranging from 85-100 percent. It is assumed that the plants in this report will have yearly availability factors above and below the selected value with the effective or levelized availability for the life of the plant being the selected value.

2.5 Raw Water Withdrawal and Consumption

A water balance was performed for each case on the major water consumers in the process. The total water demand for each subsystem was determined, and internal recycle water available from various sources like boiler feedwater (BFW) blowdown and condensate from syngas (in CO₂ capture cases) 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 municipal 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 from which it was withdrawn.

Raw water makeup was assumed to be provided 50 percent by a publicly owned treatment works (POTW) and 50 percent from groundwater. Raw water withdrawal is defined as the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, slurry preparation makeup, syngas humidification, and

quench system makeup. 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.

BFW blowdown and a portion of the sour water stripper blowdown were assumed to be treated and recycled to the cooling tower. The cooling tower blowdown and the balance of the SWS blowdown streams were assumed to be treated and 90 percent returned to the water source with the balance sent to the ash ponds for evaporation.

The largest consumer of raw water in all cases is cooling tower makeup. It was assumed that all cases utilized a mechanical draft, evaporative cooling tower, and all process blowdown streams were assumed to be treated and recycled to the cooling tower. The design ambient wet bulb temperature of 11°C (51.5°F) (Exhibit 2-1 and Exhibit 2-2) was used to achieve a cooling water temperature of 16°C (60°F) using an approach of 5°C (8.5°F). The cooling water range was assumed to be 11°C (20°F). The cooling tower makeup rate was determined using the following (26):

- Evaporative losses of 0.8 percent of the circulating water flow rate per 10°F of range
- Drift losses of 0.001 percent of the circulating water flow rate
- Blowdown losses (BDL) were calculated as follows:

$$BDL = \frac{EL}{CC - 1}$$

Where:

EL – Evaporative Losses

CC – Cycles of concentration

The cycles of concentration are a measure of water quality and a mid-range value of four was chosen for this report.

The water balances presented in subsequent sections include the water demand of the major water consumers within the process, the amount provided by internal recycle, the amount of raw water withdrawal by difference, the amount of process water returned to the source, and the raw water consumption, again by difference.

2.6 Cost Estimating Methodology

Detailed information pertaining to topics such as contracting strategy, engineering, procurement, and construction (EPC) contractor services, estimation of capital cost contingencies, owner's costs, cost estimate scope, economic assumptions, finance structures, cost of electricity, etc. are available in the April 2011 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." (2) Select portions are repeated in this report for completeness.

Capital Costs:

The capital cost estimates documented in this report reflect an uncertainty range of -15 percent/+30 percent, consistent with Association for the Advancement of Cost Engineering (AACE) Class 4 cost estimates (i.e., feasibility study) (4) (5) (6), based on the level of engineering design performed. In all cases, this report intends to represent the next commercial offering and relies on vendor cost estimates for component technologies. It also applies process

contingencies at the appropriate subsystem levels in an attempt to account for expected but undefined costs, which can be a challenge for emerging technologies.

Costs of Emerging Technologies and Designs:

The cost estimates for plant designs that include technologies that are not yet fully mature (e.g., IGCC plants and any plant with CO₂ capture) use the same cost estimating methodology as for mature plant designs (e.g., PC and NGCC power plants without CO₂ capture), which does not fully account for the unique cost premiums associated with the initial, complex integrations of emerging technologies in a commercial application. Thus, it is anticipated that initial deployments of IGCC plants – both with and without CO₂ capture – may incur costs higher than those reflected within this report.

Other Factors:

Actual reported project costs for all of the plant types are also expected to deviate from the cost estimates in this report due to project- and site-specific considerations (e.g., contracting strategy, local labor costs, seismic conditions, water quality, financing parameters, local environmental concerns, weather delays, etc.) that may make construction more costly. Such variations are not captured by the reported cost uncertainty.

Future Cost Trends:

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

2.6.1 Capital Costs

As illustrated in Exhibit 2-7, this report defines capital cost at five levels: BEC, EPCC, TPC, TOC, and TASC. BEC, EPCC, TPC, and TOC are “overnight” costs and are expressed in “base-year” dollars. The base year is the first year of capital expenditure. TASC is expressed in mixed, current-year dollars over the entire capital expenditure period, which is assumed in most NETL studies to last five years for coal plants.

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

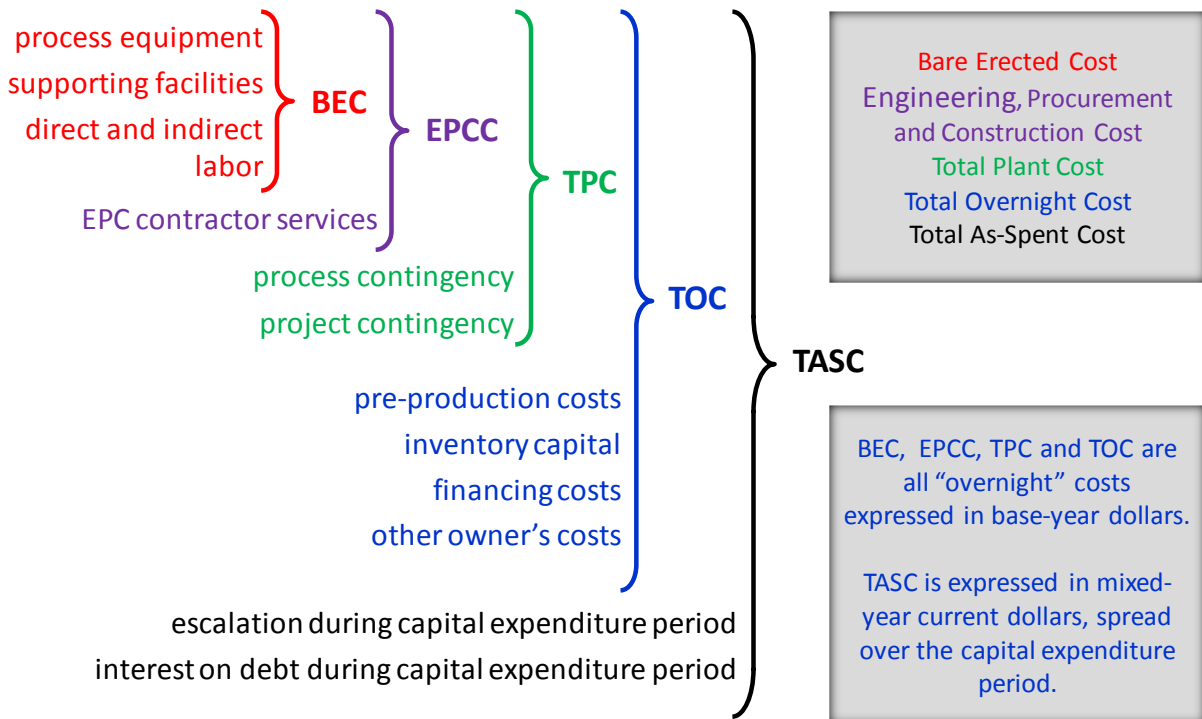
The Engineering, Procurement and Construction Cost (EPCC) comprises the BEC plus the cost of services provided by the EPC contractor. EPC services include: detailed design, contractor permitting (i.e., those permits that individual contractors must obtain to perform their scopes of work, as opposed to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

The Total Plant Cost (TPC) comprises the EPCC plus project and process contingencies. TPC is an overnight cost expressed in base-year dollars.

The Total Overnight Cost (TOC) comprises the TPC plus all other overnight costs, including owner’s costs. TOC is an “overnight” cost, expressed in base-year dollars and as such does not include escalation during construction or interest during construction.

The Total As-Spent Cost (TASC) is the sum of all capital expenditures as they are incurred during the capital expenditure period including their escalation. TASC also includes interest during construction. Accordingly, TASC is expressed in mixed, current-year dollars over the capital expenditure period.

Exhibit 2-7 Capital cost levels and their elements



Source: NETL

2.6.1.1 Cost Estimate Basis and Classification

The TPC and Operation and Maintenance (O&M) costs for each of the cases in the report were estimated by WorleyParsons using an in-house database and conceptual estimating models. Costs were further calibrated using a combination of adjusted vendor-furnished data and scaled estimates from previous design/build projects.

2.6.1.2 System Code-of-Accounts

The costs are grouped according to a process/system oriented code of accounts. This type of code-of-account structure has the advantage of grouping all reasonably allocable components of a system or process so they are included in the specific system account. (This would not be the case had a facility, area, or commodity account structure been chosen instead).

2.6.1.3 Estimate Scope

The estimates represent a complete power plant facility on a generic site. The plant boundary limit is defined as the total plant facility within the "fence line" including coal receiving and water supply system, but terminating at the high voltage side of the main power transformers. CO₂ T&S cost is not included in the reported capital cost or O&M costs, but is treated separately and added to the COE.

2.6.1.4 Capital Cost Assumptions

WorleyParsons developed the capital cost estimates for each plant using the company's in-house database and conceptual estimating models for each of the specific technologies. This database and the respective models are maintained by WorleyParsons as part of a commercial power plant design base of experience for similar equipment in the company's range of power and process projects. A reference bottom-up estimate for each major component provides the basis for the estimating models.

Other key estimate considerations include the following:

- Labor costs are based on Midwest, Merit Shop. The estimating models are based on U.S. Gulf Coast and the labor has been factored to Midwest. The basis for the factors is the PAS, Inc. (PAS) "Merit Shop Wage & Benefit Survey," which is published annually. Based on the data provided in PAS, WorleyParsons used the weighted average payroll plus fringe rate for a standard craft distribution as developed for the estimating models. PAS presents information for eight separate regions. For this report, Region 5 (IL, IN, MI, MN, OH, and WI) was selected.
- The estimates are based on a competitive bidding environment, with adequate skilled craft labor available locally.
- Labor is based on a 50-hour work-week (5-10s). No additional incentives such as per-diem allowances or bonuses have been included to attract craft labor.
- While not included at this time, labor incentives may ultimately be required to attract and retain skilled labor depending on the amount of competing work in the region, and the availability of skilled craft in the area at the time the projects proceed to construction.
- The estimates are based on a greenfield site.
- The site is considered to be Seismic Zone 1, relatively level, and free from hazardous materials, archeological artifacts, or excessive rock. Soil conditions are considered adequate for spread footing foundations. The soil bearing capability is assumed adequate such that piling is not needed to support the foundation loads.
- Engineering and Construction Management are estimated at 8-10 percent of BEC. These costs consist of all home office engineering and procurement services as well as field construction management costs. Site staffing generally includes a construction manager, resident engineer, scheduler, and personnel for project controls, document control, materials management, site safety, and field inspection.

2.6.1.5 Price Fluctuations

During the course of this report, the prices of equipment and bulk materials fluctuated substantially. Some reference quotes pre-dated the 2011 year cost basis while others were received post-2011. All vendor quotes used to develop these estimates were adjusted to June 2011 dollars accounting for the price fluctuations. Price indices, e.g., The Chemical Engineering Plant Cost Index (27) and the Gross Domestic Product Chain-type Price Index (28), were used as needed for these adjustments.

2.6.1.6 Cross-comparisons

In all technology comparison studies, the relative differences in costs are often more significant than the absolute level of TPC. This requires cross-account comparison between technologies to review the consistency of the direction of the costs.

In performing such a comparison, it is important to reference the technical parameters for each specific item, as these are the basis for establishing the costs. Scope or assumption differences can quickly explain any apparent anomalies. There are a number of cases where differences in design philosophy occur. Some key examples are:

- The CT account in the GEE cases includes a syngas expander, which is not required for the E-GasTM or Shell cases.
- The CTs for the capture cases include an additional cost for firing a high hydrogen content fuel.
- The Shell gasifier syngas cooling configuration is different between the CO₂-capture and non-CO₂-capture cases, resulting in a significant differential in thermal duty between the syngas coolers for the two cases.

2.6.1.6.1 Process Contingency

Process contingencies were applied to the estimates in this report as follows:

- Slurry Prep and Feed – 5 percent on GEE cases - systems are operating at approximately 800 psia as compared to 600 psia for the other cases
- Gasifiers and Syngas Coolers – 15 percent on all cases - next-generation commercial offering and integration with the power island
- Two-Stage Selexol – 20 percent on all capture cases - unproven technology at commercial scale in IGCC service
- Mercury Removal – 5 percent on all cases - minimal commercial scale experience in IGCC applications
- CTG – 5 percent on all non-capture cases - syngas firing and ASU integration; 10 percent on all capture cases – high hydrogen firing.
- Instrumentation and Controls – 5 percent on all accounts– integration issues

2.6.1.7 Owner's Costs

Two examples of what could be included in the “other” owner’s costs are rail spur and switch yard costs.

Switch yard costs are dependent on voltage, configuration, number of breakers, layout, and air-insulated vs. gas-insulated. As a rule of thumb, a 345 kV switchyard (air-insulated, ring bus) would cost roughly \$850,000 per breaker.

On-site only rails (excludes long runs) would be expected to cost in the range of \$850,000 to \$950,000 per mile (relatively flat level terrain) plus the costs of any switches/turnouts (approximately \$50,000 each) and road crossings (approximately \$300 per linear foot).

Additional details and explanation of owner’s costs are available in the April 2011 revision of the QGESS document “Cost Estimation Methodology for NETL Assessment of Power Plant Performance.” (2)

2.6.2 Operation and Maintenance Costs

The production costs or operating costs and related maintenance expenses (O&M) pertain to those charges associated with operating and maintaining the power plants over their expected life. These costs include:

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

There are two components of O&M costs; fixed O&M, which is independent of power generation, and variable O&M, which is proportional to power generation. Taxes and insurance are included as fixed O&M costs, totaling 2 percent of the TPC.

2.6.2.1 Operating Labor

Operating labor cost was determined based on of the number of operators required for each specific case. The average base labor rate used to determine annual cost is \$39.70/hour. The associated labor burden is estimated at 30 percent of the base labor rate.

2.6.2.2 Maintenance Material and Labor

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

2.6.2.3 Administrative and Support Labor

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

2.6.2.4 Consumables

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

Quantities for major consumables such as fuel and sorbent were taken from technology-specific heat and mass balance diagrams developed for each plant application. Other consumables were evaluated on the basis of the quantity required using reference data.

The quantities for initial fills and daily consumables were calculated on a 100 percent operating capacity basis. The annual cost for the daily consumables was then adjusted to incorporate the annual plant operating basis, or CF.

Initial fills of the consumables, fuels and chemicals, are different from the initial chemical loadings, which are included with the equipment pricing in the capital cost.

2.6.2.5 Waste Disposal

Waste quantities and disposal costs were determined/evaluated similarly to the consumables. In this report slag is considered a waste with a disposal cost of \$27.80/tonne (\$25.11/ton). The carbon used for mercury control is considered a hazardous waste with disposal cost of \$1,433/tonne (\$1,300/ton).

2.6.2.6 Co-Products and By-Products

By-product quantities were also determined similarly to the consumables. However, due to the variable marketability of these by-products, specifically sulfur, no credit was taken for its potential salable value.

It should be noted that by-product credits and/or disposal costs could potentially be an additional determining factor in the choice of technology for some companies and in selecting some sites. A high local value of the product can establish whether or not added capital should be included in the plant costs to produce a particular co-product. Slag is a potential by-product in certain markets. As stated above, the slag is considered a waste in this report with a concomitant disposal cost.

2.6.3 CO₂ Transport and Storage

The cost of CO₂ transport and storage (T&S) in a deep saline formation is estimated using the FE/NETL CO₂ Transport Cost Model and the FE/NETL CO₂ Saline Storage Cost Model. Additional detail on development of these costs is available in the May 2014 revision of the QGESS document “Carbon Dioxide Transport and Storage Costs in NETL Studies.” (29)

T&S costs are reported as first-year costs in \$/tonne of CO₂, increasing at a nominal rate of 3 percent per year, consistent with the general inflation rate assumed in NETL’s energy systems studies. From the perspective of the CO₂ source (e.g., a power plant or other energy conversion facility), these costs are treated as a disposal cost for each tonne of CO₂ captured during the assumed 30-year operational period. From the pipeline and storage site’s perspective, this cost represents the minimum price that the operators must charge so that they receive the revenue needed across the 30-year operational period to cover all costs and provide the required internal rate of return on equity (IRROE). All costs are reported in 2011 dollars.

The Transport Cost Model provides cost estimates for the construction and operation of a dedicated pipeline for transporting CO₂ from a CO₂ source to a CO₂ storage site. The pipeline is assumed to be buried in the shallow subsurface and the CO₂ along the pipeline is assumed to be at the temperature of the surrounding soil, typically about 50 °F. The CO₂ entering the pipeline at the plant gate is assumed to be at a pressure of 2,200 psig and exiting the pipeline at the storage site at a pressure of 1,200 psig. At these pressures, CO₂ is a liquid at temperatures from approximately -64°F to 88°F.

It is further assumed that the pipeline is 100 km long (62.1 mi) and transports 3.2 million tonnes of CO₂ each year on average or approximately 8,770 tonnes per day on average. The pipeline is designed to operate at 80 percent of its maximum mass flow capacity, so the design maximum daily mass flow rate of CO₂ is about 10,960 tonnes/day. The model determines the smallest standard diameter pipe that can transport this mass flow of CO₂ the required distance without boosting the pressure. The model also determines the smallest standard diameter pipe that can transport this mass flow of CO₂ the required distance assuming boost pumps are placed at equal intervals along the pipeline to boost the pressure from 1,200 psig to 2,200 psig. The model then determines which of these configurations (e.g., no boost pumps, one boost pump, two boost pumps, three boost pumps, etc.) is least expensive. The capital and operating costs used in the model were taken from the open literature. The capital costs for the pipeline are based on capital costs for natural gas pipelines reported in the Oil and Gas Journal with adjustments for the higher pressures used in CO₂ pipelines.

The financial parameters used in the FE/NETL CO₂ Transport Cost Model are:

- Debt to equity ratio: 50%/50%
- Nominal interest rate on debt: 4.5%/year
- Nominal IRROE: 12%
- Escalation rate: 3%
- Tax rate: 38%
- Project contingency factor of 15%

The FE/NETL CO₂ Transport Cost Model determines the first-year break-even price that needs to be charged to transport the CO₂ in order to cover all costs including the minimum return on equity. The model uses a weighted average cost of capital methodology in determining the break-even first-year CO₂ price. From the perspective of the CO₂ source, the break-even price is also the minimum cost of transporting CO₂. Transport costs are estimated to be \$2.24/tonne in 2011 dollars.

Storage costs are based on the FE/NETL CO₂ Saline Storage Cost Model. This model provides detailed cost estimates for the injection and monitoring of CO₂ under EPA regulations for Class VI injection wells as well as monitoring and reporting requirements under Subpart RR of the Greenhouse Gas Reporting Rule.

Inputs to the FE/NETL CO₂ Saline Storage Cost Model that have a significant influence on cost include financial parameters, timelines for the various stages of storage, and important activities occurring in each stage. The financial parameters include:

- Debt to equity ratio: 45%/55%
- Nominal interest rate on debt: 5.5%/year
- Nominal IRROE: 12%
- Escalation rate: 3%
- Financial responsibility requirements for post-injection site care and site closure are met by pre-funding a modified trust fund over the period of injection operations
- Project contingency factor of 15% and process contingency factor of 20%

In the FE/NETL CO₂ Saline Storage Cost Model, the sequestration process is divided into six stages. The timelines and important activities impacting costs for these stages are as follows:

- Regional evaluation and initial site selection: 1 year
- Site characterization: 3 years; four sites undergo site characterization with one successful site selected; pore space rights are leased
- Permitting: 2 years; drill, test, and complete injection wells
- Operations: 30 years; installation of buildings, surface equipment, monitoring wells and other monitoring equipment; comply with permit requirements; fund modified trust fund to cover financial responsibility requirements for post-injection site care and site closure
- Post-injection site care and site closure: 50 years; continue monitoring, verification and accounting (mega volt-amps [MVA]) per permit; costs are covered by storage site operator's trust fund
- Long-term stewardship: This stage is not explicitly included in the model; however, the possible financial implication of long-term stewardship is included in the model as a state-sponsored trust fund that the storage operator pays into during operations

Due to the variances in the geologic formations that make up saline formations across the U.S., region-specific storage and monitoring costs are developed to correspond to the plant locations used in NETL techno-economic studies of energy conversion facilities. Results from the FE/NETL CO₂ Saline Storage Cost Model for storage and monitoring costs were aligned with the NETL studies by taking four generic plant locations and overlaying them with possible sequestration basins from the cost model resulting in the following pairings:

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

CO₂ storage supply-cost curves were developed for each of the 4 basins of interest with the resulting cost for each basin at 25 gigatonnes (Gt) of potential storage shown in Exhibit 2-8. Choosing this point on the supply-cost curves provides a conservative estimate of the storage cost since many decades, if not more than a century, will pass before 25 Gt of CO₂ is stored in any of the 4 individual basins. For example, 25 Gt of storage would be sufficient for 125 GW of coal power with 90 percent CO₂ capture operating over 30 years.

The far right column of Exhibit 2-8 shows the total T&S costs used in NETL system studies for each plant location rounded to the nearest whole dollar. Only the \$11/tonne value is used in this volume of the baseline study report since all cases are located in the Midwest.

Exhibit 2-8 CO₂ transport and storage costs

Plant Location	Basin	Transport (2011 \$/tonne)	Storage Cost at 25 Gt (2011 \$/tonne)	T&S Value for System Studies ^A (2011\$/tonne)
Midwest	Illinois	2.24	8.69	11
Texas	East Texas		8.83	11
North Dakota	Williston		13.95	16
Montana	Powder River		21.81	24

^AThe sum of transport and storage costs is rounded to the nearest dollar

2.6.4 Cost of CO₂ Captured and Avoided

The cost of captured CO₂ represents the minimum CO₂ plant gate sales price that will incentivize carbon capture in lieu of a defined reference non-capture plant. The cost of captured CO₂ is calculated using the following formula:

$$\text{Cost of CO}_2 \text{ Captured} = \frac{(COE_{CCS} - COE_{Non\ CCS})}{CO_2 \text{ Captured}}$$

The cost of CO₂ avoided represents the minimum CO₂ emissions price, when applied to both the capture and non-capture plant that will incentivize carbon capture in lieu of a defined reference non-capture plant. The cost of CO₂ avoided is calculated using the following formula:

$$\text{Cost of CO}_2 \text{ Avoided} = \frac{(COE_{CCS\ with\ T\ \&\ S} - COE_{Non\ CCS})}{CO_2 \text{ Emissions}_{Non\ CCS} - CO_2 \text{ Emissions}_{CCS}}$$

where:

- CCS – the capture plant for which the cost of CO₂ captured/avoided is being calculated
- Non-CCS – the reference non-capture plant, as described below
- COE – the cost of electricity, reported in \$/MWh
 - o The CCS plant includes compression to 2,215 psia
 - o For CO₂ Captured, the COE excludes T&S costs
 - o For CO₂ Avoided, the COE includes T&S costs
- CO₂ Captured – the rate of CO₂ captured, reported in tonne/MWh
- CO₂ Emissions – the rate of CO₂ emitted out the stack, reported in tonne/MWh

For today’s greenfield coal with CCS plants, the reference non-capture plant used to calculate the cost of captured and avoided CO₂ is a supercritical pulverized coal (SCPC) plant without capture.

2.7 Differences in IGCC Study Cost Estimates Compared to Industry Estimates

2.7.1 Project Scope

For this report, the scope of work is generally limited to work inside the project “fence line.” For outgoing power, the scope stops at the high voltage side of the main power transformers.

Some typical examples of items outside the fenceline include:

- New access roads and railroad tracks
- Upgrades to existing roads to accommodate increased traffic
- Makeup water pipe outside the fence line
- Landfill for on-site waste (slag) disposal
- Electrical transmission lines & substation

Estimates in this report are based on a generic mid-western greenfield site having “normal” characteristics. Accordingly, the estimates do not address items such as:

- Piles or caissons
- Rock removal
- Excessive dewatering
- Expansive soil considerations
- Excessive seismic considerations
- Extreme temperature considerations
- Hazardous or contaminated soils
- Demolition or relocation of existing structures
- Leasing of offsite land for parking or laydown
- Busing of craft labor to site
- Costs of offsite storage

This report is based on a reasonably “standard” plant. Some unusual or extraordinary process equipment that is not included in the analysis are as follows:

- Excessive water treatment equipment

- Air-cooled condenser
- Automated coal reclaim
- Zero liquid discharge equipment
- SCR catalyst

This report is based on plant equipment sized for non-capture only; none of the equipment has been sized to accommodate a future conversion to CO₂ capture.

2.7.2 Labor

This report is based on Merit Shop (non-union) labor. If a project is to use union labor, there is a strong likelihood that overall labor costs will be greater than those estimated in this report.

This report is based on a 50 hour work week, with an adequate local supply of skilled craft labor. No additional incentives such as per-diem allowances or bonuses have been included to attract and retain skilled craft labor.

2.7.3 Contracting Methodology

The estimates in this report are based on a competitively bid, multiple subcontract approach, often referred to as Engineering/Procurement/Construction Management (EPCM). Accordingly, the estimates do not include premiums associated with an EPC approach. It is believed that, given current market conditions, the premium charged by an EPC contractor could be as much as 30 percent or more over an EPCM approach.

3 IGCC Power Plants

Seven IGCC power plant configurations were evaluated, and the results are presented in this section. Each design is based on a market-ready technology that is assumed to be commercially available to support startup.

The seven cases are based on the GEE gasifier, the CB&I E-Gas™ gasifier, and the Shell gasifier, each with and without CO₂ capture. As discussed in Section 1, the net output for the seven cases varies because of the constraint imposed by the fixed CT output and the high auxiliary loads imparted by the CO₂ capture process.

The CT is based on a state-of-the-art 2008 F-class design. The HRSG/steam turbine cycle varies based on the CT exhaust conditions. Steam conditions are nominally 12.4 MPa/560°C/560°C (1800 psig/1040°F/1040°F) for all of the non-CO₂ capture cases and 12.4 MPa/534°C/534°C (1800 psig/994°F/994°F) for all of the CO₂ capture cases. The capture cases have a lower main and reheat steam temperature primarily because the turbine firing temperature is reduced to allow for a parts life equivalent to NGCC operation with a high-hydrogen content fuel, which results in a lower turbine exhaust temperature.

The evaluation scope included developing heat and mass balances and estimating plant performance. Equipment lists were developed for each design to support plant capital and operating cost estimates. The evaluation basis details, including site ambient conditions, fuel composition, and environmental targets are provided in Section 2. Section 3.1 covers general information that is common to all of the cases; case-specific information is subsequently presented in Sections 3.2, 3.3, and 3.4.

3.1 IGCC Common Process Areas

The cases have process areas, which are common to each plant configuration such as coal receiving and storage, oxygen supply, gas cleanup, power generation, etc. As detailed descriptions of these process areas for each case would be burdensome and repetitious, they are presented in this section for general background information. Where there is case-specific performance information, the performance features are presented in the relevant case sections.

3.1.1 Coal Receiving and Storage

The function of the coal receiving and storage system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to and including the slide gate valves at the outlet of the coal storage silos. Coal receiving and storage is identical in design for all seven cases; however, coal preparation and feed are gasifier-specific.

Operation Description – The coal is delivered to the site by 100-car unit trains comprised of 91 tonne (100 ton) rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal into two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 8 cm x 0 (3" x 0) coal from the feeder is discharged onto a belt conveyor. Two conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker, which transfer the coal to either the long-term storage pile or to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

The reclaimer loads the coal into two vibratory feeders located in the reclaim hopper under the pile. The feeders transfer the coal onto a belt conveyor that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3 cm x 0 (1¼" x 0) by the crusher. A conveyor then transfers the coal to a transfer tower. In the transfer tower the coal is routed to the tripper, which loads the coal into one of three silos. Two sampling systems are supplied: the as-received sampling system and the as-fired sampling system. Data from the analyses are used to support the reliable and efficient operation of the plant.

3.1.2 Air Separation Unit (ASU) Choice and Integration

In order to economically and efficiently support IGCC projects, air separation equipment has been modified and improved in response to production requirements and the consistent need to increase single train output. “Elevated pressure” air separation designs have been implemented that result in distillation column operating pressures that are about twice as high as traditional plants. In this report, the main air compressor discharge pressure was set at 1.3 MPa (190 psia) compared to a traditional ASU plant operating pressure of about 0.7 MPa (105 psia). (30) For IGCC designs the elevated pressure ASU process minimizes power consumption and decreases the size of some of the equipment items. When the air supply to the ASU is integrated with the CT, the ASU operates at or near the supply pressure from the CT’s air compressor.

The power requirement for the ASU used in this report is 0.039 kW/lb-O₂ product for all cases and the total power requirement, including the oxygen and nitrogen compressors, ranges from 0.064 kW/lb-O₂ product in the E-GasTM CO₂ capture case (B4B) to 0.072 kW/lb-O₂ product in the Shell non-capture case (B1A).

3.1.2.1 Residual Nitrogen Injection

The residual nitrogen that is available after gasifier oxygen and nitrogen requirements have been met is often compressed and sent to the CT. Since all product streams are being compressed, the ASU air feed pressure is optimized to reduce the total power consumption and to provide a good match with available compressor frame sizes.

Increasing the diluent flow to the CT by injecting residual nitrogen from the ASU can have a number of benefits, depending on the design of the CT:

- Increased diluent increases mass flow through the turbine, thus increasing the power output of the CT while maintaining optimum firing temperatures for syngas operation. This is particularly beneficial for locations where the ambient temperature and/or elevation are high and the CT would normally operate at reduced output.
- By mixing with the syngas or by being injected directly into the combustor, the diluent nitrogen lowers the firing temperature (relative to natural gas) and reduces the formation of thermal NO_x.

In this report, the ASU nitrogen product was used as the primary diluent with a design target of reducing the syngas lower heating value (LHV) to 4.4-4.7 MJ/Nm³ (119-125 Btu/scf). If the amount of available nitrogen was not sufficient to meet this target, additional dilution was provided through syngas humidification; if still more dilution was required, the third option was steam injection.

3.1.2.2 Air Integration

Integration between the ASU and the CT can be practiced by extracting some, or all, of the ASU's air requirement from the CT. Medium Btu syngas streams result in a higher mass flow than natural gas to provide the same heat content to the CT. Some CT designs may need to extract air to maintain stable compressor or turbine operation in response to increased fuel flow rates. Other CTs may balance air extraction against injection of all of the available nitrogen from the ASU. The amount of air extracted can also be varied as the ambient temperature changes at a given site to optimize year-round performance.

An important aspect of air-integrated designs is the need to efficiently recover the heat of compression contained in the air extracted from the CT. Extraction air temperature is normally in the range of 399 - 454°C (750 - 850°F) and must be cooled to the last stage main air compressor discharge temperature prior to admission to the ASU. High-level recovery from the extracted air occurs by transferring heat to the nitrogen stream to be injected into the CT with a gas-to-gas heat exchanger.

It was not a goal of this project to optimize the integration of the CT and the ASU, although several papers have shown that providing 25-30 percent of the ASU air from the turbine compressor provides the best balance between maximizing plant output and efficiency without compromising plant availability or reliability. (31), (32)

3.1.2.3 Elevated Pressure ASU Experience in Gasification

The Buggenum, Netherlands unit built for Demkolec was the first elevated-pressure, fully integrated ASU to be constructed. It was designed to produce up to 1,796 tonnes/day (1,980 tons per day [tpd]) of 95 percent purity oxygen for a Shell coal-based gasification unit that fuels a Siemens V94.2 CT. In normal operation at the Buggenum plant the ASU receives all of its air supply from and sends all residual nitrogen to the CT. (23)

The Polk County, Florida ASU for the Tampa Electric IGCC is also an elevated-pressure, 95 percent purity oxygen design that provides 1,832 tonnes/day (2,020 tpd) of oxygen to a GEE coal-based gasification unit, which fuels a General Electric 7FA CT. All of the nitrogen produced in the ASU is used in the CT. The original design did not allow for air extraction from the CT. After a CT air compressor failure in January, 2005, a modification was made to allow air extraction, which in turn eliminated a bottleneck in ASU capacity and increased overall power output. (33)

3.1.2.4 ASU Basis

For this report, air integration is used for the non-carbon capture cases only. In the CO₂ capture cases, once the syngas is diluted to the target heating value, all of the available combustion air is required to maintain mass flow through the turbine and hence maintain power output.

The amount of air extracted from the CT in the non-capture cases is determined through a process that includes the following constraints:

- The output of each CT must be maintained at 232 MW
- The diluted syngas must meet heating value requirements specified by a CT vendor, which ranged from 4.4-4.7 MJ/Nm³ (119-125 Btu/scf)

Meeting the above constraints resulted in different levels of air extraction in the three non-carbon capture cases as shown in Exhibit 3-1.

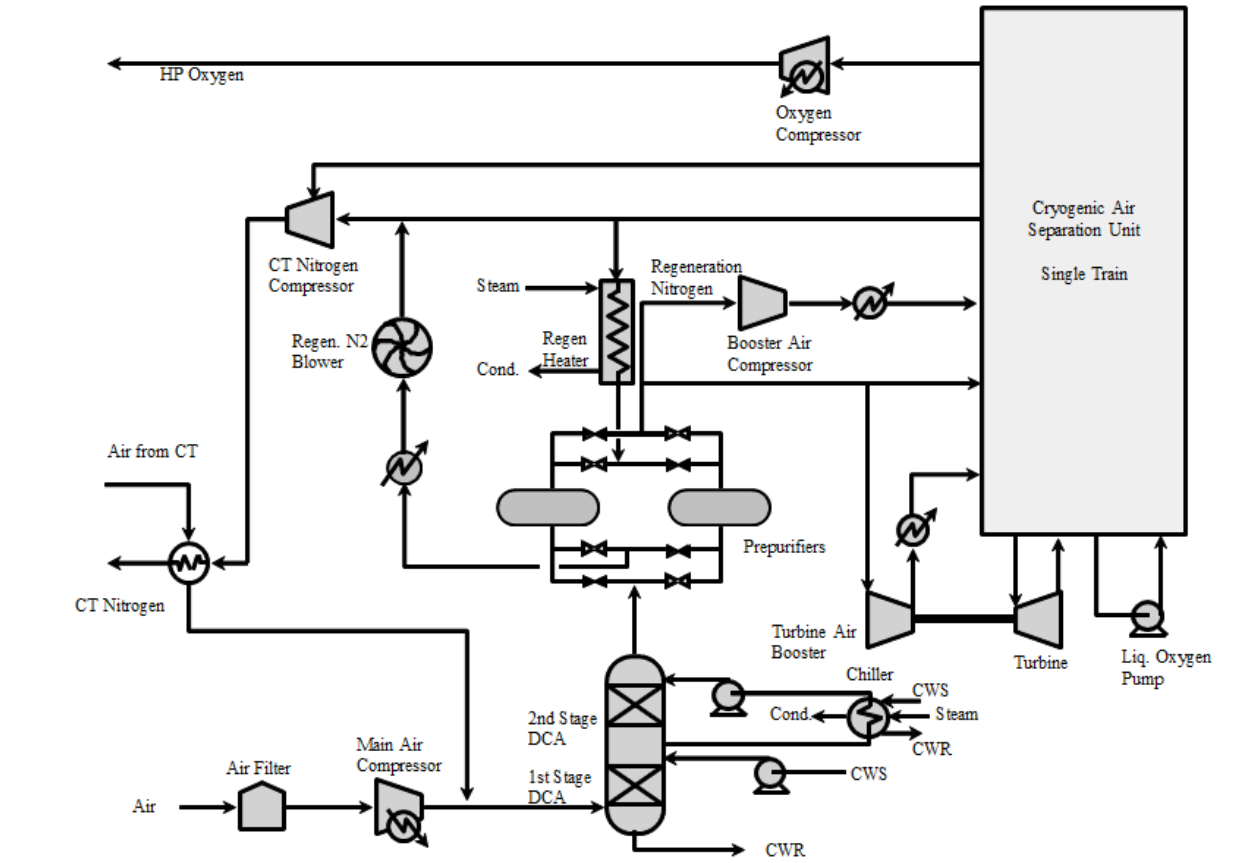
Exhibit 3-1 Air extracted from the combustion turbine and supplied to the ASU in non-carbon capture cases

Case No.	B1A	B4A	B5A
Gasifier	Shell	E-Gas™	GEE
Air Extracted from Combustion Turbine, %	4.0	4.0	4.0
Air Provided to ASU, % of ASU Total	19.4	18.9	16.8

3.1.2.5 Air Separation Plant Process Description

The air separation plant is designed to produce 95 percent by volume O₂ for use in the gasifier and Claus plant. The ASU is designed with two production trains, one for each gasifier. The air compressor is powered by an electric motor. Nitrogen, containing less than 2 percent by volume of O₂, is recovered, compressed, and used as a diluent in the CT combustor. A process schematic of a typical ASU is shown in Exhibit 3-2. (34)

The air feed to the ASU is supplied from two sources. A portion of the air is extracted from the compressor of the CT (non-CO₂ capture cases only). The remaining air is supplied from a stand-alone compressor. Air to the stand-alone compressor is first filtered in a suction filter upstream of the compressor. This air filter removes particulate, which may tend to cause compressor wheel erosion and foul intercoolers. The filtered air is then compressed in the centrifugal compressor, with intercooling and aftercooling.



Source: NETL

Air from the stand-alone compressor is combined with the extraction air, and the combined stream is cooled and fed to an adsorbent-based pre-purifier system. The adsorbent removes water, CO₂, and C₄+ saturated hydrocarbons in the air. After passing through the adsorption beds, the air is filtered with a dust filter to remove any adsorbent fines that may be present. Downstream of the dust filter a small stream of air is withdrawn to supply the instrument air requirements of the ASU.

Regeneration of the adsorbent in the pre-purifiers is accomplished by passing a hot nitrogen stream through the off-stream bed(s) in a direction counter current to the normal airflow. The nitrogen is heated against extraction steam (1.7 MPa [250 psia]) in a shell and tube heat exchanger. (34) The regeneration nitrogen drives off the adsorbed contaminants. Following regeneration, the heated bed is cooled to near normal operating temperature by passing a cool nitrogen stream through the adsorbent beds. The bed is re-pressurized with air and placed on stream so that the current on-stream bed(s) can be regenerated.

The air from the pre-purifier is then split into three streams. About 70 percent of the air is fed directly to the cold box. About 25 percent of the air is compressed in an air booster compressor. This boosted air is then cooled in an aftercooler against cooling water in the first stage and against chilled water in the second stage before it is fed to the cold box. The chiller utilizes low-pressure (LP) process steam at 0.3 MPa (50 psia) to drive the absorption refrigeration cycle. The

remaining five percent of the air is fed to a turbine-driven, single-stage, centrifugal booster compressor. This stream is cooled in a shell and tube aftercooler against cooling water before it is fed to the cold box. (34)

All three air feeds are cooled in the cold box to cryogenic temperatures against returning product oxygen and nitrogen streams in plate-and-fin heat exchangers. The large air stream is fed directly to the first distillation column to begin the separation process. The second largest air stream is liquefied against boiling liquid oxygen before it is fed to the distillation columns. The third, smallest air stream is fed to the cryogenic expander to produce refrigeration to sustain the cryogenic separation process.

Inside the cold box the air is separated into oxygen and nitrogen products. The oxygen product is withdrawn from the distillation columns as a liquid and is pressurized by a cryogenic pump. The pressurized liquid oxygen is then vaporized against the high-pressure (HP) air feed before being warmed to ambient temperature. The gaseous oxygen exits the cold box and is fed to the centrifugal compressor with intercooling between each stage of compression. The compressed oxygen is then fed to the gasification unit.

Nitrogen is produced from the cold box at two pressure levels. LP nitrogen is split into two streams. The majority of the LP nitrogen is compressed and fed to the CT as diluent nitrogen. A small portion of the nitrogen is used as the regeneration gas for the pre-purifiers and recombined with the diluent nitrogen. A HP nitrogen stream is also produced from the cold box and is further compressed before it is also supplied to the CT.

3.1.3 Water Gas Shift Reactors

Selection of Technology

In the cases with CO₂ separation and capture, the gasifier product must be converted to hydrogen-rich syngas. The first step is to convert most of the syngas CO to CO₂ by reacting the CO with water over a bed of catalyst, producing hydrogen (H₂) as a byproduct. The H₂O:CO molar ratio in the shift reaction, shown below, is adjusted to approximately 2:1 by the addition of steam to the syngas stream thus promoting a high conversion of CO. In the cases without CO₂ separation and capture, CO shift converters are not required.



The CO shift converter can be located either upstream of the AGR step (sour gas shift) or immediately downstream (sweet gas shift). If the CO converter is located downstream of the AGR, then the metallurgy of the unit is less stringent but additional equipment must be added to the process. This is because the CO converter promotes carbonyl sulfide (COS) hydrolysis without a separate catalyst bed. Products from the gasifier are humidified with steam or water and contain a portion of the water vapor necessary to meet the water-to-gas criteria at the reactor inlet. If the CO converter is located downstream of the AGR, then the gasifier product would first have to be cooled and the free water separated and treated. Then additional steam would have to be generated and re-injected into the CO converter feed to meet the required water-to-gas ratio. Therefore, for this report the CO converter was located upstream of the AGR unit and is referred to as SGS.

Process Description - The SGS consists of two paths of parallel fixed-bed reactors arranged in series. Two reactors in series are used in each parallel path to achieve sufficient conversion to

meet the 90 percent carbon capture target. In the E-GasTM case, a third shift reactor is added to each path to increase the CO conversion because of the relatively high amount of CH₄ present in the syngas. With the third reactor added, carbon capture is 89.9 percent in the E-GasTM case.

Cooling is provided between the series of reactors to control the exothermic temperature rise. The parallel set of reactors is required due to the high gas mass flow rate. In all three CO₂ capture cases the heat exchanger after the first SGS reactor is used to vaporize water that is then used to adjust the syngas H₂O:CO ratio to 2:1 on a molar basis. The heat exchanger after the second SGS reactor is used to raise intermediate-pressure (IP) steam, which then passes through the reheater (RH) section of the HRSG in the GEE and E-GasTM cases, and is used to preheat the syngas prior to the first SGS reactor in the Shell case. Approximately 97 percent conversion of the CO is achieved in the GEE and Shell cases, and about 98 percent conversion is achieved in the E-GasTM case.

3.1.4 Mercury Removal

An IGCC power plant has the potential of removing mercury in a more simple and cost-effective manner than conventional plants (e.g., PC). This is because mercury can be removed from the syngas at elevated pressure and prior to combustion where syngas volumes are much smaller than combusted flue gas volumes in conventional plants. A conceptual design for a sulfur-impregnated, activated carbon bed adsorption system was developed for mercury control. Data on the performance of carbon bed systems were obtained from the Eastman Chemical Company, which uses carbon beds at its syngas facility in Kingsport, Tennessee. (18) The coal mercury content (0.15 ppm dry) and carbon bed removal efficiency (greater than 99 percent) were discussed previously in Sections 2.2 and 2.3, respectively.

Carbon Bed Location – The packed carbon bed vessels are located upstream of the sulfur recovery unit (SRU) and syngas enters at a temperature near 38°C (100°F). Consideration was given to locating the beds further upstream before the COS hydrolysis unit (in non-CO₂ capture cases) at a temperature near 204°C (400°F). However, while the mercury removal efficiency of carbon has been found to be relatively insensitive to pressure variations, temperature adversely affects the removal efficiency. (35) Eastman Chemical also operates their beds ahead of their SRU at a temperature of 30°C (86°F). (18)

Consideration was also given to locating the beds downstream of the SRU. However, it was felt that removing the mercury and other contaminants before the SRU would enhance the performance of the SRU and increase the life of the various solvents.

Process Parameters – An empty vessel basis gas residence time of approximately 20 seconds was used based on Eastman Chemical's experience. (18) Allowable gas velocities are limited by considerations of particle entrainment, bed agitation, and pressure drop. One-foot-per-second superficial velocity is in the middle of the range normally encountered (35) and was selected for this application.

The bed density of 30 lb/ft³ was based on the Calgon Carbon Corporation HGR-P sulfur-impregnated pelletized activated carbon. (36) These parameters determined the size of the vessels and the amount of carbon required. Each gasifier train has two mercury removal beds and there are two gasifier trains in each case, resulting in four carbon beds per case.

Carbon Replacement Time – Eastman Chemicals replaces its bed every 18 to 24 months. (18) However, bed replacement is not because of mercury loading, but for other reasons including:

- A buildup in pressure drop
- A buildup of water in the bed
- A buildup of other contaminants

For this report a 24 month carbon replacement cycle was assumed. Under these assumptions, the mercury loading in the bed would build up to 0.6 - 1.1 weight percent (wt%). Mercury capacity of sulfur-impregnated carbon can be as high as 20 wt%. (37) The mercury laden carbon is considered to be a hazardous waste, and the disposal cost estimate reflects this categorization.

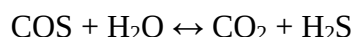
3.1.5 Acid Gas Removal (AGR) Process Selection

Gasification of coal to generate power produces a syngas that must be treated prior to further utilization. A portion of the treatment consists of AGR and sulfur recovery. This includes all sulfur species, but in particular the total of COS and H₂S, thereby resulting in stack gas emissions of less than 4 ppmv SO₂.

3.1.5.1 COS Hydrolysis

The use of COS hydrolysis pretreatment in the feed to the AGR process converts the COS to a more easily capturable H₂S species. This method was first commercially proven at the Buggenum plant, and was also used at both the Tampa Electric and Wabash River IGCC projects. Several catalyst manufacturers, including Haldor Topsoe and Porocel, offer a catalyst that promotes the COS hydrolysis reaction. The non-carbon capture COS hydrolysis reactor designs are based on information from Porocel. In cases with CO₂ capture, the SGS reactors reduce COS to H₂S as discussed in Section 3.1.3.

The COS hydrolysis reaction is equimolar with a slightly exothermic heat of reaction, as shown in the following reaction:



Since the reaction is exothermic, higher conversion is achieved at lower temperatures. However, at lower temperatures the reaction kinetics are slower. Based on the feed gas for this evaluation, Porocel recommended a temperature of 177 to 204°C (350 to 400°F). Since the exit gas COS concentration is critical to the amount of H₂S that must be removed with the AGR process, a retention time of 50-75 seconds was used to achieve 99.5 percent conversion of the COS. The Porocel activated alumina-based catalyst, designated as Hydrocel 640 catalyst, promotes the COS hydrolysis reaction without promoting reaction of H₂S and CO to form COS and H₂.

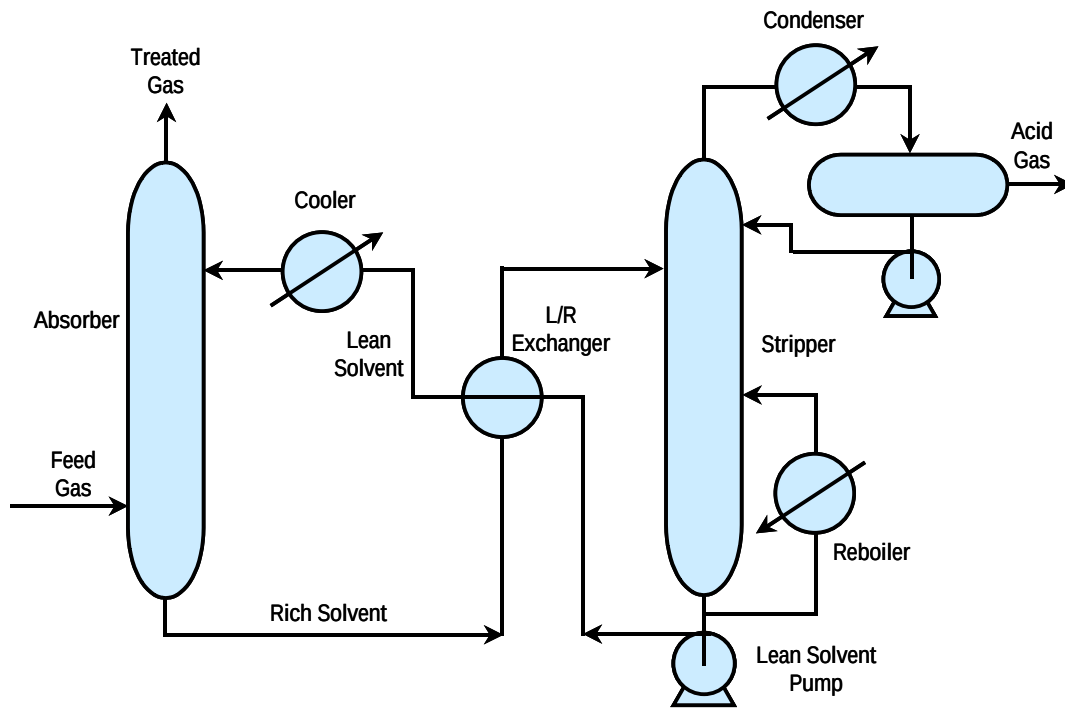
Although the reaction is exothermic, the heat of reaction is dissipated among the large amount of non-reacting components. Therefore, the reaction is essentially isothermal. The product gas, now containing less than 4 ppmv of COS, is cooled prior to entering the mercury removal process and the AGR.

3.1.5.2 Sulfur Removal

H₂S removal generally consists of absorption by a regenerable solvent. The most commonly used technique is based on counter current contact with the solvent. Acid-gas-rich solution from the absorber is stripped of its acid gas in a regenerator, usually by application of heat. The

regenerated lean solution is then cooled and recirculated to the top of the absorber, completing the cycle. Exhibit 3-3 is a simplified diagram of the AGR process. (38)

Exhibit 3-3 Flow diagram for a conventional AGR unit



Source: NETL

There are well over 30 AGR processes in common commercial use throughout the oil, chemical, and natural gas industries. However, in a 2002 report by SFA Pacific a list of 42 operating and planned gasifiers shows that only six AGR processes are represented: Rectisol, Sulfinol, MDEA, Selexol, aqueous di-isopropanol (ADIP) amine, and FLEXSORB. (39) These processes can be separated into three general types: chemical reagents, physical solvents, and hybrid solvents.

3.1.5.3 AGR/Gasifier Pairings used in Non-capture Cases

There are numerous commercial AGR processes that could meet the sulfur environmental target of this report. The most frequently used AGR systems (Selexol, Sulfinol, MDEA, and Rectisol) have all been used with the Shell and GEE gasifiers in various applications. Both existing E-GasTM gasifiers use MDEA, but could in theory use any of the existing AGR technologies. (38) The following selections were made for the AGR process in non-CO₂ capture cases:

- GEE gasifier: Selexol was chosen based on the GE gasifier operating at the highest pressure (815 psia versus 615 psia for E-GasTM and Shell), which favors the physical solvent used in the Selexol process.
- E-GasTM gasifier: Refrigerated MDEA was chosen because the two operating E-GasTM gasifiers use MDEA and because Chicago Bridge & Iron Company (CB&I) lists MDEA as the selected AGR process on their website. (39) Refrigerated MDEA was chosen over conventional MDEA because the sulfur emissions environmental target chosen is just outside of the range of conventional (higher temperature) MDEA.

- Shell gasifier: The Sulfinol process was chosen for this case because it is a Shell owned technology. While the Shell gasifier can and has been used with other AGR processes, it was concluded the most likely pairing would be with the Sulfinol process.

The two-stage Selexol process is used in all three cases that require CO₂ capture. According to the previously referenced SFA Pacific report, “For future IGCC with CO₂ removal for sequestration, a two-stage Selexol process presently appears to be the preferred AGR process – as indicated by ongoing engineering studies at EPRI and by various engineering firms with IGCC interests.” (40)

3.1.5.4 AGR used in CO₂ Capture Cases

This section provides a brief process description of the two-stage selexol process used for capture cases.

Syngas enters the first of two absorbers where H₂S is preferentially removed using loaded solvent from the CO₂ absorber. The gas exiting the H₂S absorber passes through the second absorber where CO₂ is removed using first flash regenerated, chilled solvent followed by thermally regenerated solvent added near the top of the column. The treated gas exits the absorber and is sent either directly to the CT or is partially humidified prior to entering the CT. A portion of the gas can also be used for coal drying, when required.

The amount of hydrogen recovered from the syngas stream is dependent on the Selexol process design conditions. In this report, hydrogen recovery is 99.4 percent. The minimal hydrogen slip to the CO₂ sequestration stream maximizes the overall plant efficiency. The Selexol plant cost estimates are based on a plant designed to recover this high percentage of hydrogen.

The CO₂ loaded solvent exits the CO₂ absorber. A portion is sent to the H₂S absorber, a portion is sent to a reabsorber, and the remainder is sent to a series of flash drums for regeneration. The CO₂ product stream is obtained from the three flash drums. After flash regeneration, the solvent is chilled and returned to the CO₂ absorber.

The rich solvent exiting the H₂S absorber is combined with the rich solvent from the reabsorber, and the combined stream is heated using the lean solvent from the stripper. The hot, rich solvent enters the H₂S concentrator and partially flashes. The remaining liquid contacts nitrogen from the ASU, and a portion of the CO₂ along with lesser amounts of H₂S and COS are stripped from the rich solvent. The stripped gases from the H₂S concentrator are sent to the reabsorber where the H₂S and COS that were co-stripped in the concentrator are transferred to a stream of loaded solvent from the CO₂ absorber. The clean gas from the reabsorber is combined with the clean gas from the H₂S absorber and sent to the CT.

The solvent exiting the H₂S concentrator is sent to the stripper where the absorbed gases are liberated by hot gases flowing up the column from the steam heated reboiler. Water in the overhead vapor from the stripper is condensed and returned as reflux to the stripper or exported as necessary to maintain the proper water content of the lean solvent. The acid gas from the stripper is sent to the Claus plant for further processing. The lean solvent exiting the stripper is first cooled by providing heat to the rich solvent, then further cooled by exchange with the product gas and finally chilled in the lean chiller before returning to the top of the CO₂ absorber.

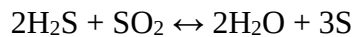
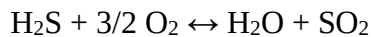
3.1.6 Sulfur Recovery/Tail Gas Cleanup Process Selection

Currently, most of the world's sulfur is produced from the acid gases coming from gas treating. The Claus process remains the mainstay for sulfur recovery. Conventional three-stage Claus plants, with indirect reheat and feeds with a high H₂S content, can approach 98 percent sulfur recovery efficiency. However, since environmental regulations have become more stringent, sulfur recovery plants are required to recover sulfur with over 99.8 percent efficiency. To meet these stricter regulations, the Claus process underwent various modifications and add-ons.

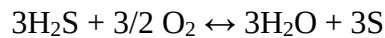
The add-on modification to the Claus plant selected for this report can be considered a separate option from the Claus process. In this context, it is often called a tail gas treating unit (TGTU) process.

The Claus Process

The Claus process converts H₂S to elemental sulfur via the following reactions:



The second reaction, the Claus reaction, is equilibrium limited. The overall reaction is:

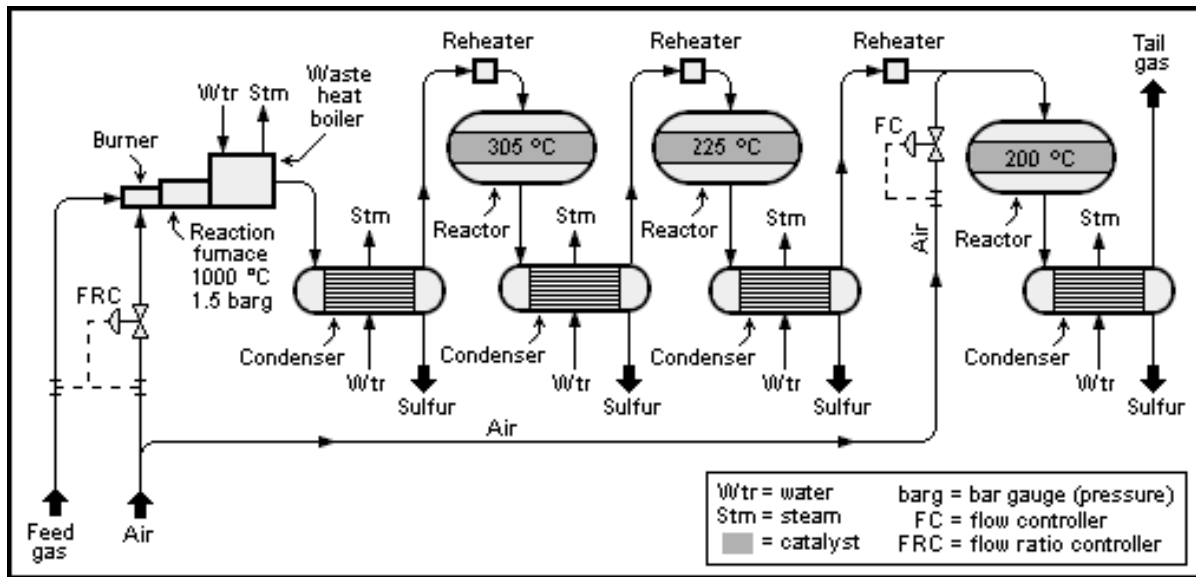


The sulfur in the vapor phase exists as S₂, S₆, and S₈ molecular species, with the S₂ predominant at higher temperatures, and S₈ predominant at lower temperatures.

A simplified process flow diagram of a typical three-stage Claus plant is shown in Exhibit 3-4. (40) One-third of the H₂S is burned in the furnace with oxygen from the air to give sufficient SO₂ to react with the remaining H₂S. Since these reactions are highly exothermic, a waste heat boiler that recovers this heat to generate HP steam usually follows the furnace. Sulfur is condensed in a condenser that follows the HP steam recovery section. LP steam is raised in the condenser. The tail gas from the first condenser then goes to several catalytic conversion stages, usually two to three, where the remaining sulfur is recovered via the Claus reaction. Each catalytic stage consists of gas preheat, a catalytic reactor, and a sulfur condenser. The liquid sulfur goes to the sulfur pit, while the tail gas proceeds to the incinerator or for further processing in a TGTU.

3.1.6.1 Claus Plant Sulfur Recovery Efficiency

The Claus reaction is equilibrium limited, and sulfur conversion is sensitive to the reaction temperature. The highest sulfur conversion in the thermal zone is limited to about 75 percent. Typical furnace temperatures are in the range of 1093 to 1427°C (2000 to 2600°F), and as the temperature decreases, conversion increases dramatically.

Exhibit 3-4 Typical three-stage Claus sulfur plant

Source: Milton Beychok (41)

Claus plant sulfur recovery efficiency depends on many factors:

- H_2S concentration of the feed gas
- Number of catalytic stages
- Gas reheat method

In order to keep Claus plant recovery efficiencies approaching 94 to 96 percent for feed gases that contain about 20 to 50 percent H_2S , a split-flow design is often used. In this version of the Claus plant, part of the feed gas is bypassed around the furnace to the first catalytic stage, while the rest of the gas is oxidized in the furnace to mostly SO_2 . This results in a more stable temperature in the furnace.

3.1.6.2 Oxygen-Blown Claus

Large diluent streams in the feed to the Claus plant, such as N_2 from combustion air, or a high CO_2 content in the feed gas, lead to higher cost Claus processes and any add-on or tail gas units. One way to reduce diluent flows through the Claus plant and to obtain stable temperatures in the furnace for dilute H_2S streams is the oxygen-blown Claus process.

The oxygen-blown Claus process was originally developed to increase capacity at existing conventional Claus plants and to increase flame temperatures of low H_2S content gases. The process has also been used to provide the capacity and operating flexibility for sulfur plants where the feed gas is variable in flow and composition such as often found in refineries. The application of the process has now been extended to grass roots installations, even for rich H_2S feed streams, to provide operating flexibility at lower costs than would be the case for conventional Claus units. At least four of the gasification plants in Europe use oxygen enriched Claus units.

Oxygen enrichment results in higher temperatures in the front-end furnace, potentially reaching temperatures as high as 1593 to 1649°C (2900 to 3000°F) as the enrichment moves beyond 40 to 70 vol% O_2 in the oxidant feed stream. Although oxygen enrichment has many benefits, its

primary benefit for lean H₂S feeds is a stable furnace temperature. Sulfur recovery is not significantly enhanced by oxygen enrichment. Because the IGCC process already requires an ASU, the oxygen-blown Claus plant was chosen for all cases.

3.1.6.3 Tail Gas Treating

In many refinery and other conventional Claus applications, tail gas treating involves the removal of the remaining sulfur compounds from gases exiting the SRU. Tail gas from a typical Claus process, whether a conventional Claus or one of the extended versions of the process, usually contains small but varying quantities of COS, CS₂, H₂S, SO₂, and elemental sulfur vapors. In addition, there may be H₂, CO, and CO₂ in the tail gas. In order to remove the rest of the sulfur compounds from the tail gas, all of the sulfur-bearing species must first be converted to H₂S. Then, the resulting H₂S is absorbed into a solvent and the clean gas vented or recycled for further processing. The clean gas resulting from the hydrolysis step can undergo further cleanup in a dedicated absorption unit or be integrated with an upstream AGR unit. The latter option is particularly suitable with physical absorption solvents. The approach of treating the tail gas in a dedicated amine absorption unit and recycling the resulting acid gas to the Claus plant is the one used by the Shell Claus Off-gas Treating (SCOT) process. With tail gas treatment, Claus plants can achieve overall removal efficiencies in excess of 99.9 percent.

In the case of IGCC applications, the tail gas from the Claus plant can be catalytically hydrogenated and then recycled back into the system with the choice of location being technology dependent, or it can be treated with a SCOT-type process. In each of the seven cases the Claus plant tail gas is hydrogenated, water is separated, the tail gas is compressed and then returned to the AGR process for further treatment.

3.1.6.4 Flare Stack

A self-supporting, refractory-lined, carbon steel (CS) flare stack is typically provided to combust and dispose of unreacted gas during startup, shutdown, and upset conditions. However, in all seven cases a flare stack was provided for syngas dumping during startup, shutdown, etc. This flare stack eliminates the need for a separate Claus plant flare.

3.1.7 Slag Handling

The slag handling system conveys, stores, and disposes of slag removed from the gasification process. Spent material drains from the gasifier bed into a water bath in the bottom of the gasifier vessel. A slag crusher receives slag from the water bath and grinds the material into pea-sized fragments. A slag/water slurry that is between 5 and 10 percent solids leaves the gasifier pressure boundary through either a proprietary pressure letdown device (E-GasTM) or through the use of lockhoppers (GEE and Shell) to a series of dewatering bins.

The general aspects of slag handling are the same for all three technologies. The slag is dewatered, the water is clarified and recycled, and the dried slag is transferred to a storage area for disposal. The specifics of slag handling vary among the gasification technologies regarding how the water is separated and the end uses of the water recycle streams.

In this report the slag bins were sized for a nominal holdup capacity of 72 hours of full-load operation. At periodic intervals, a convoy of slag-hauling trucks will transit the unloading station underneath the hopper and remove a quantity of slag for disposal. Approximately ten truckloads per day are required to remove the total quantity of slag produced by the plant

operating at nominal rated power. While the slag is suitable for use as a component of road paving mixtures, it was assumed in this report that the slag would be landfilled at a specified cost.

3.1.8 Power Island

3.1.8.1 Combustion Turbine

The CT generator selected for this application is representative of the state-of-the-art 2008 F-class turbines. This machine is an axial flow, single spool, and constant speed unit, with variable inlet guide vane (IGV). The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature than the previous generation machines. The standard production version of this machine is fired with natural gas and is also commercially offered for use with IGCC derived syngas, although only earlier versions of the turbine are currently operating on syngas. For the purposes of this report, it was assumed that the state-of-the-art 2008 F-class turbine will be commercially available for use on both conventional and high hydrogen content syngas representative of the cases with CO₂ capture. High H₂ fuel combustion issues like flame stability, flashback, and NO_x formation were assumed to be solved in the time frame needed to support deployment. However, because these are FOAK applications, process contingencies were included in the cost estimates as described in Section 2.6. Performance typical of a state-of-the-art 2008 F-class turbine on natural gas at ISO conditions is presented in Exhibit 3-5.

Exhibit 3-5 State-of-the-art 2008 F-class combustion turbine performance characteristics using natural gas (42)

	State-of-the-art 2008 F-Class
Firing Temperature Class, °C (°F)	1,371+ (2,500+)
Airflow, kg/s (lb/s)	431 (950)
Pressure Ratio	18.5
Simple Cycle Output, MW	185
Combined cycle performance	
Net Output, MW	280
Net Efficiency (LHV), %	57.5
Net Heat Rate (LHV), kJ/kWh (Btu/kWh)	6,256 (5,934)

In this service, with syngas from an IGCC plant, the machine requires some modifications to the burner and turbine nozzles in order to properly combust the low-Btu gas and expand the combustion products in the turbine section of the machine.

The modifications to the machine include a redesign of the original can-annular combustors. A second modification involves increasing the nozzle areas of the turbine to accommodate the mass and volume flow of low-Btu fuel gas combustion products, which are increased relative to those produced when firing natural gas. Other modifications include rearranging the various auxiliary skids that support the machine to accommodate the spatial requirements of the plant's general arrangement. The generator is a standard hydrogen-cooled machine with static exciter.

3.1.8.2 Combustion Turbine Package Scope of Supply

The CT is typically supplied in several fully shop-fabricated modules, complete with all mechanical, electrical, and control systems as required for CT operation. Site CT installation involves module inter-connection, and linking CT modules to the plant systems.

3.1.8.3 CT Firing Temperature Control Issue for Low Calorific Value Fuel

A CT when fired on LCV syngas has the potential to increase power output due to the increase in flow rate through the turbine. The higher turbine flow and moisture content of the combustion products can contribute to overheating of turbine components, affect rating criteria for the parts lives, and require a reduction in syngas firing temperatures (compared to the natural gas firing) to maintain design metal temperature. (43) Uncontrolled syngas firing temperature could result in more than 50 percent life cycle reduction of stage 1 buckets. Control systems for syngas applications include provisions to compensate for these effects by maintaining virtually constant generation output for the range of the specified ambient conditions. IGVs and firing temperature are used to maintain the turbine output at the maximum torque rating, producing a flat rating up to the IGV full open position. Beyond the IGV full open position, flat output may be extended to higher ambient air temperatures by steam/nitrogen injection.

In this report the firing temperature (defined as inlet rotor temperature) in the non-capture cases is 1333-1343°C (2432-2449°F) and in the CO₂ capture cases is 1317-1322°C (2402-2412°F). The reduction in firing temperature in the CO₂ capture cases is done to maintain parts life as the H₂O content of the combustion products increases from 6-9 vol% in the non-capture cases to 12-14 vol% in the capture cases. The decrease in temperature also results in the lower temperature steam cycle in the CO₂ capture cases, ranging from 12.4 MPa/534°C/534°C (1800 psig/993°F/993°F) to 12.4 MPa/534°C/534°C (1800 psig/994°F/994°F) for all of the CO₂ capture cases versus 12.4 MPa/559°C/559°C (1800 psig/1038°F/1038°F) to 12.4 MPa/562°C/562°C (1800 psig/1043°F/1043°F) for all of the non-CO₂ capture cases.

3.1.8.4 Combustion Turbine Syngas Fuel Requirements

Typical fuel specifications and contaminant levels for successful CT operation are provided in GE Power System's "Specification for Fuel Gases for Combustion in Heavy-Duty Turbines" and presented for state-of-the-art 2008 F-class machines in Exhibit 3-6 and Exhibit 3-7. (44) The vast majority of published CT performance information is specific to natural gas operation. Turbine performance using syngas requires vendor input as was obtained for this report.

Exhibit 3-6 Typical fuel specification for state-of-the-art 2008 F-class machines

	Max	Min
LHV, kJ/m ³ (Btu/scf)	None	3.0 (100)
Gas Fuel Pressure, MPa (psia)	3.1 (450)	
Gas Fuel Temperature, °C (°F)	A	Varies with gas pressure ^B
Flammability Limit Ratio, Rich-to-Lean, Volume Basis	C	2:2.1
Sulfur	D	

^A The maximum fuel temperature is defined in GEK-4189 (45)

^B To ensure that the fuel gas supply to the CT is 100 percent free of liquids the minimum fuel gas temperature must meet the required superheat over the respective dew point. This requirement is independent of the hydrocarbon and moisture concentration. Superheat calculation shall be performed as described in GEI-4140G (44)

^C Maximum flammability ratio limit is not defined. Fuel with flammability ratio significantly larger than those of natural gas may require start-up fuel

^D The quantity of sulfur in syngas is not limited by specification. Experience has shown that fuel sulfur levels up to one percent by volume do not significantly affect oxidation/corrosion rates

3.1.8.5 Normal Operation

Inlet air is compressed in a single spool compressor to a pressure ratio of approximately 16:1. This pressure ratio was vendor specified and less than the 17:1 ratio used in natural gas applications. The majority of compressor discharge air remains on board the machine and passes to the burner section to support combustion of the syngas. Compressed air is also used in burner, transition, and film cooling services. About 4-7 percent of the compressor air is extracted and integrated with the air supply of the ASU in non-carbon capture cases. It may be technically possible to integrate the CT and ASU in CO₂ capture cases as well; however, in this report integration was considered only for non-carbon capture cases.

Exhibit 3-7 Allowable gas fuel contaminant level for state-of-the-art 2008 F-class machines

	Turbine Inlet Limit, ppbw	Fuel Limit, ppmw		
		Turbine Inlet Flow/Fuel Flow		
		50	12	4
Lead	20	1.0	0.240	.080
Vanadium	10	0.5	0.120	0.040
Calcium	40	2.0	0.480	0.160
Magnesium	40	2.0	0.480	0.160
Sodium + Potassium				
Na/K = 28 ^A	20	1.0	0.240	0.080
Na/K = 3	10	0.5	0.120	0.40
Na/K ≤ 1	6	0.3	0.072	0.024
Particulates Total ^B	600	30	7.2	2.4
Above 10 microns	6	0.3	0.072	0.024

^ANominal sea salt ratio

^BThe fuel gas delivery system shall be designed to prevent generation or admittance of solid particulate to the CT gas fuel system

Pressurized syngas is combusted in 14 parallel diffusion combustors, and syngas dilution is used to limit NO_x formation. As described in Section 3.1.2, nitrogen from the ASU is used as the primary diluent followed by syngas humidification and finally by steam dilution, if necessary, to achieve an LHV of 4.4-4.7 MJ/Nm³ (119-125 Btu/scf). The advantages of using nitrogen as the primary diluent include:

- Nitrogen from the ASU is already partially compressed and using it for dilution eliminates wasting the compression energy.

- Limiting the water content reduces the need to de-rate firing temperature, particularly in the high-hydrogen (CO₂ capture) cases.

There are some disadvantages to using nitrogen as the primary diluent, and these include:

- There is a significant auxiliary power requirement to further compress the large nitrogen flow from the ASU pressures of 0.4 and 1.3 MPa (56 and 182 psia) to the CT pressure of 3.2 MPa (465 psia).
- Low-quality heat not otherwise useful for other applications can be used to preheat water for the syngas humidification process.
- Nitrogen is not as efficient as water in limiting NO_x emissions.

It is not clear that one dilution method provides a significant advantage over the other. However, in this report nitrogen was chosen as the primary diluent based on suggestions by turbine industry experts during peer review of the report.

Hot combustion products are expanded in the three-stage turbine-expander. Given the assumed ambient conditions, back-end loss, and HRSG pressure drop, the CT exhaust temperature is nominally 588°C (1,090°F) for non-CO₂ capture cases and 562°C (1,044°F) for capture cases.

Gross turbine power, as measured prior to the generator terminals, is 232 MW per train. The CT generator is a standard hydrogen-cooled machine with static exciter.

3.1.9 Steam Generation Island

3.1.9.1 Heat Recovery Steam Generator

The HRSG is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the CT exhaust gas. High-temperature flue gas exiting the CT is conveyed through the HRSG to recover the large quantity of thermal energy that remains. Flue gas travels through the HRSG gas path and exits at 132°C (270°F) for all 7 cases.

The HP drum produces steam at main steam pressure, while the IP drum produces process steam and turbine dilution steam, if required. The HRSG drum pressures are nominally 12.4/3.1 MPa (1,800/443 psia) for the HP/IP turbine sections, respectively. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater (FW) heating, and also provides deaeration of the condensate.

Natural circulation of steam is accomplished in the HRSG by utilizing differences in densities due to temperature differences of the steam. The natural circulation HRSG provides the most cost-effective and reliable design.

The HRSG drums include moisture separators, internal baffles, and piping for FW/steam. All tubes, including economizers, superheaters, and headers and drums, are equipped with drains.

Safety relief valves are furnished in order to comply with appropriate codes and ensure a safe work place.

Superheater, boiler, and economizer sections are supported by shop-assembled structural steel. Inlet and outlet duct is provided to route the gases from the CT outlet to the HRSG inlet and the HRSG outlet to the stack. A diverter valve is included in the inlet duct to bypass the gas when appropriate. Suitable expansion joints are also included.

3.1.9.2 Steam Turbine Generator and Auxiliaries

The steam turbine consists of an HP section, an IP section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing. The LP turbine has a last stage bucket length of 76 cm (30 in).

Main steam from the HRSG and gasifier island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at either 12.4 MPa/559°C to 562°C (1,800 psig/1,038°F to 1,043°F) for the non-carbon capture cases, or 12.4 MPa/534°C (1,800 psig/993°F to 994°F) for the carbon capture cases. The steam initially enters the turbine near the middle of the HP span, flows through the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 3.1 MPa/558°C to 561°C (443 psig/1,036°F to 1,041°F) for the non-carbon capture cases or 3.1 MPa/532°C to 533°C (443 psig/990°F to 992°F) for the carbon capture cases. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

Turbine bearings are lubricated by a closed-loop (CL), water-cooled, pressurized oil system. The oil is contained in a reservoir located below the turbine floor. During startup or unit trip, an emergency oil pump mounted on the reservoir pumps the oil. When the turbine reaches 95 percent of synchronous speed, the main pump mounted on the turbine shaft pumps oil. The oil flows through water-cooled heat exchangers prior to entering the bearings. The oil then flows through the bearings and returns by gravity to the lube oil reservoir.

Turbine shafts are sealed against air in-leakage or steam blowout using a modern positive pressure variable clearance shaft sealing design arrangement connected to an LP steam seal system. During startup, seal steam is provided from the main steam line. As the unit increases load, HP turbine gland leakage provides the seal steam. Pressure-regulating valves control the gland header pressure and dump any excess steam to the condenser. A steam packing exhauster maintains a vacuum at the outer gland seals to prevent leakage of steam into the turbine room. Any steam collected is condensed in the packing exhauster and returned to the condensate system.

The generator is a hydrogen-cooled synchronous type, generating power at 24 kV. A static, transformer type exciter is provided. The generator is cooled with a hydrogen gas recirculation system using fans mounted on the generator rotor shaft. The heat absorbed by the gas is removed as it passes over finned tube gas coolers mounted in the stator frame. Gas is prevented from escaping at the rotor shafts by a CL oil seal system. The oil seal system consists of storage tank, pumps, filters, and pressure controls, all skid-mounted.

The STG is controlled by a triple-redundant, microprocessor-based electro-hydraulic control system. The system provides digital control of the unit in accordance with programmed control algorithms, color cathode ray tube (CRT) operator interfacing, and datalink interfaces to the balance-of-plant DCS, and incorporates on-line repair capability.

3.1.9.3 Condensate System

The condensate system transfers condensate from the condenser hotwell to the deaerator, through the gland steam condenser, gasifier, and the low-temperature economizer section in the HRSG.

The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the HRSG. Condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

3.1.9.4 Feedwater System

The function of the FW system is to pump the various FW streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided for each of 3 pressure levels, HP, IP, and LP. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

The FW pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high-bearing temperature. FW pump suction pressure and temperature are also monitored. In addition, the suction of each boiler feed pump is equipped with a startup strainer.

3.1.9.5 Main and Reheat Steam Systems

The function of the main steam system is to convey main steam generated in the synthesis gas cooler (SGC) and HRSG from the HRSG superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the HRSG RH, and to the turbine reheat stop valves.

Main steam at approximately 12.4 MPa/559°C to 562°C (1,800 psig/1,038°F to 1,043°F) (non-CO₂ capture cases) or 12.4 MPa/534°C (1,800 psig/993°F to 994°F) (CO₂ capture cases) exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to the HP turbine. Cold reheat steam at approximately 3.5 MPa/349°C to 372°C (501 psia/661°F to 702°F) exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 3.1 MPa/558 to 561°C (443 psig/1,036°F to 1,041°F) for the non-carbon capture cases and 3.1 MPa/532°C to 533°C (443 psig/990°F to 992°F) for the CO₂ capture cases exits the HRSG RH through a motor-operated gate valve and is routed to the IP turbines.

Steam piping is sloped from the HRSG to the drip pots located near the steam turbine for removal of condensate from the steam lines. Condensate collected in the drip pots and in low-point drains is discharged to the condenser through the drain system.

Steam flow is measured by means of flow nozzles in the steam piping. The flow nozzles are located upstream of any branch connections on the main headers.

Safety valves are installed to comply with appropriate codes and to ensure the safety of personnel and equipment.

3.1.9.6 Circulating Water System

The circulating water system (CWS) is a closed-cycle cooling water system that supplies cooling water to the condenser to condense the main turbine exhaust steam. The system also supplies cooling water to the AGR plant as required, and to the auxiliary cooling system. The auxiliary

cooling system is a CL process that utilizes a higher quality water to remove heat from compressor intercoolers, oil coolers, and other ancillary equipment and transfers that heat to the main circulating cooling water system in plate and frame heat exchangers. The heat transferred to the circulating water in the condenser and other applications is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical CWPs, a mechanical draft evaporative cooling tower, and CS cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counterflow mechanical draft cooling tower.

The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or for plugging tubes. This can be done during normal operation at reduced load.

The condenser is equipped with an air extraction system to evacuate the condenser steam space for removal of non-condensable gases during steam turbine operation and to rapidly reduce the condenser pressure from atmospheric pressure before unit startup and admission of steam to the condenser.

3.1.9.7 Raw Water, Fire Protection, and Cycle Makeup Water Systems

The raw water system supplies cooling tower makeup, cycle makeup, service water and potable water requirements. The water source is 50 percent from a POTW and 50 percent from groundwater. Booster pumps within the plant boundary provide the necessary pressure.

The fire protection system provides water under pressure to the fire hydrants, hose stations, and fixed water suppression system within the buildings and structures. The system consists of pumps, underground and aboveground supply piping, distribution piping, hydrants, hose stations, spray systems, and deluge spray systems. One motor-operated booster pump is supplied on the intake structure of the cooling tower with a diesel engine backup pump installed on the water inlet line.

The cycle makeup water system provides high-quality demineralized water for makeup to the HRSG cycle, for steam injection ahead of the SGS reactors in CO₂ capture cases, and for injection steam to the auxiliary boiler for control of NO_x emissions, if required.

The cycle makeup system consists of two 100 percent trains, each with a full-capacity activated carbon filter, primary cation exchanger, primary anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. The equipment is skid-mounted and includes a control panel and associated piping, valves, and instrumentation.

3.1.10 Accessory Electric Plant

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

3.1.11 Waste Treatment/Miscellaneous Systems

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash. It is anticipated that the treated water will be suitable for discharge into existing systems and be within the EPA standards for suspended solids, oil and grease, pH, and miscellaneous metals.

The waste treatment system is minimal and consists, primarily, of neutralization and oil/water separators (along with the associated pumps, piping, etc.).

A natural gas supply line has been included in all cases for start-up or emergency fuel.

Miscellaneous systems consisting of natural gas, service air, instrument air, and service water are provided. A natural gas supply line has been included in all cases for start-up or emergency fuel. All truck roadways and unloading stations inside the fence area are provided.

3.1.12 Instrumentation and Control

An integrated plant-wide distributed control system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally DCS. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to be operational and accessible 99.5 percent of the time that it is required (99.5 percent availability). The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are manually implemented, with operator selection of modular automation routines available. The exception to this, and an important facet of the control system for gasification, is the critical controller system, which is a part of the license package from the gasifier supplier and is a dedicated and distinct hardware segment of the DCS.

This critical controller system is used to control the gasification process. The partial oxidation of the fuel feed and oxygen feed streams to form a syngas product is a stoichiometric, temperature- and pressure-dependent reaction. The critical controller utilizes a redundant microprocessor executing calculations and dynamic controls at 100- to 200-millisecond intervals. The enhanced execution speeds as well as evolved predictive controls allow the critical controller to mitigate process upsets and maintain the reactor operation within a stable set of operating parameters.

3.1.13 Performance Summary Metrics

This section details the methodologies of several metrics reported in the performance summaries.

Cold Gas Efficiency

The cold gas efficiency is calculated by subtracting the heating value of any gas recycled to the gasifier from the heating value of the gas exiting the gasifier and dividing that difference by the thermal input to the gasifier. This calculation is represented by the equation:

$$CGE = \frac{(GO - RI)}{TI}$$

Where:

CGE – cold gas efficiency

GO – heating value of gas exiting the gasifier

RI – heating value of gas recycled to gasifier

TI – thermal input to into the gasifier

The thermal input to the gasifier is calculated by taking the coal feed rate and multiplying by the heating value and converting the units to kW.

The components considered for the heating value of the gasifier exit gas and recycle gas are CO, H₂, CH₄, H₂S, NH₃, and COS. Their mole fraction is extracted from the model, multiplied by the molar flow of the stream, and then multiplied by their individual heating values.

Combustion Turbine Efficiency

The combustion turbine efficiency is calculated by taking the combustion turbine power produced and dividing it by the thermal input to the turbines. This calculation is represented by the equation:

$$CTE = \frac{CTP}{TI}$$

Where:

CTE – combustion turbine efficiency

CTP – combustion turbine power

TI – thermal input to turbines

The thermal input is calculated by taking the mole fraction of the individual gases (CO, H₂, CH₄, H₂S, NH₃, and COS) multiplied by the molar flow rate of the total stream entering the combustion turbine and then multiplying each by their individual heating values.

Steam Turbine Efficiency

The steam turbine efficiency is calculated by taking the steam turbine power produced and dividing it by the difference between the thermal input and thermal output. This calculation is represented by the equation:

$$STE = \frac{STP}{(TI - TO)}$$

Where:

STE – steam turbine efficiency

STP – steam turbine power

TI – thermal input

TO – thermal output

The thermal input is considered to be the main steam, the makeup water, and the low pressure steam from the BFW.

The thermal output is considered to be any extractions such as the gasifier steam, the 250 psia header, and the BFW from the condenser.

The IP blowdown, HP blowdown, and superheater losses are also credited to the thermal input as they are extracted from the cycle prior to the main steam but after the condensate BFW.

Steam Turbine Heat Rate

The steam turbine heat rate is calculated by taking the inverse of the steam turbine efficiency. This calculation is represented by the equation:

$$STHR = \frac{1}{STE} * 3,412$$

Where:

STHR – steam turbine heat rate, Btu/kWh

STE – steam turbine efficiency, fraction

3.2 Shell Global Solutions IGCC Cases

This section contains an evaluation of plant designs for cases B1A and B1B, which are based on the Shell gasifier. Cases B1A and B1B are very similar in terms of process, equipment, scope and arrangement, except that Case B1B employs a syngas quench and includes SGS reactors, CO₂ absorption/regeneration, and compression/transport systems. There are no provisions for CO₂ removal in Case B1A.

The balance of this section is organized as follows:

- Gasifier Background – provides information on the development and status of the Shell gasification technology.
- Process and System Description – provides an overview of the technology operation as applied to Case B1A. The systems that are common to all gasifiers were covered in Section 3.1 and only features that are unique to Case B1A are discussed further in this section.
- Key Assumptions – provides a summary of study and modeling assumptions relevant to cases B1A and B1B.
- Sparing Philosophy – provided for cases B1A and B1B
- Performance Results – provides the main modeling results from Case B1A, including the performance summary, environmental performance, carbon balance, sulfur balance, water balance, mass and energy balance diagrams, and mass and energy balance tables.
- Equipment List – provides an itemized list of major equipment for Case B1A with account codes that correspond to the cost accounts in the Cost Estimates section.
- Cost Estimates – provides a summary of capital and operating costs for Case B1A.

Process and System Description, Performance Results, Equipment List and Cost Estimates are repeated for Case B1B.

3.2.1 Gasifier Background

The “Coal Gasification Guidebook: Status, Application, and Technologies” report published by the EPRI provides a detailed history of the development of several types of gasifier technology, including the Shell gasifier, as well as gasifier capacity, distinguishing characteristics, and important coal characteristics. (46)

As of 2009, Shell reported ten gasifiers in operation producing 100,000-150,000 Nm³/hr and three of the same size in construction. Another three ranging from 150,000-250,000 Nm³/hr are also in construction. (47) The large gasifier operating in The Netherlands has a bituminous coal-handling capacity of 1,633 tonnes/day (1,800 tpd) and produces dry gas at a rate of 158,575 Nm³/hr (5.6 million scf/hr) with an energy content of about 1,792 GJ/hr (1,700 MMBtu/hr) (HHV). This gasifier was sized to match the fuel gas requirements for the Siemens/Kraftwerk Union V-94.2 CT and could easily be scaled up to match state-of-the-art 2008 F-class turbine requirements. (47)

Shell gasifiers are capable of utilizing a wide variety of coal types, and compared to slurry fed gasifiers, the dry-fed, cooled-refractory lined, Shell gasifier has a lower oxygen requirement and produces a gas with a higher H_2S/CO_2 ratio, which improves sulfur recovery. (46)

While the use of dry feed allows for lower oxygen consumption, the feed system – which includes the coal drying system – is more costly and complicated. (46)

Coal characteristics that are favorable when selecting a coal type for use with a Shell gasifier include low concentrations of ash with a moderate fusion temperature. If a coal is selected that has a high ash fusion temperature, flux addition may be necessary. The negative impact that high ash coals have on the operation of gasifiers are reduced in dry feed systems in comparison to slurry fed gasifiers. (46)

3.2.2 Process Description

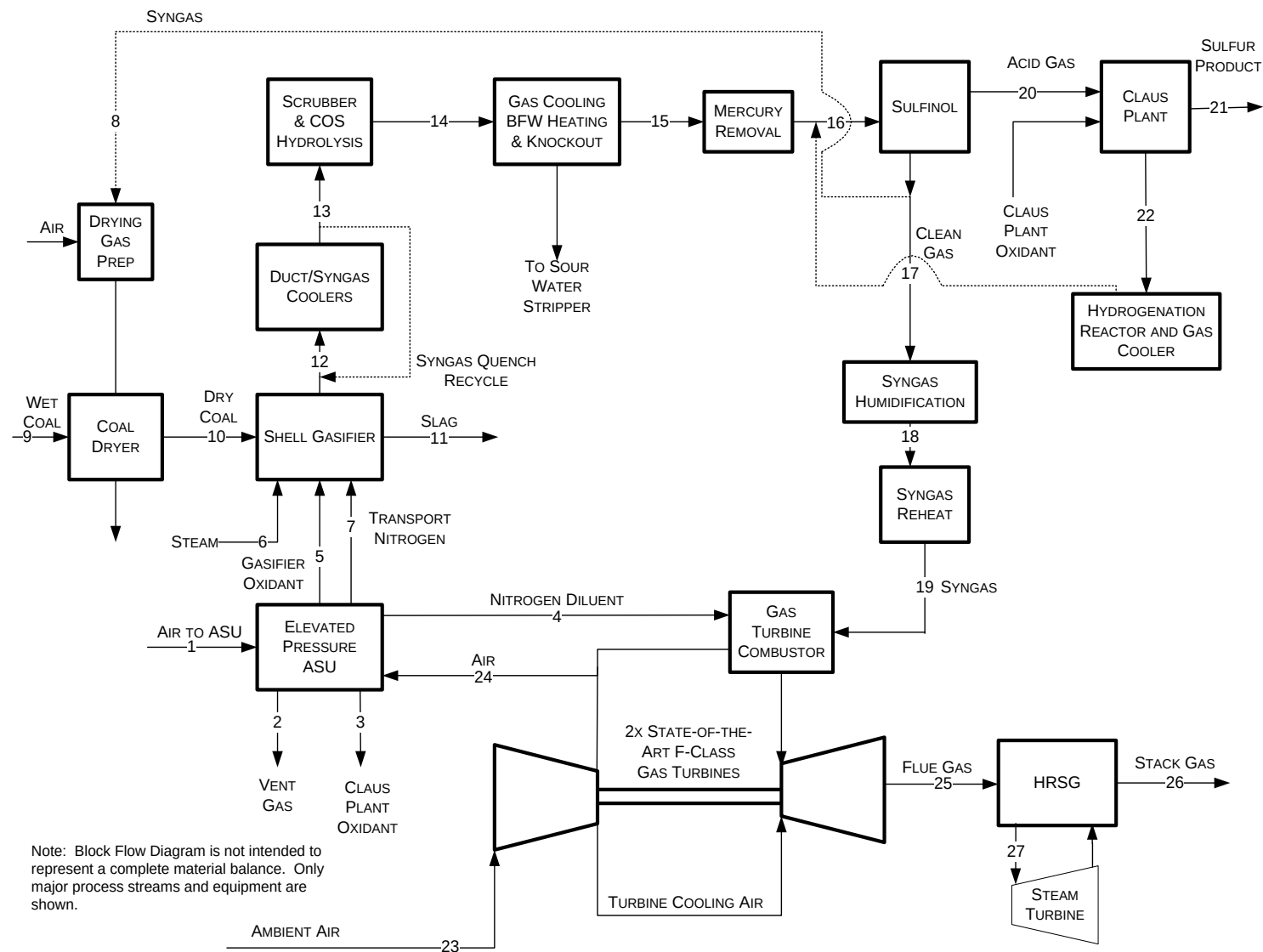
In this section the overall Shell gasification process for Case B1A is described. The system description follows the block flow diagram (BFD) in Exhibit 3-8 and stream numbers reference the same exhibit. The tables in Exhibit 3-9 provide process data for the numbered streams in the BFD.

3.2.2.1 Coal Preparation and Feed Systems

Coal receiving and handling is common to all cases and was covered in Section 3.1.1. The receiving and handling subsystem ends at the coal silo. The Shell process uses a dry feed system, which is sensitive to the coal moisture content. Coal moisture consists of two parts, surface moisture and inherent moisture. For coal to flow smoothly through the lock hoppers, the surface moisture must be removed. The Illinois No. 6 coal used in this report contains 11.12 percent total moisture on an as-received basis (stream 9). It was assumed that the coal must be dried to 5 percent moisture to allow for smooth flow through the dry feed system (stream 10).

The coal is simultaneously crushed and dried in the coal mill then delivered to a surge hopper with an approximate 2-hour capacity. The drying medium is provided by combining the off-gas from the Claus plant TGTU and a slipstream of clean syngas (stream 8) and passing them through an incinerator. The incinerator flue gas, with an oxygen content of 6 vol%, is then used to dry the coal in the mill.

The coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses nitrogen from the ASU to convey the coal to the gasifiers.

Exhibit 3-8 Case B1A block flow diagram, Shell IGCC without CO₂ capture

Source: NETL

Exhibit 3-9 Case B1A stream table, Shell IGCC without capture

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0262	0.0318	0.0023	0.0318	0.0000	0.0023	0.0100	0.0000	0.0000	0.0000	0.0095	0.0095	0.0096
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000	0.0006	0.0006	0.0006
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.5980	0.0000	0.0000	0.0000	0.5797	0.5797	0.5833
CO ₂	0.0003	0.0092	0.0000	0.0000	0.0000	0.0000	0.0000	0.0158	0.0000	0.0000	0.0000	0.0143	0.0143	0.0151
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007	0.0007	0.0000
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3142	0.0000	0.8090	0.0000	0.3006	0.3006	0.3024
H ₂ O	0.0099	0.2378	0.0000	0.0003	0.0000	1.0000	0.0004	0.0013	0.0000	0.0935	0.0000	0.0252	0.0252	0.0194
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0030	0.0000	0.0009	0.0009	0.0001
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0081	0.0081	0.0089
N ₂	0.7732	0.5018	0.0178	0.9919	0.0178	0.0000	0.9919	0.0601	0.0000	0.0163	0.0000	0.0567	0.0567	0.0570
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0036	0.0036	0.0036
O ₂	0.2074	0.2251	0.9504	0.0054	0.9504	0.0000	0.0054	0.0000	0.0000	0.0782	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	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	18,364	812	88	16,375	4,591	1,061	882	252	0	0	0	25,096	17,567	17,460
V-L Flowrate (kg/hr)	529,935	21,917	2,828	459,480	147,752	19,111	24,747	5,088	0	0	0	509,993	356,995	354,790
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	198,059	185,222	19,837	0	0	0
Temperature (°C)	15	21	32	93	32	343	32	45	15	16	1,427	1,079	191	177
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	5.10	5.41	3.65	0.10	5.79	4.24	4.24	4.03	3.90
Steam Table Enthalpy (kJ/kg) ^A	30.23	37.32	26.67	92.52	26.67	3,063.97	21.22	60.75	---	---	---	1,749.13	330.23	297.65
AspenPlus Enthalpy (kJ/kg) ^B	-97.58	-2,651.61	4.13	63.38	4.13	-12,907.38	-8.35	-3,577.38	-2,116.95	-768.33	2,165.43	-2,101.56	-3,520.45	-3,497.87
Density (kg/m ³)	1.2	1.6	11.0	24.4	11.0	20.1	60.5	27.8	---	16.5	---	7.6	21.0	20.9
V-L Molecular Weight	28.857	26.992	32.181	28.060	32.181	18.015	28.060	20.195	---	---	---	20.322	20.322	20.320
V-L Flowrate (lb _{mole} /hr)	40,486	1,790	194	36,101	10,122	2,339	1,944	555	0	0	0	55,328	38,729	38,493
V-L Flowrate (lb/hr)	1,168,307	48,319	6,236	1,012,980	325,738	42,132	54,557	11,218	0	0	0	1,124,342	787,039	782,178
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	436,646	408,345	43,732	0	0	0
Temperature (°F)	59	70	90	199	90	650	90	112	59	60	2,600	1,974	375	351
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	740.0	785.0	530.0	14.7	840.0	615.0	615.0	585.0	565.0
Steam Table Enthalpy (Btu/lb) ^A	13.0	16.0	11.5	39.8	11.5	1,317.3	9.1	26.1	---	---	---	752.0	142.0	128.0
AspenPlus Enthalpy (Btu/lb) ^B	-42.0	-1,140.0	1.8	27.2	1.8	-5,549.2	-3.6	-1,538.0	-910.1	-330.3	931.0	-903.5	-1,513.5	-1,503.8
Density (lb/ft ³)	0.076	0.101	0.687	1.521	0.687	1.257	3.774	1.734	---	1.029	---	0.475	1.310	1.303

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 3-9 Case B1A stream table, Shell IGCC without capture (continued)

	15	16	17	18	19	20	21	22	23	24	25	26	27
V-L Mole Fraction													
Ar	0.0098	0.0098	0.0100	0.0079	0.0079	0.0004	0.0000	0.0058	0.0092	0.0092	0.0086	0.0086	0.0000
CH ₄	0.0006	0.0006	0.0006	0.0005	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.5962	0.5845	0.5980	0.4741	0.4741	0.0167	0.0000	0.1034	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0155	0.0257	0.0158	0.0125	0.0125	0.4460	0.0000	0.2613	0.0003	0.0003	0.0752	0.0752	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.3091	0.3072	0.3142	0.2491	0.2491	0.0096	0.0000	0.0429	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0014	0.0015	0.0013	0.2083	0.2083	0.0063	0.0000	0.4259	0.0099	0.0099	0.0906	0.0906	1.0000
HCl	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
H ₂ S	0.0091	0.0089	0.0000	0.0000	0.0000	0.3872	0.0000	0.0016	0.0000	0.0000	0.0000	0.0000	0.0000
N ₂	0.0583	0.0618	0.0601	0.0476	0.0476	0.1338	0.0000	0.1585	0.7732	0.7732	0.7210	0.7210	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	0.0000
O ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.2074	0.1046	0.1046	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	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	17,081	17,426	16,771	21,155	21,155	402	0	510	110,253	4,410	137,428	137,428	39,521
V-L Flowrate (kg/hr)	347,994	358,740	338,688	417,666	417,666	14,964	0	13,723	3,181,557	127,262	3,962,154	3,962,154	711,987
Solids Flowrate (kg/hr)	0	0	0	0	0	0	4,959	0	0	0	0	0	0
Temperature (°C)	35	34	45	161	193	45	178	232	15	432	587	132	559
Pressure (MPa, abs)	3.72	3.69	3.65	3.21	3.17	3.65	0.41	0.41	0.10	1.62	0.10	0.10	12.51
Steam Table Enthalpy (kJ/kg) ^A	45.28	44.26	60.75	710.39	762.94	13.88	---	1,007.13	30.23	463.79	791.04	280.13	3,496.26
AspenPlus Enthalpy (kJ/kg) ^B	-3,555.60	-3,652.99	-3,577.38	-5,259.88	-5,207.33	-5,041.46	142.76	-7,809.89	-97.58	335.98	-1,163.40	-1,674.31	-12,474.05
Density (kg/m ³)	29.5	29.6	27.8	17.7	16.2	60.9	5,278.6	2.6	1.2	7.9	0.4	0.9	35.2
V-L Molecular Weight	20.373	20.587	20.195	19.743	19.743	37.190	---	26.923	28.857	28.857	28.831	28.831	18.015
V-L Flowrate (lb _{mole} /hr)	37,657	38,417	36,974	46,639	46,639	887	0	1,124	243,066	9,723	302,978	302,978	87,130
V-L Flowrate (lb/hr)	767,195	790,887	746,680	920,796	920,796	32,989	0	30,253	7,014,133	280,565	8,735,054	8,735,054	1,569,663
Solids Flowrate (lb/hr)	0	0	0	0	0	0	10,933	0	0	0	0	0	0
Temperature (°F)	94	94	112	321	380	112	353	450	59	810	1,088	270	1,038
Pressure (psia)	540.0	535.0	530.0	465.0	460.0	530.0	59.3	58.9	14.7	234.9	15.2	15.2	1,814.7
Steam Table Enthalpy (Btu/lb) ^A	19.5	19.0	26.1	305.4	328.0	6.0	---	433.0	13.0	199.4	340.1	120.4	1,503.1
AspenPlus Enthalpy (Btu/lb) ^B	-1,528.6	-1,570.5	-1,538.0	-2,261.3	-2,238.7	-2,167.4	61.4	-3,357.6	-42.0	144.4	-500.2	-719.8	-5,362.9
Density (lb/ft ³)	1.844	1.850	1.734	1.103	1.011	3.805	329.535	0.163	0.076	0.495	0.026	0.056	2.200

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

3.2.2.2 Gasifier

There are two Shell dry feed, pressurized, upflow, entrained, slagging gasifiers, operating at 4.2 MPa (615 psia) and processing a total of 4,753 tonnes/day (5,240 tpd) of as-received coal. The air separation plant supplies 3,614 tonnes/day (3,984 tpd) of 95 percent oxygen to the gasifiers (stream 5) and the Claus plant (stream 3). Coal reacts with oxygen and steam at a temperature of 1427°C (2600°F) in the gasifier to produce principally hydrogen and carbon monoxide with little carbon dioxide formed.

The gasifier's refractory-lined water wall is protected by molten slag that solidifies on the cooled walls.

3.2.2.3 Raw Gas Cooling/Particulate Removal

High-temperature heat recovery in each gasifier train is accomplished in three steps, including the gasifier jacket, which cools and solidifies slag touching the gasifier walls and maintains the syngas temperature at 1,427°C (2,600°F). The product gas from the gasifier is cooled to approximately 1,093°C (2,000°F) by adding cooled recycled fuel gas to lower the temperature below the ash melting point. Gas (stream 12) then goes through a duct cooler and syngas cooler, which lower the gas temperature from approximately 1,093°C (2,000°F) to 316°C (600°F), and produce HP steam for use in the steam cycle.

After passing through the duct cooler and syngas cooler, the syngas passes through a cyclone and a raw gas candle filter where a majority of the fine particles are removed and returned to the gasifier with the coal fuel. The filter consists of an array of ceramic candle elements in a pressure vessel. Fines produced by the gasification system are recirculated to extinction. The ash that is not carried out with the gas forms slag and runs down the interior walls, exiting the gasifier in liquid form. The slag is solidified in a quench tank for disposal (stream 11). Lockhoppers are used to reduce the pressure of the solids from 4.2 to 0.1 MPa (615 to 15 psia). The syngas scrubber removes additional PM further downstream.

After passing through the cyclone and ceramic candle filter array, the syngas is further cooled to 191°C (375°F) (stream 13) by raising IP steam.

3.2.2.4 Quench Gas Compressor

About 30 percent of the cooled syngas is recycled back to the gasifier exit as quench gas. A single-stage compressor is utilized to boost the pressure of the cooled fuel gas stream from 4.0 MPa (575 psia) to 4.0 MPa (585 psia) to provide quench gas to cool the syngas stream from the gasifier.

3.2.2.5 Syngas Scrubber/Sour Water Stripper

The raw syngas exiting the final raw gas cooler at 191°C (375°F) (stream 13) then enters the scrubber for removal of chlorides, SO₂, NH₃ and remaining particulate. The quench scrubber washes the syngas in a counter-current flow in two packed beds. The syngas leaves the scrubber saturated at a temperature of 94°C (201°F). The quench scrubber removes essentially all traces of entrained particles, principally unconverted carbon, slag, and metals. The bottoms from the scrubber are sent to the sour water stripper for processing.

The sour water stripper removes NH₃, SO₂, and other impurities from the waste stream of the scrubber. The sour gas stripper consists of a sour drum that accumulates sour water from the gas

scrubber and condensate from SGCs. Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the SRU. Remaining water is sent to wastewater treatment.

3.2.2.6 COS Hydrolysis, Mercury Removal and AGR

H₂S and COS are at significant concentrations, requiring removal for the power plant to achieve the low design level of SO₂ emissions. H₂S is removed in an AGR process; however, because COS is not readily removable, it is first catalytically converted to H₂S in a COS hydrolysis unit.

Following the water scrubber, the gas is reheated to 177°C (350°F) and fed to the COS hydrolysis reactor. The COS in the sour gas is hydrolyzed with steam over a catalyst bed to H₂S, which is more easily removed by the AGR solvent. Before the raw fuel gas can be treated in the AGR process (stream 14), it must be cooled to about 35°C (95°F). During this cooling through a series of heat exchangers, part of the water vapor condenses. This water, which contains some NH₃, is sent to the sour water stripper. The cooled syngas (stream 15) then passes through a series of 2 carbon beds to remove greater than 95 percent of the Hg (Section 3.1.4).

The Sulfinol process, developed by Shell in the early 1960s, is a combination process that uses a mixture of amines and a physical solvent. The solvent consists of an aqueous amine and sulfolane. Sulfinol-D uses diisopropanolamine (DIPA), while Sulfinol-M uses MDEA. The mixed solvents allow for better solvent loadings at high acid gas partial pressures and higher solubility of COS and organic sulfur compounds than straight aqueous amines. Sulfinol-M was selected for this application.

The sour syngas is fed directly into an HP contactor. The HP contactor is an absorption column in which the H₂S, COS, CO₂, and small amounts of H₂ and CO are removed from the gas by the Sulfinol solvent. The overhead gas stream from the HP contactor is then washed with water in the sweet gas scrubber before leaving the unit as the feed gas to the sulfur polishing unit.

The rich solvent from the bottom of the HP contactor flows through a hydraulic turbine and is flashed in the rich solvent flash vessel. The flashed gas is then scrubbed in the LP contactor with lean solvent to remove H₂S and COS. The overhead from the LP contactor is flashed in the LP KO drum. This gas can be used as a utility fuel gas, consisting primarily of H₂ and CO, at 0.8 MPa (118 psia) and 38°C (101°F). The solvent from the bottom of the LP contactor is returned to the rich solvent flash vessel.

Hot, lean solvent in the lean/rich solvent exchanger then heats the flashed rich solvent before entering the stripper. The stripper strips the H₂S, COS, and CO₂ from the solvent at LP with heat supplied through the stripper reboiler. The acid gas stream to sulfur recovery/tail gas cleanup is recovered as the flash gas from the stripper accumulator. The lean solvent from the bottom of the stripper is cooled in the lean/rich solvent exchanger and the lean solvent cooler. Most of the lean solvent is pumped to the HP contactor. A small amount goes to the LP contactor.

The Sulfinol process removes essentially all of the CO₂ along with the H₂S and COS. The acid gas fed to the SRU contains 39 vol% H₂S and 45 vol% CO₂. The CO₂ passes through the SRU, the TGTU and ultimately is vented through the coal dryer. Since the amount of CO₂ in the syngas is initially small, this does not have a significant effect on the mass flow reaching the CT. However, the costs of the sulfur recovery/tail gas cleanup are higher than that for a sulfur removal process producing an acid gas stream with a higher sulfur concentration.

3.2.2.7 Claus Unit

The SRU is a Claus bypass type SRU utilizing oxygen (stream 3) instead of air. The Claus plant produces molten sulfur (stream 21) by reacting approximately one third of the H_2S in the feed to SO_2 , then reacting the H_2S and SO_2 to sulfur and water. The use of Claus technology results in an overall sulfur recovery exceeding 99 percent.

Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant produces approximately 119 tonnes/day (131 tpd) of elemental sulfur. Feed for this case consists of acid gas from both the acid gas cleanup unit (stream 20) and a vent stream from the sour water stripper in the gasifier section. A slipstream of clean syngas (stream 8) is passed through an incinerator and combusted with air. The hot, nearly inert incinerator off gas is used to dry coal before being vented to the atmosphere.

In the furnace waste heat boiler, 11,991 kg/hr (26,435 lb/hr) of 3.0 MPa (430 psia) steam is generated. This steam is used to satisfy all Claus process preheating and reheating requirements as well as to provide some steam to the medium-pressure steam header. The sulfur condensers produce 0.34 MPa (50 psig) steam for the LP steam header.

3.2.2.8 Power Block

Clean syngas exiting the Sulfinol absorber (stream 17) is humidified because there is not sufficient nitrogen from the ASU to provide the level of dilution required. The moisturized syngas (stream 18) is reheated (stream 19), further diluted with nitrogen from the ASU (stream 4) and steam, and enters the state-of-the-art 2008 F-class CT burner. The CT compressor provides combustion air to the burner and also 19 percent of the air requirements in the ASU (stream 24). The exhaust gas exits the CT at 587°C (1,088°F) (stream 25) and enters the HRSG where additional heat is recovered until the flue gas exits the HRSG at 132°C (270°F) (stream 26) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced, commercially available steam turbine using a 12.4 MPa/559°C/559°C (1,800 psig/1,038°F/1,038°F) steam cycle.

3.2.2.9 Air Separation Unit (ASU)

The ASU is designed to produce a nominal output of 3,614 tonnes/day (3,984 tpd) of 95 mol% O_2 for use in the gasifier (stream 5) and SRU (stream 3). The plant is designed with two production trains. The air compressor is powered by an electric motor. Approximately 11,028 tonnes/day (12,156 tpd) of nitrogen are also recovered, compressed, and used as dilution in the CT combustor. About 4 percent of the CT air is used to supply approximately 19 percent of the ASU air requirements, in the non-capture case.

3.2.2.10 Balance of Plant

Balance of plant items were covered in Sections 3.1.9, 3.1.10, and 3.1.12.

3.2.3 Key System Assumptions

System assumptions for Cases B1A and B1B (Shell IGCC with and without CO_2 capture) are compiled in Exhibit 3-10.

3.2.3.1 Balance of Plant – Cases B1A and B1B

The balance of plant assumptions are common to all cases and presented in Exhibit 3-11.

Exhibit 3-10 Shell IGCC plant study configuration matrix

Case	B1A	B1B
Gasifier Pressure, MPa (psia)	4.2 (615)	4.2 (615)
O ₂ :Coal Ratio, kg O ₂ /kg dry coal	0.84	0.84
Carbon Conversion, %	99.5	99.5
Syngas HHV at Gasifier Outlet, kJ/Nm ³ (Btu/scf)	10,841 (291)	10,849 (291)
Steam Cycle, MPa/°C/°C (psig/°F/°F)	12.4/559/559 (1,800/1,038/1,038)	12.4/534/534 (1,800/993/993)
Condenser Pressure, mm Hg (in Hg)	51 (2.0)	51 (2.0)
CT	2x State-of-the-art 2008 F-Class (232 MW output each)	2x State-of-the-art 2008 F-Class (232 MW output each)
Gasifier Technology	Shell	Shell
Oxidant	95 vol% Oxygen	95 vol% Oxygen
Coal	Illinois No. 6	Illinois No. 6
Coal Feed Moisture Content, %	5	5
COS Hydrolysis	Yes	Occurs in SGS
SGS	No	Yes
H ₂ S Separation	Sulfinol-M	Selexol 1 st Stage
Sulfur Removal, %	99.7	99.9
Sulfur Recovery	Claus Plant with Tail Gas Treatment / Elemental Sulfur	Claus Plant with Tail Gas Treatment / Elemental Sulfur
Particulate Control	Cyclone, Candle Filter, Scrubber, and AGR Absorber	Cyclone, Candle Filter, Scrubber, and AGR Absorber
Mercury Control	Carbon Bed	Carbon Bed
NO _x Control	MNQC (LNB), N ₂ Dilution, Humidification and steam dilution	MNQC (LNB), N ₂ Dilution and Humidification
CO ₂ Separation	N/A	Selexol 2 nd Stage
Overall Carbon Capture	N/A	90.0%
CO ₂ Sequestration	N/A	Off-site Saline Formation

Exhibit 3-11 Balance of plant assumptions

Parameters	Values
<u>Cooling system</u>	Recirculating Wet Cooling Tower
<u>Fuel and Other storage</u>	
Coal	30 days
Slag	30 days
Sulfur	30 days
Sorbent	30 days
<u>Plant Distribution Voltage</u>	
Motors below 1 hp	110/220 V
Motors between 1 hp and 250 hp	480 V
Motors between 250 hp and 5,000 hp	4,160 V
Motors above 5,000 hp	13,800 V
Steam and Combustion Turbine Generators	24,000 V
Grid Interconnection Voltage	345 kV
<u>Water and Waste Water</u>	
Makeup Water	The water supply is 50 percent from a local POTW and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources
Process Wastewater	Water associated with gasification activity and storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge.
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant was sized for 5.68 cubic meters per day (1,500 gallons per day)
Water Discharge	Most of the process wastewater is recycled to the cooling tower basin. Blowdown is treated for chloride and metals, and discharged.

3.2.4 Sparing Philosophy

The sparing philosophy for cases B1A and B1B is provided below. Dual trains are used to accommodate the size of commercial CTs. There is no redundancy other than normal sparing of rotating equipment. The plant design consists of the following major subsystems:

- Two ASUs (2 x 50%)
- Two trains of coal drying and dry feed systems (2 x 50%)
- Two trains of gasification, including gasifier, SGC, cyclone, and barrier filter (2 x 50%)
- Two trains of syngas clean-up process (2 x 50%)
- Two trains of Sulfinol-M acid gas removal in Case B1A and two-stage Selexol in Case B1B (2 x 50%)

- Two CO₂ compression systems (2 x 50%) in Case B1B
- One train of Claus-based sulfur recovery (1 x 100%)
- Two CT/HRSG tandems (2 x 50%)
- One steam turbine (1 x 100%)

3.2.5 Case B1A - Performance Results

The plant produces a net output of 629 MWe at a net plant efficiency of 42.1 percent (HHV basis). Shell has reported expected efficiencies using bituminous coal of around 44-45 percent (HHV basis), although this value excluded the net power impact of coal drying. (48) Accounting for coal drying would reduce the efficiency by only about 0.5-1 percentage points so the efficiency results for the Shell case are still lower in this report than those reported by the vendor.

Overall performance for the entire plant is summarized in Exhibit 3-12; Exhibit 3-13 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for approximately 79 percent of the total auxiliary load distributed between the main air compressor, oxygen compressor, nitrogen compressor, and ASU auxiliaries. The cooling water system, including the CWP and cooling tower fan, accounts for approximately 5 percent of the auxiliary load, and the BFW pumps account for an additional 4 percent. All other individual auxiliary loads are 3 percent or less of the total.

Exhibit 3-12 Case B1A plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	273
Total Gross Power, MWe	737
Air Separation Unit Main Air Compressor, kWe	45,190
Oxygen Compressor, kWe	8,890
Nitrogen Compressors, kWe	29,850
CO ₂ Compression, kWe	0
Acid Gas Removal, kWe	620
Balance of Plant, kWe	23,550
Total Auxiliaries, MWe	108
Net Power, MWe	629
HHV Net Plant Efficiency (%)	42.1%
HHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	8,546 (8,100)
HHV Cold Gas Efficiency, %	84.5%
HHV Combustion Turbine Efficiency, %	38.3%
LHV Net Plant Efficiency (%)	43.7%
LHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	8,242 (7,812)
LHV Cold Gas Efficiency, %	82.8%
LHV Combustion Turbine Efficiency, %	40.5%
Steam Turbine Cycle Efficiency, %	44.8%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	8,037 (7,618)
Condenser Duty, GJ/hr (MMBtu/hr)	1,397 (1,324)
As-Received Coal Feed, kg/hr (lb/hr)	198,059 (436,646)
HHV Thermal Input, kWt	1,492,878
LHV Thermal Input, kWt	1,439,899
Raw Water Withdrawal, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.025 (6.6)
Raw Water Consumption, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.020 (5.3)
O ₂ :Coal	0.705

Exhibit 3-13 Case B1A plant power summary

Power Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	273
Total Gross Power, MWe	737
Auxiliary Load Summary	
Coal Handling, kWe	440
Coal Milling, kWe	2,040
Sour Water Recycle Slurry Pump, kWe	0
Slag Handling, kWe	520
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	45,190
Oxygen Compressor, kWe	8,890
Nitrogen Compressors, kWe	29,850
Feedwater Pumps, kWe	4,500
Quench Water Pump, kWe	0
Syngas Recycle Compressor, kWe	680
CO ₂ Compression, kWe	0
Scrubber Pumps, kWe	770
Acid Gas Removal, kWe	620
Combustion Turbine Auxiliaries, kWe	1,000
Steam Turbine Auxiliaries, kWe	100
Condensate Pumps, kWe	230
Circulating Water Pumps, kWe	3,400
Ground Water Pumps, kWe	380
Cooling Tower Fans, kWe	1,760
Claus Plant/TGTU Auxiliaries, kWe	250
Claus Plant TG Recycle Compressor, kWe	890
Miscellaneous Balance of Plant ^A , kWe	3,000
Transformer Losses, kWe	2,590
Total Auxiliaries, MWe	108
Net Power, MWe	629

^AIncludes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.2.5.1 Environmental Performance

The environmental targets for emissions of Hg, NO_x, SO₂, and PM were presented in Section 2.3. A summary of the plant air emissions for Case B1A is presented in Exhibit 3-14. All HCl is assumed to be removed and is, therefore, not reported.

Exhibit 3-14 Case B1A air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year) ^A	kg/MWh (lb/MWh) ^B
SO ₂	0.002 (0.004)	68 (75)	0.013 (0.029)
NO _x	0.025 (0.059)	957 (1,055)	0.185 (0.409)
Particulates	0.003 (0.0071)	115 (127)	0.022 (0.049)
Hg	1.87E-7 (4.34E-7)	0.007 (0.008)	1.36E-6 (3.00E-6)
CO ₂	85 (197)	3,188,643 (3,514,877)	617 (1,361)
CO ₂ ^C	-	-	723 (1,595)

^ACalculations based on an 80 percent capacity factor

^BEmissions are based on gross power except where otherwise noted

^CCO₂ emissions based on net power instead of gross power

The low level of SO₂ emissions is achieved by capturing the sulfur in the gas by the Sulfinol-M AGR process. The AGR process removes over 99 percent of the sulfur compounds in the fuel gas down to a level of less than 6 ppmv. This results in a concentration in the HRSG flue gas of less than 2 ppmv. The H₂S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is compressed and recycled back to the AGR to capture most of the remaining sulfur. The SO₂ emissions in Exhibit 3-14 include both the stack emissions and the coal dryer emissions.

NO_x emissions are limited by the use of nitrogen dilution, humidification, and steam dilution to 15 ppmvd (as NO₂ @ 15 percent O₂). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process and destroyed in the Claus plant burner. This helps lower NO_x levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a cyclone and a barrier filter in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of the mercury is captured from the syngas by dual activated carbon beds.

CO₂ emissions represent the uncontrolled discharge from the process.

The carbon balance for the plant is shown in Exhibit 3-15. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not neglected here since the Aspen model accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag and as CO₂ in the stack gas and ASU vent gas.

Exhibit 3-15 Case B1A carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	126,252 (278,339)	Stack Gas	124,177 (273,764)
Air (CO ₂)	508 (1,120)	Coal Dryer Stack	1,863 (4,106)
		CO ₂ Product	0 (0)
		ASU Vent	89 (197)
		Slag	631 (1,392)
Total	126,760 (279,459)	Total	126,760 (279,459)

Exhibit 3-16 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur in the coal drying gas, and sulfur emitted in the stack gas. Sulfur in the slag is considered to be negligible.

Exhibit 3-16 Case B1A sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	4,964 (10,944)	Stack Gas	5 (11)
		Coal Dryer Stack	0 (0)
		CO ₂ Product	0 (0)
		Elemental Sulfur	4,959 (10,933)
Total	4,964 (10,944)	Total	4,964 (10,944)

Exhibit 3-17 shows the overall water balance for the plant.

Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is defined as the water removed from the ground or diverted from a municipal source for use in the plant and was assumed to be provided 50 percent by a POTW and 50 percent from groundwater. Raw water withdrawal can be represented by the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, quench system makeup, and slag handling makeup. 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 3-17 Case B1A water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)
Slag Handling	0.43 (114)	0.17 (45)	0.26 (69)	–	0.26 (69)
Slurry Water	–	–	–	–	–
Quench/Wash	–	–	–	–	–
Humidifier	1.38 (365)	–	1.38 (365)	–	1.38 (365)
SWS Blowdown	–	–	–	0.00 (0.45)	0.00 (-0.45)
Condenser Makeup	1.07 (282)	–	1.07 (282)	–	1.07 (282)
BFW Makeup	0.23 (62)	–	0.23 (62)	–	0.23 (62)
Gasifier Steam	0.32 (84)	–	0.32 (84)	–	0.32 (84)
Shift Steam	–	–	–	–	–
CT Steam Dilution	0.51 (135)	–	0.51 (135)	–	0.51 (135)
Cooling Tower	13.25 (3,501)	0.25 (67)	13.00 (3,435)	2.98 (787)	10.02 (2,647)
BFW Blowdown	–	0.23 (62)	-0.23 (-62)	–	-0.23 (-62)
SWS Blowdown	–	0.02 (4.50)	-0.02 (-4.50)	–	-0.02 (-4.50)
Total	16.13 (4,261)	0.42 (112)	15.71 (4,150)	2.98 (788)	12.73 (3,362)

3.2.5.2 Heat and Mass Balance Diagrams

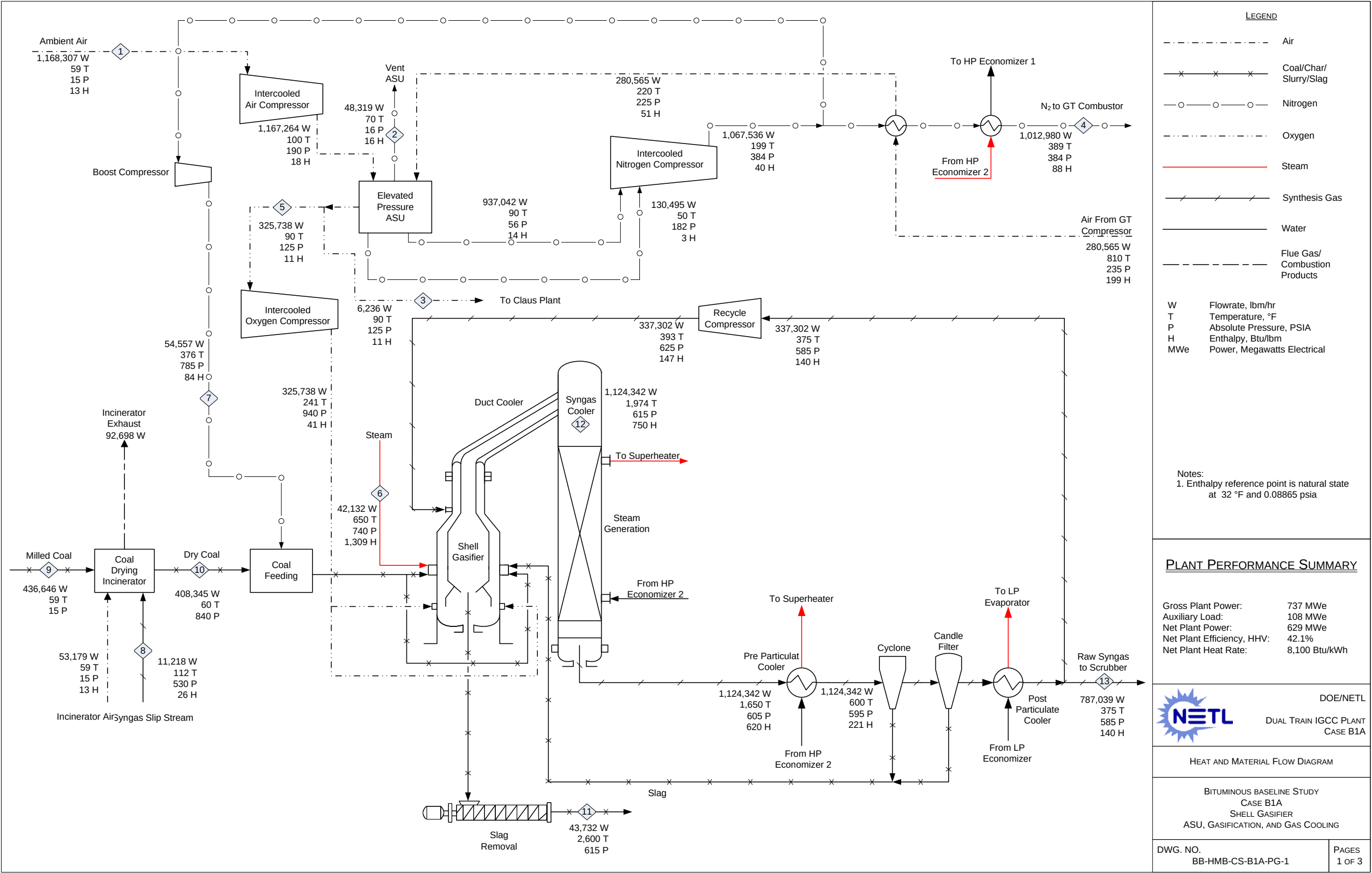
Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-18 through Exhibit 3-20:

- Coal gasification and ASU
- Syngas cleanup, sulfur recovery, and tail gas recycle
- Combined cycle power generation, steam, and FW

An overall plant energy balance is provided in tabular form in Exhibit 3-21. The power out is the combined CT and steam turbine power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-12) is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

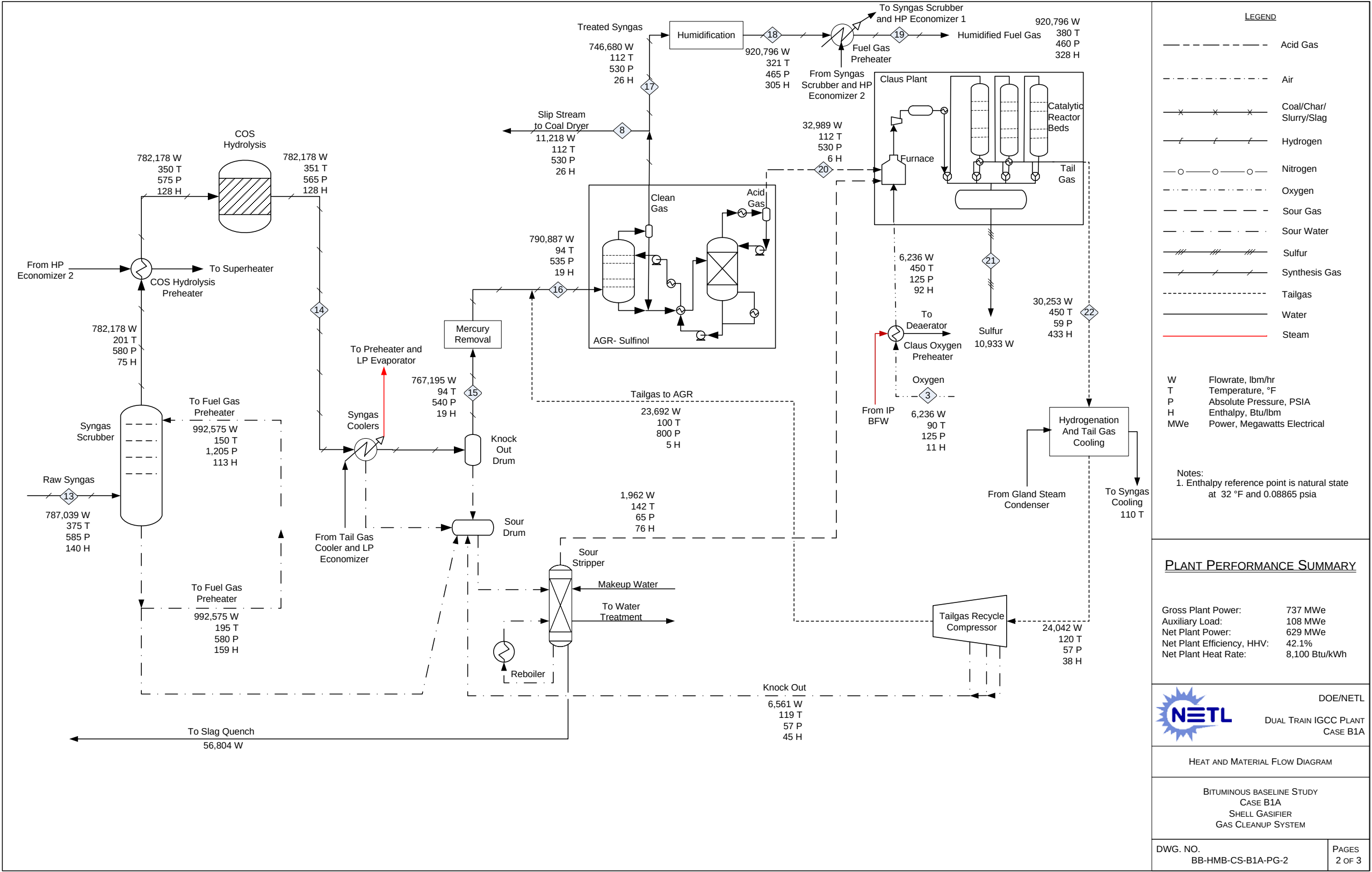
This page intentionally left blank

Exhibit 3-18 Case B1A coal gasification and ASU heat and mass balance schematic



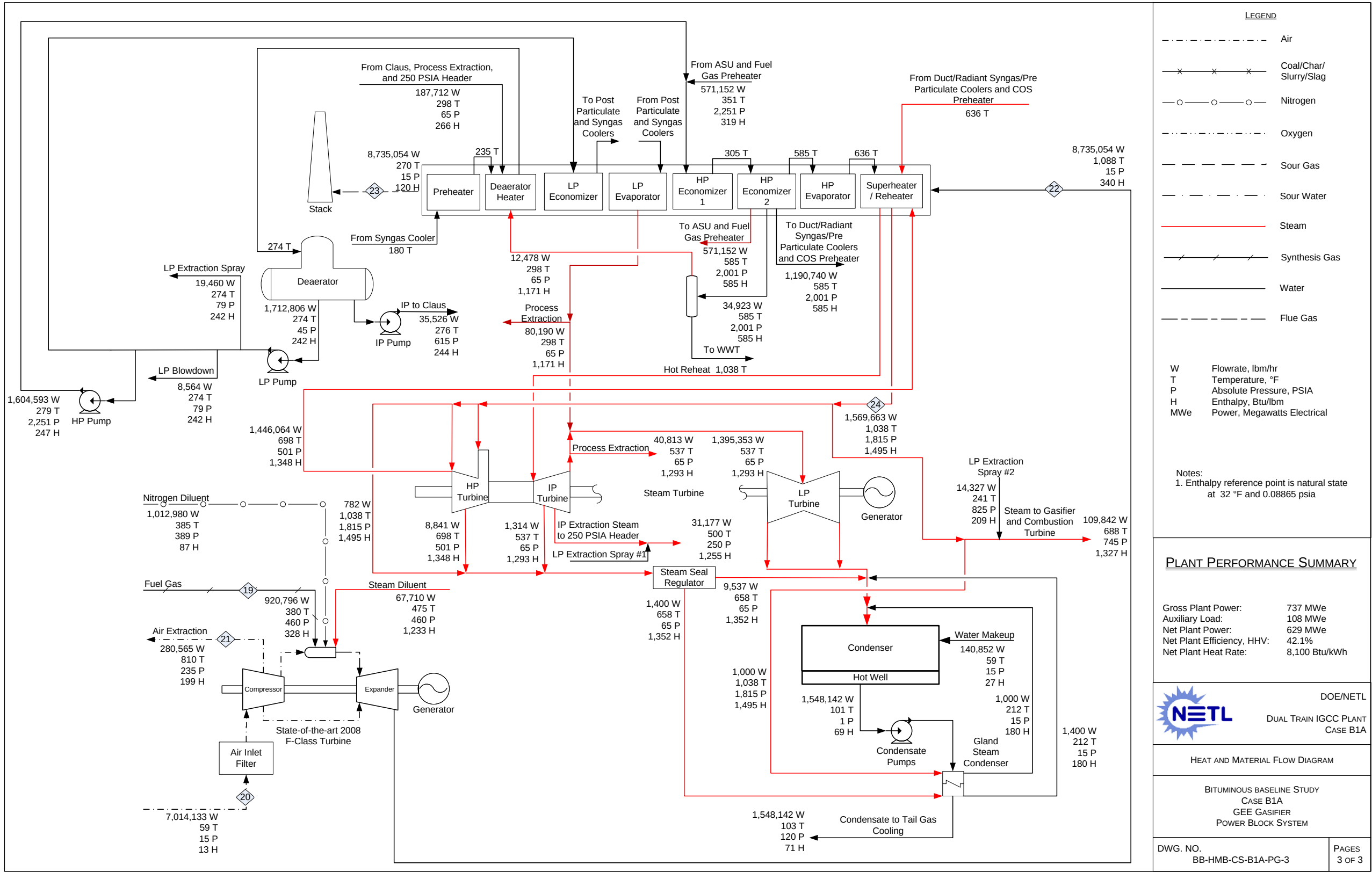
Source: NETL

Exhibit 3-19 Case B1A syngas cleanup heat and mass balance schematic



Source: NETL

Exhibit 3-20 Case B1A combined cycle power generation heat and mass balance schematic



Source: NETL

This page intentionally left blank

Exhibit 3-21 Case B1A overall energy balance (0°C [32°F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In (GJ/hr)				
Coal	5,374 (5,094)	4.5 (4.3)	–	5,379 (5,098)
Air	–	112.9 (107.0)	–	112.9 (107.0)
Raw Water Makeup	–	59.1 (56.0)	–	59.1 (56.0)
Auxiliary Power	–	–	389 (369)	389 (369)
TOTAL	5,374 (5,094)	176.5 (167.3)	389 (369)	5,940 (5,630)
Heat Out (GJ/hr)				
ASU Vent	–	0.8 (0.8)	–	0.8 (0.8)
Slag	20.7 (19.6)	33.5 (31.7)	–	54.2 (51.4)
Stack Gas	–	1,110 (1,052)	–	1,110 (1,052)
Sulfur	46.0 (43.6)	0.6 (0.5)	–	46.5 (44.1)
Motor Losses and Design Allowances	–	–	51.8 (49.1)	51.8 (49.1)
Condenser	–	1,397 (1,324)	–	1,397 (1,324)
Non-Condenser Cooling Tower Loads	–	246 (233)	–	246 (233)
CO ₂	–	–	–	–
Blowdown	–	22.9 (21.7)	–	22.9 (21.7)
CO ₂ Capture Losses	–	–	–	–
<i>Ambient Losses^A</i>	–	113.9 (108.0)	–	113.9 (108.0)
Power	–	–	2,653 (2,515)	2,653 (2,515)
TOTAL	66.7 (63.2)	2,925 (2,772)	2,705 (2,564)	5,696 (5,399)
Unaccounted Energy ^B	–	244 (231)	–	244 (231)

^AAmbient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers.

^BBy difference

3.2.6 Case B1A - Major Equipment List

Major equipment items for the Shell gasifier with no CO₂ capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 3.2.7. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

Case B1A – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	40 tonne (50 ton)	2	1
7	Feeder	Vibratory	160 tonne/hr (180 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	330 tonne/hr (360 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	160 tonne (180 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/ tripper	330 tonne/hr (360 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	330 tonne/hr (360 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	730 tonne (800 ton)	3	0

Case B1A – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	70 tonne/hr (80 tph)	3	0
2	Conveyor No. 6	Belt w/ tripper	220 tonne/hr (240 tph)	1	0
3	Roller Mill Feed Hopper	Dual Outlet	440 tonne (480 ton)	1	0
4	Weigh Feeder	Belt	110 tonne/hr (120 tph)	2	0
5	Coal Dryer and Pulverizer	Rotary	110 tonne/hr (120 tph)	2	0
6	Coal Dryer Feed Hopper	Vertical Hopper	220 tonne (240 ton)	2	0

Case B1A – Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1,165,000 liters (308,000 gal)	2	0
2	Condensate Pumps	Vertical canned	6,480 lpm @ 90 m H ₂ O (1,710 gpm @ 300 ft H ₂ O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	436,000 kg/hr (962,000 lb/hr)	2	0
4	Intermediate-Pressure Feedwater Pump	Horizontal centrifugal, single stage	7,410 lpm @ 30 m H ₂ O (1,960 gpm @ 90 ft H ₂ O)	2	1
5	High-Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 6,940 lpm @ 1,800 m H ₂ O (1,830 gpm @ 6,100 ft H ₂ O)	2	1
6	High-Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 2,470 lpm @ 220 m H ₂ O (650 gpm @ 730 ft H ₂ O)	2	1
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded screw	28 m ³ /min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m ³ /min(1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	135 GJ/hr (128 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	48,500 lpm @ 20 m H ₂ O (12,800 gpm @ 70 ft H ₂ O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H ₂ O (1,000 gpm @ 350 ft H ₂ O)	1	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H ₂ O (700 gpm @ 250 ft H ₂ O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	4,400 lpm @ 20 m H ₂ O (1,160 gpm @ 60 ft H ₂ O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	2,930 lpm @ 270 m H ₂ O (780 gpm @ 880 ft H ₂ O)	3	1
16	Filtered Water Pumps	Stainless steel, single suction	1,780 lpm @ 50 m H ₂ O (470 gpm @ 160 ft H ₂ O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	852,000 liter (225,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	320 lpm (90 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0

Case B1A – Account 4: Gasifier, ASU, and Accessories Including Low Temperature Heat Recovery and Fuel Gas Saturation

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized dry-feed, entrained bed	2,600 tonne/day, 4.2 MPa (2,900 tpd, 615 psia)	2	0
2	Synthesis Gas Cooler	Convective spiral-wound tube boiler	280,000 kg/hr (618,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	280,000 kg/hr (618,000 lb/hr) Design efficiency 90%	2	0
4	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0
5	Syngas Scrubber Including Sour Water Stripper	Vertical upflow	196,000 kg/hr (433,000 lb/hr)	2	0
6	Raw Gas Coolers	Shell and tube with condensate drain	193,000 kg/hr (425,000 lb/hr)	8	0
7	Raw Gas Knockout Drum	Vertical with mist eliminator	192,000 kg/hr, 35°C, 3.8 MPa (423,000 lb/hr, 94°F, 545 psia)	2	0
8	Saturation Water Economizers	Shell and tube	131 GJ/hr (124 MMBtu/hr)	2	0
9	Fuel Gas Saturator	Vertical tray tower	230,000 kg/hr, 161°C, 3.3 MPa (506,000 lb/hr, 321°F, 480 psia)	2	0
10	Saturator Water Pump	Centrifugal	3,600 lpm @ 13 m H ₂ O (900 gpm @ 42 ft H ₂ O)	2	2
11	Synthesis Gas Reheater	Shell and tube	186,000 kg/hr (411,000 lb/hr)	2	0
12	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	196,000 kg/hr (433,000 lb/hr) syngas	2	0
13	ASU Main Air Compressor	Centrifugal, multi-stage	4,000 m ³ /min @ 1.3 MPa (141,000 scfm @ 190 psia)	2	0
14	Cold Box	Vendor design	2,000 tonne/day (2,200 tpd) of 95% purity oxygen	2	0
15	Oxygen Compressor	Centrifugal, multi-stage	1,000 m ³ /min (35,000 scfm) Suction - 1.0 MPa (130 psia) Discharge - 6.5 MPa (940 psia)	2	0
16	Primary Nitrogen Compressor	Centrifugal, multi-stage	3,000 m ³ /min (116,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0
17	Secondary Nitrogen Compressor	Centrifugal, single-stage	460 m ³ /min (16,000 scfm) Suction - 1.3 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
18	Transport Nitrogen Boost Compressor	Centrifugal, single-stage	190 m ³ /min (7,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 5.4 MPa (790 psia)	2	0
19	Extraction Air Heat Exchanger	Gas-to-gas, vendor design	70,000 kg/hr, 432°C, 1.6 MPa (154,000 lb/hr, 810°F, 235 psia)	2	0

Case B1A – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	191,000 kg/hr (422,000 lb/hr), 35°C (94°F), 3.7 MPa (540 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	191,000 kg/hr (422,000 lb/hr), 35°C (94°F), 3.7 MPa (535 psia)	2	0
3	Sulfur Plant	Claus type	131 tonne/day (144 tpd)	1	0
4	COS Hydrolysis Reactor	Fixed bed, catalytic	195,000 kg/hr (430,000 lb/hr), 177°C (350°F), 4.0 MPa (580 psia)	2	0
5	Acid Gas Removal Plant	Sulfinol	197,000 kg/hr (435,000 lb/hr), 34°C (94°F), 3.7 MPa (535 psia)	2	0
6	Hydrogenation Reactor	Fixed bed, catalytic	15,000 kg/hr (33,000 lb/hr), 232°C (450°F), 0.4 MPa (58.9 psia)	1	0
7	Tail Gas Recycle Compressor	Centrifugal	12,000 kg/hr (26,000 lb/hr) each	1	0

Case B1A – Account 6: Combustion Turbine and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	State-of-the-art 2008 F-class	230 MW	2	0
2	Combustion Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0

Case B1A – Account 7: HRSG, Ducting, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.4 m (28 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 391,593 kg/hr, 12.4 MPa/559°C (863,314 lb/hr, 1,800 psig/1,038°F)	2	0

Case B1A – Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	287 MW 12.4 MPa/559°C/559°C (1,800 psig/ 1,038°F/1,038°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	320 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,540 GJ/hr (1,460 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0

Case B1A – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	342,000 lpm @ 30 m (90,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 1,900 GJ/hr (1,800 MMBtu/hr) heat duty	1	0

Case B1A – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	208,000 liters (55,000 gal)	2	0
2	Slag Crusher	Roll	11 tonne/hr (12 tph)	2	0
3	Slag Depressurizer	Lock Hopper	11 tonne/hr (12 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	125,000 liters (33,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	56,000 liters (15,000 gal)	2	0
6	Slag Conveyor	Drag chain	11 tonne/hr (12 tph)	2	0
7	Slag Separation Screen	Vibrating	11 tonne/hr (12 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	11 tonne/hr (12 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	178,000 liters (47,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	50 lpm @ 14 m H ₂ O (10 gpm @ 46 ft H ₂ O)	2	2
11	Grey Water Storage Tank	Field erected	57,000 liters (15,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	200 lpm @ 430 m H ₂ O (50 gpm @ 1,420 ft H ₂ O)	2	2
13	Slag Storage Bin	Vertical, field erected	800 tonne (900 tons)	2	0
14	Unloading Equipment	Telescoping chute	90 tonne/hr (100 tph)	1	0

Case B1A – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	24 kV/345 kV, 290 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 47 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 23 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 3 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case B1A – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.2.7 Case B1A - Cost EstimatingCosts Results

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-22 shows a detailed breakdown of the capital costs; Exhibit 3-23 shows the owner's costs, TOC, and TASC; Exhibit 3-24 shows the initial and annual O&M costs; and Exhibit 3-25 shows the COE breakdown.

The estimated TPC of the Shell gasifier with no CO₂ capture is \$2,725/kW. Process contingency represents 2.8 percent of the TPC, and project contingency represents 13.8 percent. The COE is \$107.0/MWh.

Exhibit 3-22 Case B1A total plant cost details

Case:		B1A – Shell IGCC w/o CO ₂				Estimate Type:		Conceptual			
Plant Size (MW,net):		629				Cost Base:		Jun 2011			
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
1		Coal & Sorbent Handling									
1.1	Coal Receive & Unload	\$4,254	\$0	\$2,050	\$0	\$6,305	\$630	\$0	\$1,387	\$8,322	\$13
1.2	Coal Stackout & Reclaim	\$5,498	\$0	\$1,314	\$0	\$6,812	\$681	\$0	\$1,499	\$8,992	\$14
1.3	Coal Conveyors & Yd Crush	\$5,111	\$0	\$1,301	\$0	\$6,412	\$641	\$0	\$1,411	\$8,464	\$13
1.4	Other Coal Handling	\$1,337	\$0	\$301	\$0	\$1,638	\$164	\$0	\$360	\$2,162	\$3
1.9	Coal & Sorbent Hnd. Foundations	\$0	\$2,841	\$7,425	\$0	\$10,265	\$1,027	\$0	\$2,258	\$13,550	\$22
	Subtotal	\$16,200	\$2,841	\$12,391	\$0	\$31,432	\$3,143	\$0	\$6,915	\$41,491	\$66
2		Coal & Sorbent Prep & Feed									
2.1	Coal Crushing & Drying	\$48,619	\$2,932	\$6,987	\$0	\$58,539	\$5,854	\$0	\$12,878	\$77,271	\$123
2.2	Prepared Coal Storage & Feed	\$2,303	\$553	\$356	\$0	\$3,212	\$321	\$0	\$707	\$4,240	\$7
2.3	Dry Coal Injection System	\$75,787	\$873	\$6,942	\$0	\$83,603	\$8,360	\$0	\$18,393	\$110,356	\$175
2.4	Misc. Coal Prep & Feed	\$1,266	\$925	\$2,725	\$0	\$4,917	\$492	\$0	\$1,082	\$6,490	\$10
2.9	Coal & Sorbent Feed Foundation	\$0	\$4,645	\$3,986	\$0	\$8,631	\$863	\$0	\$1,899	\$11,392	\$18
	Subtotal	\$127,976	\$9,929	\$20,996	\$0	\$158,901	\$15,890	\$0	\$34,958	\$209,749	\$334
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	\$3,734	\$6,444	\$3,379	\$0	\$13,557	\$1,356	\$0	\$2,983	\$17,896	\$28
3.2	Water Makeup & Pretreating	\$688	\$71	\$378	\$0	\$1,137	\$114	\$0	\$375	\$1,626	\$3
3.3	Other Feedwater Subsystems	\$2,099	\$694	\$620	\$0	\$3,413	\$341	\$0	\$751	\$4,505	\$7
3.4	Service Water Systems	\$402	\$802	\$2,765	\$0	\$3,969	\$397	\$0	\$1,310	\$5,676	\$9
3.5	Other Boiler Plant Systems	\$2,168	\$810	\$1,994	\$0	\$4,971	\$497	\$0	\$1,094	\$6,562	\$10
3.6	FO Supply Sys & Nat Gas	\$17,736	\$712	\$660	\$0	\$19,108	\$1,911	\$0	\$4,204	\$25,223	\$40
3.7	Waste Treatment Equipment	\$931	\$0	\$576	\$0	\$1,507	\$151	\$0	\$497	\$2,155	\$3
3.8	Misc. Power Plant Equipment	\$1,267	\$169	\$658	\$0	\$2,094	\$209	\$0	\$691	\$2,994	\$5
	Subtotal	\$29,025	\$9,703	\$11,029	\$0	\$49,757	\$4,976	\$0	\$11,904	\$66,637	\$106
4		Gasifier & Accessories									
4.1	Syngas Cooler Gasifier System	\$203,552	\$0	\$88,237	\$0	\$291,789	\$29,179	\$40,524	\$55,413	\$416,905	\$663
4.2	Syngas Cooler	w/4.1	\$0	w/ 4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.3	ASU & Oxidant Compression	\$177,729	\$0	w/equip.	\$0	\$177,729	\$17,773	\$0	\$19,550	\$215,052	\$342
4.4	LT Heat Recovery & FG Saturation	\$18,617	\$0	\$7,073	\$0	\$25,691	\$2,569	\$0	\$5,652	\$33,911	\$54
4.5	Misc. Gasification Equipment	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.6	Flare Stack System	\$0	\$1,110	\$449	\$0	\$1,559	\$156	\$0	\$343	\$2,058	\$3
4.8	Major Component Rigging	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.9	Gasification Foundations	\$0	\$9,370	\$5,588	\$0	\$14,957	\$1,496	\$0	\$4,113	\$20,567	\$33
	Subtotal	\$399,897	\$10,480	\$101,347	\$0	\$511,725	\$51,172	\$40,524	\$85,072	\$688,493	\$1,095
5A		Gas Cleanup & Piping									
5A.1	Sulfinol System	\$45,130	\$0	\$38,032	\$0	\$83,162	\$8,316	\$0	\$18,296	\$109,774	\$175
5A.2	Elemental Sulfur Plant	\$11,584	\$2,258	\$14,843	\$0	\$28,684	\$2,868	\$0	\$6,311	\$37,863	\$60
5A.3	Mercury Removal	\$1,400	\$0	\$1,058	\$0	\$2,458	\$246	\$123	\$565	\$3,392	\$5
5A.4	COS Hydrolysis	\$3,475	\$0	\$4,507	\$0	\$7,981	\$798	\$0	\$1,756	\$10,535	\$17
5A.5	Particulate Removal	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0

Case:		B1A – Shell IGCC w/o CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		629					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
5A.6	Blowback Gas Systems	\$1,945	\$327	\$183	\$0	\$2,455	\$245	\$0	\$540	\$3,240	\$5
5A.7	Fuel Gas Piping	\$0	\$979	\$640	\$0	\$1,619	\$162	\$0	\$356	\$2,137	\$3
5A.9	HGCU Foundations	\$0	\$886	\$597	\$0	\$1,483	\$148	\$0	\$489	\$2,120	\$3
	Subtotal	\$63,533	\$4,449	\$59,860	\$0	\$127,842	\$12,784	\$123	\$28,313	\$169,062	\$269
6	Combustion Turbine & Accessories										
6.1	Combustion Turbine Generator	\$104,201	\$0	\$7,506	\$0	\$111,707	\$11,171	\$5,585	\$12,846	\$141,309	\$225
6.9	Combustion Turbine Foundations	\$0	\$923	\$1,068	\$0	\$1,992	\$199	\$0	\$657	\$2,848	\$5
	Subtotal	\$104,201	\$923	\$8,575	\$0	\$113,699	\$11,370	\$5,585	\$13,504	\$144,158	\$229
7	HRSG, Ducting, & Stack										
7.1	Heat Recovery Steam Generator	\$31,100	\$0	\$6,022	\$0	\$37,122	\$3,712	\$0	\$4,083	\$44,918	\$71
7.3	Ductwork	\$0	\$2,085	\$1,461	\$0	\$3,546	\$355	\$0	\$780	\$4,681	\$7
7.4	Stack	\$4,023	\$0	\$1,501	\$0	\$5,524	\$552	\$0	\$608	\$6,684	\$11
7.9	HRSG, Duct & Stack Foundations	\$0	\$766	\$769	\$0	\$1,535	\$153	\$0	\$506	\$2,195	\$3
	Subtotal	\$35,123	\$2,851	\$9,753	\$0	\$47,727	\$4,773	\$0	\$5,978	\$58,477	\$93
8	Steam Turbine Generator										
8.1	Steam TG & Accessories	\$38,584	\$0	\$5,922	\$0	\$44,506	\$4,451	\$0	\$4,896	\$53,852	\$86
8.2	Turbine Plant Auxiliaries	\$241	\$0	\$549	\$0	\$791	\$79	\$0	\$87	\$957	\$2
8.3	Condenser & Auxiliaries	\$3,110	\$0	\$1,755	\$0	\$4,865	\$486	\$0	\$535	\$5,886	\$9
8.4	Steam Piping	\$16,344	\$0	\$7,088	\$0	\$23,432	\$2,343	\$0	\$6,444	\$32,218	\$51
8.9	TG Foundations	\$0	\$1,137	\$2,009	\$0	\$3,146	\$315	\$0	\$1,038	\$4,499	\$7
	Subtotal	\$58,279	\$1,137	\$17,322	\$0	\$76,739	\$7,674	\$0	\$13,000	\$97,412	\$155
9	Cooling Water System										
9.1	Cooling Towers	\$3,610	\$0	\$1,090	\$0	\$4,700	\$470	\$0	\$776	\$5,946	\$9
9.2	Circulating Water Pumps	\$1,751	\$0	\$120	\$0	\$1,871	\$187	\$0	\$309	\$2,367	\$4
9.3	Circ. Water System Auxiliaries	\$157	\$0	\$22	\$0	\$179	\$18	\$0	\$30	\$227	\$0
9.4	Circ. Water Piping	\$0	\$6,965	\$1,687	\$0	\$8,652	\$865	\$0	\$1,903	\$11,421	\$18
9.5	Make-up Water System	\$395	\$0	\$542	\$0	\$937	\$94	\$0	\$206	\$1,237	\$2
9.6	Component Cooling Water Sys	\$796	\$952	\$654	\$0	\$2,403	\$240	\$0	\$529	\$3,171	\$5
9.9	Circ. Water System Foundations	\$0	\$2,272	\$4,038	\$0	\$6,310	\$631	\$0	\$2,082	\$9,024	\$14
	Subtotal	\$6,709	\$10,190	\$8,153	\$0	\$25,052	\$2,505	\$0	\$5,834	\$33,392	\$53
10	Ash & Spent Sorbent Handling Systems										
10.1	Slag Dewatering & Cooling	\$18,726	\$0	\$9,172	\$0	\$27,898	\$2,790	\$0	\$3,069	\$33,756	\$54
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.6	Ash Storage Silos	\$635	\$0	\$686	\$0	\$1,321	\$132	\$0	\$218	\$1,671	\$3
10.7	Ash Transport & Feed Equipment	\$874	\$0	\$204	\$0	\$1,078	\$108	\$0	\$178	\$1,364	\$2
10.8	Misc. Ash Handling Equipment	\$1,315	\$1,612	\$478	\$0	\$3,405	\$341	\$0	\$562	\$4,307	\$7
10.9	Ash/Spent Sorbent Foundation	\$0	\$53	\$70	\$0	\$123	\$12	\$0	\$41	\$176	\$0
	Subtotal	\$21,550	\$1,665	\$10,610	\$0	\$33,826	\$3,383	\$0	\$4,067	\$41,275	\$66
11	Accessory Electric Plant										
11.1	Generator Equipment	\$1,155	\$0	\$1,125	\$0	\$2,280	\$228	\$0	\$251	\$2,758	\$4
11.2	Station Service Equipment	\$4,293	\$0	\$395	\$0	\$4,687	\$469	\$0	\$516	\$5,672	\$9

Case:		B1A – Shell IGCC w/o CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		629					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
11.3	Switchgear & Motor Control	\$7,923	\$0	\$1,473	\$0	\$9,396	\$940	\$0	\$1,550	\$11,885	\$19
11.4	Conduit & Cable Tray	\$0	\$4,026	\$12,409	\$0	\$16,435	\$1,643	\$0	\$4,520	\$22,598	\$36
11.5	Wire & Cable	\$0	\$7,762	\$4,722	\$0	\$12,484	\$1,248	\$0	\$3,433	\$17,165	\$27
11.6	Protective Equipment	\$0	\$797	\$2,958	\$0	\$3,755	\$375	\$0	\$620	\$4,750	\$8
11.7	Standby Equipment	\$275	\$0	\$274	\$0	\$548	\$55	\$0	\$90	\$693	\$1
11.8	Main Power Transformers	\$17,455	\$0	\$173	\$0	\$17,628	\$1,763	\$0	\$2,909	\$22,299	\$35
11.9	Electrical Foundations	\$0	\$181	\$493	\$0	\$674	\$67	\$0	\$222	\$963	\$2
Subtotal		\$31,100	\$12,765	\$24,020	\$0	\$67,885	\$6,789	\$0	\$14,110	\$88,784	\$141
12		Instrumentation & Control									
12.1	IGCC Control Equipment	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.2	Combustion Turbine Control	w/6.1	\$0	w/6.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.3	Steam Turbine Control	w/8.1	\$0	w/8.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.4	Other Major Component Control	\$1,195	\$0	\$814	\$0	\$2,010	\$201	\$100	\$347	\$2,658	\$4
12.5	Signal Processing Equipment	w/12.7	\$0	w/12.7	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.6	Control Boards, Panels & Racks	\$275	\$0	\$180	\$0	\$454	\$45	\$23	\$105	\$627	\$1
12.7	Computer & Accessories	\$6,376	\$0	\$208	\$0	\$6,585	\$658	\$329	\$757	\$8,330	\$13
12.8	Instrument Wiring & Tubing	\$0	\$2,455	\$4,646	\$0	\$7,101	\$710	\$355	\$2,042	\$10,208	\$16
12.9	Other I & C Equipment	\$4,262	\$0	\$2,112	\$0	\$6,374	\$637	\$319	\$1,100	\$8,430	\$13
Subtotal		\$12,109	\$2,455	\$7,961	\$0	\$22,524	\$2,252	\$1,126	\$4,350	\$30,252	\$48
13		Improvements to Site									
13.1	Site Preparation	\$0	\$117	\$2,670	\$0	\$2,788	\$279	\$0	\$920	\$3,986	\$6
13.2	Site Improvements	\$0	\$2,089	\$2,953	\$0	\$5,042	\$504	\$0	\$1,664	\$7,210	\$11
13.3	Site Facilities	\$3,743	\$0	\$4,202	\$0	\$7,945	\$794	\$0	\$2,622	\$11,361	\$18
Subtotal		\$3,743	\$2,206	\$9,825	\$0	\$15,774	\$1,577	\$0	\$5,206	\$22,557	\$36
14		Buildings & Structures									
14.1	Combustion Turbine Area	\$0	\$317	\$179	\$0	\$496	\$50	\$0	\$109	\$655	\$1
14.2	Steam Turbine Building	\$0	\$2,909	\$4,142	\$0	\$7,051	\$705	\$0	\$1,163	\$8,919	\$14
14.3	Administration Building	\$0	\$1,008	\$731	\$0	\$1,739	\$174	\$0	\$287	\$2,199	\$3
14.4	Circulation Water Pump House	\$0	\$199	\$105	\$0	\$304	\$30	\$0	\$50	\$384	\$1
14.5	Water Treatment Buildings	\$0	\$565	\$551	\$0	\$1,117	\$112	\$0	\$184	\$1,413	\$2
14.6	Machine Shop	\$0	\$516	\$353	\$0	\$869	\$87	\$0	\$143	\$1,099	\$2
14.7	Warehouse	\$0	\$833	\$537	\$0	\$1,370	\$137	\$0	\$226	\$1,733	\$3
14.8	Other Buildings & Structures	\$0	\$499	\$388	\$0	\$887	\$89	\$0	\$195	\$1,171	\$2
14.9	Waste Treating Building & Str.	\$0	\$1,115	\$2,130	\$0	\$3,245	\$325	\$0	\$714	\$4,283	\$7
Subtotal		\$0	\$7,961	\$9,116	\$0	\$17,077	\$1,708	\$0	\$3,072	\$21,857	\$35
Total		\$909,446	\$79,555	\$310,959	\$0	\$1,299,960	\$129,996	\$47,358	\$236,282	\$1,713,596	\$2,725

Exhibit 3-23 Case B1A owner's costs

Description	\$/1,000	\$/kW
Pre-Production Costs		
6 Months All Labor	\$15,220	\$24
1 Month Maintenance Materials	\$3,484	\$6
1 Month Non-fuel Consumables	\$371	\$1
1 Month Waste Disposal	\$405	\$1
25% of 1 Months Fuel Cost at 100% CF	\$2,731	\$4
2% of TPC	\$34,272	\$54
Total	\$56,482	\$90
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	\$22,284	\$35
0.5% of TPC (spare parts)	\$8,568	\$14
Total	\$30,852	\$49
Other Costs		
Initial Cost for Catalyst and Chemicals	\$2,032	\$3
Land	\$900	\$1
Other Owner's Costs	\$257,039	\$409
Financing Costs	\$46,267	\$74
Total Overnight Costs (TOC)	\$2,107,169	\$3,351
TASC Multiplier (IOU, high-risk, 35 year)	1.140	
Total As-Spent Cost (TASC)	\$2,402,172	\$3,820

Exhibit 3-24 Case B1A initial and annual operating and maintenance costs

Case:	B1A – Shell IGCC w/o CO ₂			Cost Base:	Jun 2011	
Plant Size (MW,net):	629	Heat Rate-net (Btu/kWh):	8,100	Capacity Factor (%):	80	
Operating & Maintenance Labor						
Operating Labor			Operating Labor Requirements per Shift			
Operating Labor Rate (base):	39.70	\$/hour	Skilled Operator:	2.0		
Operating Labor Burden:	30.00	% of base	Operator:	9.0		
Labor O-H Charge Rate:	25.00	% of labor	Foreman:	1.0		
			Lab Tech's, etc.:	3.0		
			Total:	15.0		
Fixed Operating Costs						
			Annual Cost			
			(\$)	(\$/kW-net)		
Annual Operating Labor:				\$6,781,554	\$10.783	
Maintenance Labor:				\$17,570,241	\$27.938	
Administrative & Support Labor:				\$6,087,949	\$9.680	
Property Taxes and Insurance:				\$34,271,927	\$54.495	
Total:				\$64,711,671	\$102.897	
Variable Operating Costs						
			(\$)	(\$/MWh-net)		
Maintenance Material:				\$33,446,804	\$7.58890	
Consumables						
	Consumption				Cost (\$)	
	Initial Fill	Per Day	Per Unit	Initial Fill		
Water (/1000 gallons):	0	2,988	\$1.67	\$0	\$1,460,528	\$0.33139
Makeup and Waste Water Treatment Chemicals (lbs):	0	17,801	\$0.27	\$0	\$1,392,224	\$0.31589
Carbon (Mercury Removal) (lb):	116,013	199	\$5.50	\$638,069	\$319,034	\$0.07239
COS Catalyst (m³):	262	0.18	\$3,751.70	\$982,135	\$196,427	\$0.04457
Sulfinol Solution (gal):	26,146	18	\$15.73	\$411,337	\$81,630	\$0.01852
Claus Catalyst (ft³):	w/equip.	1.82	\$203.15	\$0	\$108,246	\$0.02456
Subtotal:				\$2,031,540	\$3,558,089	\$0.80731
Waste Disposal						
Spent Mercury Catalyst (lb.):	0	199	\$0.65	\$0	\$37,710	\$0.00856
Flyash (ton):	0	0	\$0.00	\$0	\$0	\$0.00000
Slag (ton):	0	525	\$25.11	\$0	\$3,847,802	\$0.87305
Subtotal:				\$0	\$3,885,512	\$0.88160
By-Products						
Sulfur (tons):	0	131	\$0.00	\$0	\$0	\$0.00000
Subtotal:				\$0	\$0	\$0.00000
Variable Operating Costs Total:				\$2,031,540	\$40,890,405	\$9.27782
Fuel Cost						
Illinois Number 6 (ton):	0	5,240	\$68.54	\$0	\$104,866,703	\$23.79370
Total:				\$0	\$104,866,703	\$23.79370

Exhibit 3-25 Case B1A COE breakdown

Component	Value, \$/MWh	Percentage
Capital	59.3	55%
Fixed	14.7	14%
Variable	9.3	9%
Fuel	23.8	22%
Total (Excluding T&S)	107.0	N/A
CO ₂ T&S	0.0	0%
Total (Including T&S)	107.0	N/A

3.2.8 Case B1B - Shell IGCC Power Plant with CO₂ Capture

This case is configured to produce electric power with CO₂ capture. The plant configuration is the same as Case B1A, namely two Shell gasifier trains, two state-of-the-art 2008 F-class turbines, two HRSGs, and one steam turbine. The gross power output is constrained by the capacity of the two CTs, and since the CO₂ capture and compression process increases the auxiliary load on the plant, the net output is significantly reduced relative to Case B1A (497 MW versus 629 MW).

The process description for Case B1B is similar to Case B1A with several notable exceptions to accommodate CO₂ capture. A BFD and stream tables for Case B1B are shown in Exhibit 3-26 and Exhibit 3-27, respectively. Instead of repeating the entire process description, only differences from Case B1A are reported here.

3.2.8.1 Coal Preparation and Feed Systems

No differences from Case B1A.

3.2.8.2 Gasification

The gasification process is the same as Case B1A with the exception that total coal feed to the two gasifiers is 5,065 tonnes/day (5,583 tpd) (stream 9) and the ASU provides 3,852 tonnes/day (4,246 tpd) of 95 mol% oxygen to the gasifier and Claus plant (streams 5 and 3).

3.2.8.3 Raw Gas Cooling/Particulate Removal

After the raw syngas is cooled to approximately 1,093°C (2,000°F) by the syngas quench recycle, syngas is further cooled to 899°C (1,650°F) by raising HP steam at 13.8 MPa (2,000 psia). A direct contact water quench follows to cool the raw syngas from 899°C (1,650°F) to 399°C (750°F) while providing a portion of the water required for SGS. The syngas is then cooled to 316°C (600°F) by raising steam, which is used in the SGS unit. After particulate filtration, the syngas is cooled to 232°C (450°F) by raising IP steam at 0.4 MPa (65 psia) before proceeding to the scrubber.

3.2.8.4 Syngas Scrubber/Sour Water Stripper

Syngas exits the scrubber at 191°C (376°F).

3.2.8.5 Sour Gas Shift

The SGS process was described in Section 3.1.3. In Case B1B the syngas after the scrubber is reheated to 197°C (386°F) and then steam (stream 14) is added to adjust the H₂O:CO molar ratio to approximately 1.8:1 prior to the first SGS reactor. The hot syngas exiting the first stage of SGS is used to preheat water used to humidify clean syngas prior to entering the CT. One more stage of SGS (for a total of two) results in 97.8 percent overall conversion of the CO to CO₂. The warm syngas from the second stage of SGS is cooled to 248°C (478°F) by preheating the syngas prior to the first stage of SGS. The SGS catalyst also serves to hydrolyze COS thus eliminating the need for a separate COS hydrolysis reactor. Following the second stage of SGS, the syngas is further cooled to 35°C (95°F) prior to the mercury removal beds.

3.2.8.6 Mercury Removal and AGR

Mercury removal is the same as in Case B1A.

The AGR process in Case B1B is a two-stage Selexol process where H₂S is removed in the first stage and CO₂ in the second stage of absorption. The process results in three product streams, the clean syngas (stream 18), a CO₂-rich stream and an acid gas feed to the Claus plant (stream 22). The acid gas contains 34 percent H₂S and 51 percent CO₂ with the balance primarily H₂. The CO₂-rich stream is discussed further in the CO₂ compression section.

3.2.8.7 CO₂ Compression and Dehydration

CO₂ from the AGR process is flashed at three pressure levels to separate CO₂ and decrease H₂ losses to the CO₂ product pipeline. The HP CO₂ stream is flashed at 2.0 MPa (289.7 psia), compressed, and recycled back to the CO₂ absorber. The MP CO₂ stream is flashed at 1.0 MPa (149.7 psia). The LP CO₂ stream is flashed at 0.1 MPa (16.7 psia), compressed to 1.0 MPa (149.5 psia), and combined with the MP CO₂ stream. The combined stream is compressed from 1.0 MPa (149.5 psia) to an SC condition at 15.3 MPa (2215 psia) using a multiple-stage, intercooled compressor. During compression, the CO₂ stream is dehydrated to a dew point of -40°C (-40°F) with triethylene glycol. The raw CO₂ stream from the Selexol process contains over 99 percent CO₂. The CO₂ (stream 21) is transported to the plant fence line and is sequestration ready. CO₂ T&S costs were estimated using the methodology described in Section 2.6.

3.2.8.8 Claus Unit

The Claus plant is the same as Case B1A with the following exceptions:

- 5,277 kg/hr (11,634 lb/hr) of sulfur (stream 23) are produced
- The waste heat boiler generates 13,697 kg/hr (30,197 lb/hr) of 3.0 MPa (430 psia) steam, which provides all of the Claus plant process needs and provides some additional steam to the medium pressure steam header.

3.2.8.9 Power Block

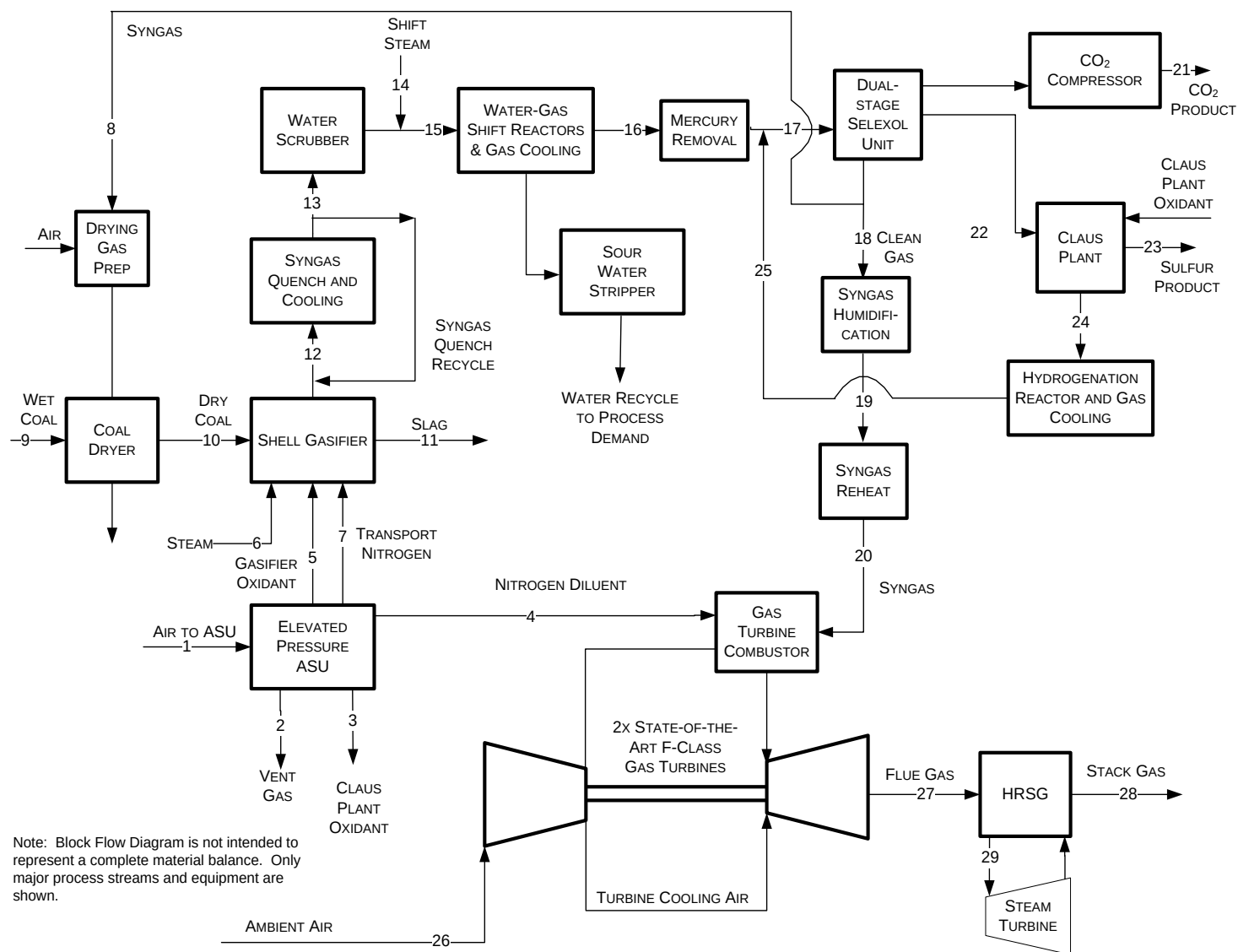
Clean syngas from the AGR plant is combined with a small amount of clean gas from the CO₂ compression process (stream 18) and partially humidified because the nitrogen available from the ASU is insufficient to provide adequate dilution. The moisturized syngas is reheated to 193°C (380°F) using HP BFW, diluted with nitrogen (stream 4), and then enters the CT burner. The exhaust gas (stream 27) exits the CT at 562°C (1,043°F) and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 132°C (270°F) (stream 28) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available steam turbine using a 12.4 MPa/534°C/534°C (1,800 psig/993°F/993°F) steam cycle.

3.2.8.10 Air Separation Unit (ASU)

The same elevated pressure ASU is used as in Case B1A and produces 3,852 tonnes/day (4,246 tpd) of 95 mol% oxygen and 12,290 tonnes/day (13,547 tpd) of nitrogen. There is no integration between the ASU and the CT in the CO₂ capture case.

3.2.8.11 Balance of Plant

Balance of plant items were covered in Sections 3.1.9, 3.1.10, and 3.1.12.

Exhibit 3-26 Case B1B block flow diagram, Shell IGCC with CO₂ capture

Source: NETL

Exhibit 3-27 Case B1B stream table, Shell IGCC with capture

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0230	0.0318	0.0023	0.0318	0.0000	0.0023	0.0096	0.0000	0.0000	0.0000	0.0085	0.0061	0.0000
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000	0.0005	0.0004	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0130	0.0000	0.0000	0.0000	0.5187	0.3712	0.0000
CO ₂	0.0003	0.0079	0.0000	0.0000	0.0000	0.0000	0.0000	0.0455	0.0000	0.0000	0.0000	0.0126	0.0090	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0004	0.0000
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.8727	0.0000	0.8090	0.0000	0.2691	0.1926	0.0000
H ₂ O	0.0099	0.1984	0.0000	0.0003	0.0000	1.0000	0.0004	0.0001	0.0000	0.0935	0.0000	0.1278	0.3758	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0030	0.0000	0.0008	0.0006	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0052	0.0000
N ₂	0.7732	0.5759	0.0178	0.9919	0.0178	0.0000	0.9919	0.0585	0.0000	0.0163	0.0000	0.0507	0.0363	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	0.0033	0.0024	0.0000
O ₂	0.2074	0.1948	0.9504	0.0054	0.9504	0.0000	0.0054	0.0000	0.0000	0.0782	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	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	24,277	1,004	102	17,310	4,886	1,130	940	308	0	5,809	0	26,493	29,249	11,003
V-L Flowrate (kg/hr)	700,548	27,328	3,289	485,713	157,225	20,363	26,368	1,900	0	37,081	0	531,672	569,894	198,214
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	211,040	160,281	21,137	0	0	0
Temperature (°C)	15	20	32	93	32	343	32	35	15	16	1,427	1,082	232	288
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	5.10	5.62	3.59	0.10	5.79	4.24	4.24	4.03	5.52
Steam Table Enthalpy (kJ/kg) ^A	30.23	36.90	26.67	92.51	26.67	3,063.97	20.78	163.14	---	---	---	2,047.11	1,220.20	2,918.18
AspenPlus Enthalpy (kJ/kg) ^B	-97.58	-2,197.43	4.13	63.47	4.13	-12,907.38	-8.79	-3,104.81	-2,116.95	-768.33	2,165.43	-2,946.51	-6,659.43	-13,053.16
Density (kg/m ³)	1.2	1.5	11.0	24.4	11.0	20.1	62.8	8.5	---	16.5	---	7.5	19.1	25.6
V-L Molecular Weight	28.857	27.206	32.181	28.060	32.181	18.015	28.060	6.159	---	6.383	---	20.069	19.484	18.015
V-L Flowrate (lb _{mole} /hr)	53,521	2,215	225	38,162	10,771	2,492	2,072	680	0	12,807	0	58,407	64,482	24,256
V-L Flowrate (lb/hr)	1,544,445	60,248	7,250	1,070,814	346,622	44,894	58,132	4,188	0	81,749	0	1,172,137	1,256,401	436,986
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	465,264	353,359	46,599	0	0	0
Temperature (°F)	59	68	90	199	90	650	90	94	59	60	2,600	1,980	450	550
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	740.0	815.0	520.0	14.7	840.0	615.0	615.0	585.0	800.0
Steam Table Enthalpy (Btu/lb) ^A	13.0	15.9	11.5	39.8	11.5	1,317.3	8.9	70.1	---	---	---	880.1	524.6	1,254.6
AspenPlus Enthalpy (Btu/lb) ^B	-42.0	-944.7	1.8	27.3	1.8	-5,549.2	-3.8	-1,334.8	-910.1	-330.3	931.0	-1,266.8	-2,863.0	-5,611.9
Density (lb/ft ³)	0.076	0.097	0.687	1.521	0.687	1.257	3.918	0.531	---	1.029	---	0.468	1.190	1.597

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 3-27 Case B1B stream table, Shell IGCC with capture (continued)

	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
V-L Mole Fraction															
Ar	0.0047	0.0062	0.0062	0.0096	0.0086	0.0086	0.0002	0.0017	0.0000	0.0065	0.0102	0.0092	0.0088	0.0088	0.0000
CH ₄	0.0003	0.0004	0.0004	0.0006	0.0005	0.0005	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.2873	0.0084	0.0084	0.0130	0.0116	0.0116	0.0002	0.0025	0.0000	0.1143	0.0060	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0070	0.3776	0.3810	0.0455	0.0407	0.0407	0.9944	0.5141	0.0000	0.2866	0.6257	0.0003	0.0080	0.0080	0.0000
COS	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.1491	0.5633	0.5593	0.8727	0.7819	0.7819	0.0050	0.1068	0.0000	0.0659	0.2694	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.5172	0.0016	0.0016	0.0001	0.1041	0.1041	0.0000	0.0300	0.0000	0.4715	0.0017	0.0099	0.1374	0.1374	1.0000
HCl	0.0001	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
H ₂ S	0.0040	0.0057	0.0057	0.0000	0.0000	0.0000	0.0000	0.3398	0.0000	0.0014	0.0057	0.0000	0.0000	0.0000	0.0000
N ₂	0.0281	0.0368	0.0374	0.0585	0.0524	0.0524	0.0002	0.0047	0.0000	0.0516	0.0813	0.7732	0.7397	0.7397	0.0000
NH ₃	0.0019	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	0.2074	0.1060	0.1060	0.1060	0.0000
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0018	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	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	37,783	28,837	29,237	18,312	20,438	20,438	10,099	488	0	630	400	110,253	139,892	139,892	28,855
V-L Flowrate (kg/hr)	723,426	562,223	574,690	112,786	151,092	151,092	442,270	17,202	0	16,585	12,466	3,181,557	3,818,362	3,818,362	519,836
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	5,277	0	0	0	0	0	0
Temperature (°C)	217	35	35	35	137	193	51	48	178	232	38	15	562	132	534
Pressure (MPa, abs)	3.96	3.62	3.59	3.59	3.21	3.17	15.27	0.16	0.12	0.08	5.51	0.10	0.10	0.10	12.51
Steam Table Enthalpy (kJ/kg) ^A	1,541.73	45.33	44.53	163.14	1,181.24	1,416.72	-161.70	85.01	---	1,114.50	8.30	30.23	866.14	371.29	3,432.00
AspenPlus Enthalpy (kJ/kg) ^B	-8,063.86	-7,695.78	-7,701.53	-3,104.81	-5,307.37	-5,071.89	-9,118.58	-6,131.59	142.40	-8,838.52	-7,960.81	-97.58	-720.70	-1,215.55	-12,538.31
Density (kg/m ³)	19.5	28.1	28.1	8.5	6.9	6.0	639.8	2.2	5,279.4	0.5	74.4	1.2	0.4	0.8	36.8
V-L Molecular Weight	19.147	19.496	19.656	6.159	7.393	7.393	43.792	35.245	---	26.336	31.160	28.857	27.295	27.295	18.015
V-L Flowrate (lb _{mole} /hr)	83,298	63,576	64,458	40,371	45,059	45,059	22,265	1,076	0	1,388	882	243,066	308,408	308,408	63,615
V-L Flowrate (lb/hr)	1,594,881	1,239,491	1,266,974	248,650	333,100	333,100	975,038	37,925	0	36,563	27,484	7,014,133	8,418,047	8,418,047	1,146,042
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	11,634	0	0	0	0	0	0
Temperature (°F)	422	95	94	94	278	380	124	119	352	450	100	59	1,043	270	993
Pressure (psia)	575.0	525.0	520.0	520.0	465.0	460.0	2,214.7	23.7	17.3	12.3	799.5	14.7	15.2	15.2	1,814.7
Steam Table Enthalpy (Btu/lb) ^A	662.8	19.5	19.1	70.1	507.8	609.1	-69.5	36.5	---	479.1	3.6	13.0	372.4	159.6	1,475.5
AspenPlus Enthalpy (Btu/lb) ^B	-3,466.8	-3,308.6	-3,311.1	-1,334.8	-2,281.8	-2,180.5	-3,920.3	-2,636.1	61.2	-3,799.9	-3,422.5	-42.0	-309.8	-522.6	-5,390.5
Density (lb/ft ³)	1.218	1.753	1.753	0.531	0.431	0.374	39.942	0.135	329.584	0.033	4.646	0.076	0.026	0.053	2.295

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

3.2.9 Case B1B - Performance Results

The Case B1B modeling assumptions were presented previously in Section 3.2.3.

The plant produces a net output of 497 MWe at a net plant efficiency of 31.2 percent (HHV basis). Overall performance for the plant is summarized in Exhibit 3-28; Exhibit 3-29 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for approximately 58 percent of the auxiliary load between the main air compressor, nitrogen compressor, oxygen compressor, and ASU auxiliaries. The two-stage Selexol process and CO₂ compression account for an additional 28 percent of the auxiliary power load. The BFW and CWS (CWP and cooling tower fan) compose approximately 4 percent of the load, leaving 10 percent of the auxiliary load for all other systems.

Exhibit 3-28 Case B1B plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	209
Total Gross Power, MWe	673
Air Separation Unit Main Air Compressor, kWe	59,740
Oxygen Compressor, kWe	9,460
Nitrogen Compressors, kWe	32,910
CO ₂ Compression, kWe	30,210
Acid Gas Removal, kWe	18,650
Balance of Plant, kWe	25,690
Total Auxiliaries, MWe	177
Net Power, MWe	497
HHV Net Plant Efficiency (%)	31.2%
HHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	11,528 (10,927)
HHV Cold Gas Efficiency, %	84.5%
HHV Combustion Turbine Efficiency, %	35.9%
LHV Net Plant Efficiency (%)	32.4%
LHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	11,119 (10,539)
LHV Cold Gas Efficiency, %	82.9%
LHV Combustion Turbine Efficiency, %	42.3%
Steam Turbine Cycle Efficiency, %	39.0%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	9,238 (8,756)
Condenser Duty, GJ/hr (MMBtu/hr)	1,335 (1,265)
As-Received Coal Feed, kg/hr (lb/hr)	211,040 (465,264)
HHV Thermal Input, kWt	1,590,722
LHV Thermal Input, kWt	1,534,271
Raw Water Withdrawal, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.043 (11.4)
Raw Water Consumption, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.035 (9.3)
O ₂ :Coal	0.704

Exhibit 3-29 Case B1B plant power summary

Power Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	209
Total Gross Power, MWe	673
Auxiliary Load Summary	
Coal Handling, kWe	460
Coal Milling, kWe	2,170
Sour Water Recycle Slurry Pump, kWe	0
Slag Handling, kWe	550
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	59,740
Oxygen Compressor, kWe	9,460
Nitrogen Compressors, kWe	32,910
Feedwater Pumps, kWe	3,500
Quench Water Pump, kWe	610
Syngas Recycle Compressor, kWe	790
CO ₂ Compression, kWe	30,210
Scrubber Pumps, kWe	360
Acid Gas Removal, kWe	18,650
Combustion Turbine Auxiliaries, kWe	1,000
Steam Turbine Auxiliaries, kWe	100
Condensate Pumps, kWe	280
Circulating Water Pumps, kWe	4,380
Ground Water Pumps, kWe	510
Cooling Tower Fans, kWe	2,270
Claus Plant/TGTU Auxiliaries, kWe	250
Claus Plant TG Recycle Compressor, kWe	1,830
Miscellaneous Balance of Plant ^A , kWe	3,000
Transformer Losses, kWe	2,630
Total Auxiliaries, MWe	177
Net Power, MWe	497

^AIncludes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.2.9.1 Environmental Performance

The environmental targets for emissions of Hg, NO_x, SO₂, CO₂, and PM were presented in Section 2.4. A summary of the plant air emissions for Case B1B is presented in Exhibit 3-30. All HCl is assumed to be removed and is, therefore, not reported.

Exhibit 3-30 Case B1B air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year) ^A	kg/MWh (lb/MWh) ^B
SO ₂	0.001 (0.002)	37 (40)	0.008 (0.017)
NO _x	0.021 (0.049)	847 (934)	0.180 (0.396)
Particulates	0.003 (0.0071)	123 (135)	0.026 (0.057)
Hg	1.60E-7 (3.72E-7)	0.006 (0.007)	1.36E-6 (3.00E-6)
CO ₂	9 (20)	344,507 (379,754)	73 (161)
CO ₂ ^C	-	-	99 (218)

^ACalculations based on an 80 percent capacity factor

^BEmissions are based on gross power except where otherwise noted

^CCO₂ emissions based on net power instead of gross power

The low level of SO₂ emissions is achieved by capturing the sulfur in the gas by the two-stage Selexol AGR process. The CO₂ capture target results in the sulfur compounds being removed to a greater extent than required in the environmental targets of Section 2.3. The clean syngas exiting the AGR process has a sulfur concentration of approximately 4 ppmv. This results in a concentration in the HRSG flue gas of less than 1 ppmv. The H₂S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is compressed and recycled back to the AGR where most of the remaining sulfur is removed.

NO_x emissions are limited by the use of nitrogen dilution and humidification to 15 ppmvd (as NO₂ @ 15 percent O₂). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process and subsequently destroyed in the Claus plant burner. This helps lower NO_x levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a cyclone and a barrier filter in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of mercury is captured from the syngas by dual activated carbon beds.

Ninety two percent of the CO₂ from the syngas is captured in the AGR system and compressed for sequestration. Because not all CO is converted to CO₂ in the shift reactors, the overall carbon removal is 90.0 percent.

The carbon balance for the plant is shown in Exhibit 3-31. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon leaves the plant as unburned carbon in the slag and as CO₂ in the stack gas, coal dryer vent gas, ASU vent gas, and the captured CO₂ product. The carbon capture efficiency is defined as one minus the amount of carbon in the stack gas relative to the total carbon in less carbon contained in the slag, represented by the following fraction:

$$\frac{\text{Carbon in Stack}}{(\text{Total Carbon In}) - (\text{Carbon in Slag})} = \left(1 - \left(\frac{29,578}{297,754 - 1,483}\right) * \right) 100 = 90.0\%$$

Exhibit 3-31 Case B1B carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	134,527 (296,581)	Stack Gas	13,416 (29,578)
Air (CO ₂)	532 (1,173)	Coal Dryer Stack	223 (491)
		CO ₂ Product	120,652 (265,992)
		ASU Vent	95 (210)
		Slag	673 (1,483)
Total	135,059 (297,754)	Total	135,059 (297,754)

Exhibit 3-32 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, and sulfur from the tail gas unit that is vented through the coal dryer. Sulfur in the slag is considered negligible.

Exhibit 3-32 Case B1B sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,290 (11,661)	Stack Gas	3 (6)
		Coal Dryer Stack	0 (0)
		CO ₂ Product	10 (22)
		Elemental Sulfur	5,277 (11,634)
Total	5,290 (11,661)	Total	5,290 (11,661)

Exhibit 3-33 shows the overall water balance for the plant. The exhibit is presented in an identical manner as for Case B1A.

Exhibit 3-33 Case B1B water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)
Slag Handling	0.46 (121)	0.46 (121)	–	–	–
Slurry Water	–	–	–	–	–
Quench/Wash	3.17 (836)	2.22 (587)	0.94 (250)	–	0.94 (250)
Humidifier	0.67 (177)	0.67 (177)	–	–	–
SWS Blowdown	–	–	–	0.03 (7.08)	-0.03 (-7.08)
Condenser Makeup	3.83 (1,012)	–	3.83 (1,012)	–	3.83 (1,012)
BFW Makeup	0.18 (48)	–	0.18 (48)	–	0.18 (48)
Gasifier Steam	0.34 (90)	–	0.34 (90)	–	0.34 (90)
Shift Steam	3.31 (874)	–	3.31 (874)	–	3.31 (874)
CT Steam Dilution	–	–	–	–	–
Cooling Tower	17.07 (4,510)	0.45 (119)	16.62 (4,391)	3.84 (1,014)	12.78 (3,377)
BFW Blowdown	–	0.18 (48)	-0.18 (-48)	–	-0.18 (-48)
SWS Blowdown	–	0.27 (70.80)	-0.27 (-70.80)	–	-0.27 (-70.80)
Total	25.20 (6,657)	3.80 (1,004)	21.40 (5,652)	3.87 (1,021)	17.53 (4,631)

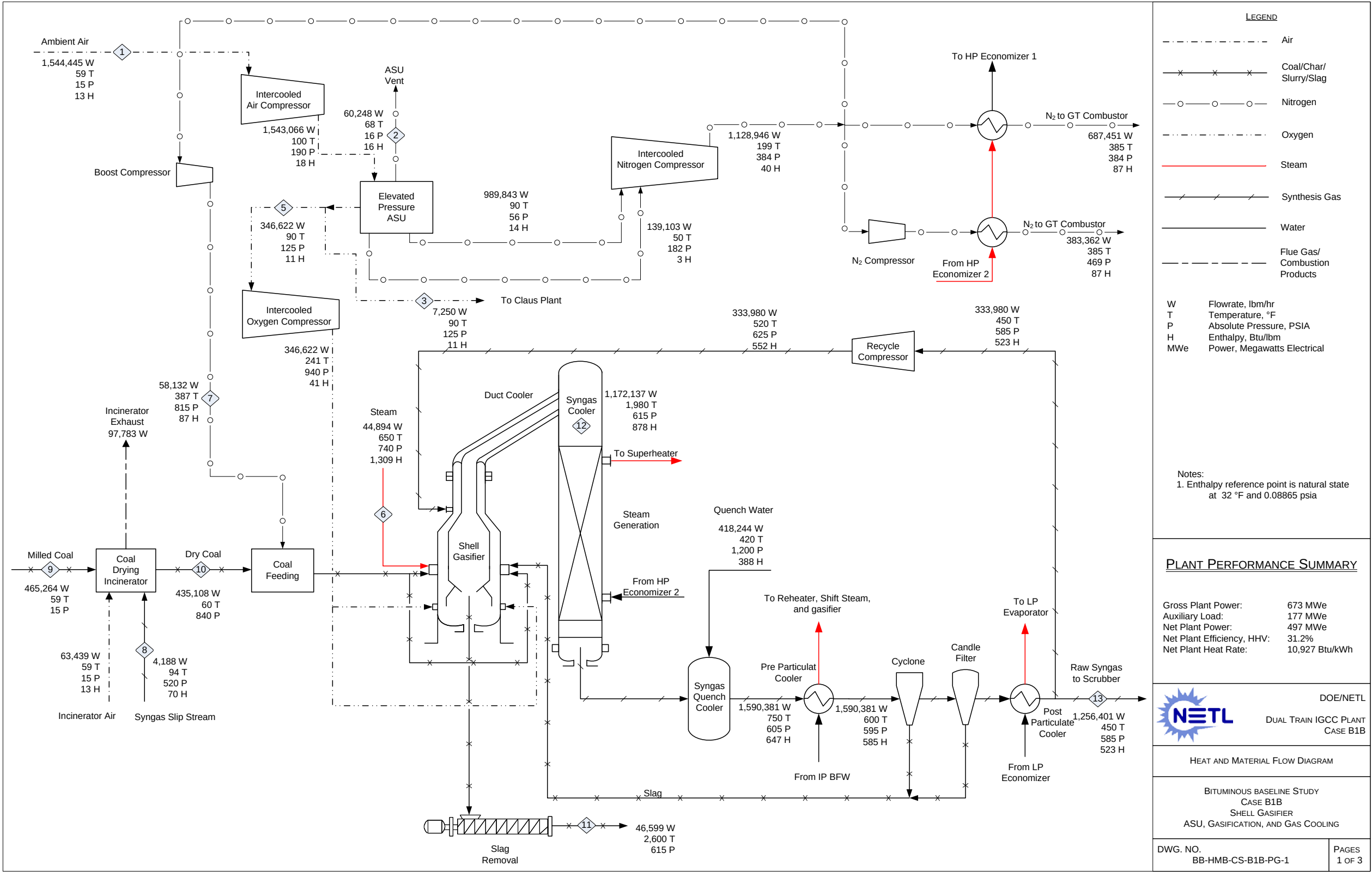
3.2.9.2 Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-34 through Exhibit 3-36:

- Coal gasification and ASU
- Syngas cleanup including sulfur recovery and tail gas recycle
- Combined cycle power generation, steam, and FW

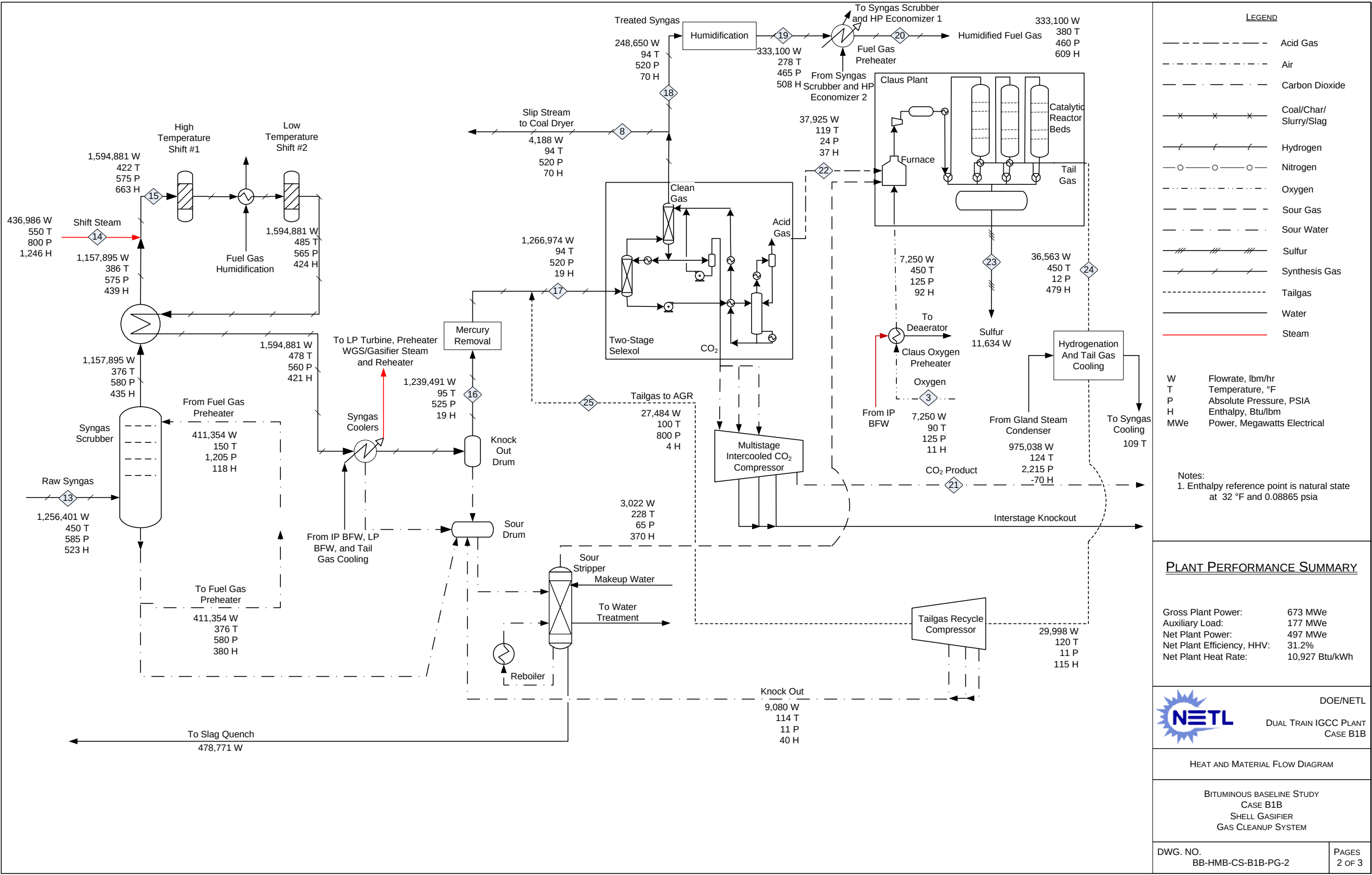
An overall plant energy balance is provided in tabular form in Exhibit 3-37. The power out is the combined CT and steam turbine power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-28) is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

Exhibit 3-34 Case B1B coal gasification and ASU heat and mass balance schematic



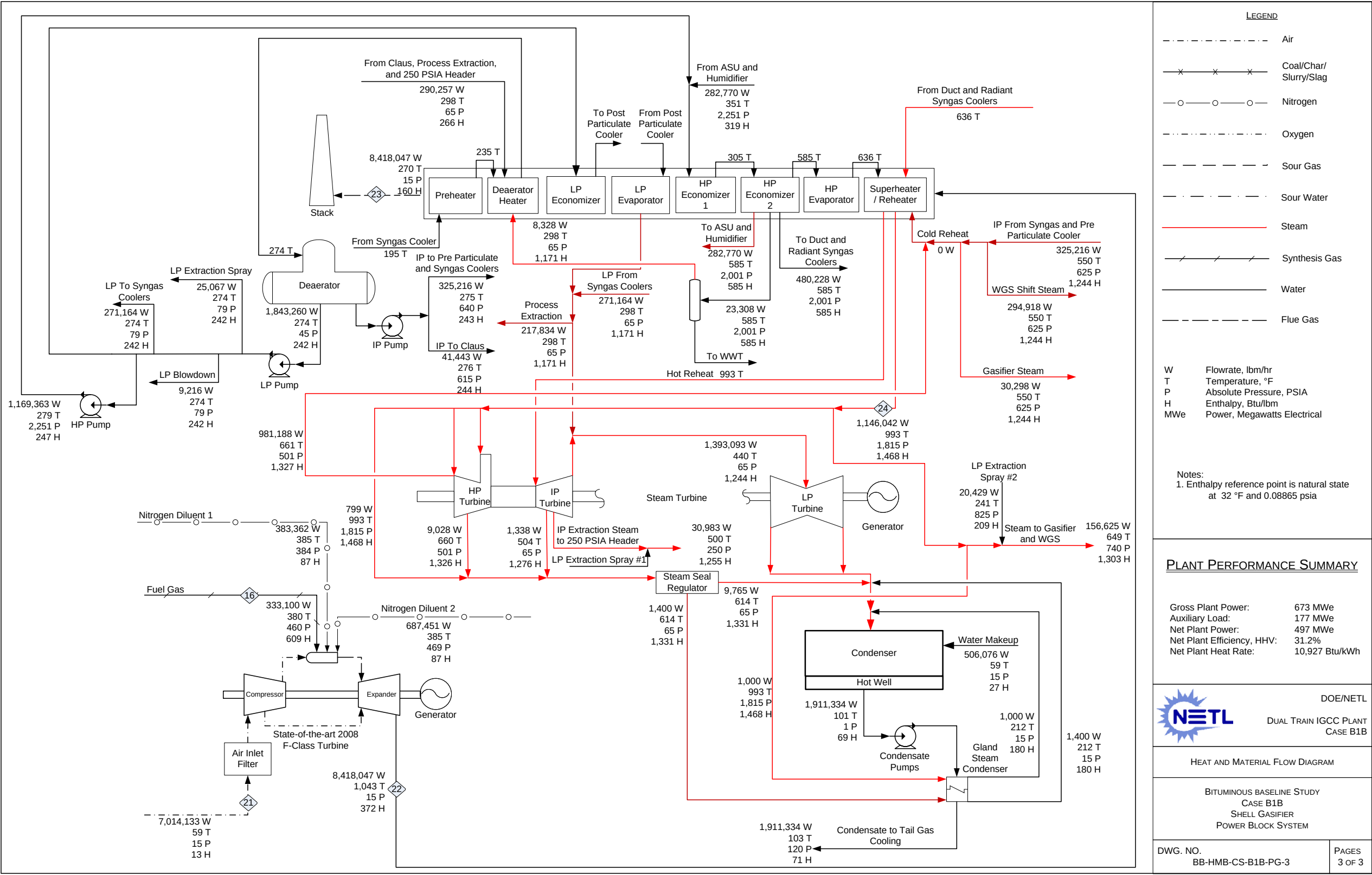
Source: NETL

Exhibit 3-35 Case B1B syngas cleanup heat and mass balance schematic



Source: NETL

Exhibit 3-36 Case B1B combined cycle power generation heat and mass balance schematic



Source: NETL

This page intentionally left blank

Exhibit 3-37 Case B1B overall energy balance (0°C [32°F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In (GJ/hr)				
Coal	5,727 (5,428)	4.8 (4.5)	–	5,731 (5,432)
Air	–	118.2 (112.1)	–	118.2 (112.1)
Raw Water Makeup	–	80.5 (76.3)	–	80.5 (76.3)
Auxiliary Power	–	–	636 (603)	636 (603)
TOTAL	5,727 (5,428)	203.5 (192.9)	636 (603)	6,566 (6,223)
Heat Out (GJ/hr)				
ASU Vent	–	1.0 (1.0)	–	1.0 (1.0)
Slag	22.1 (20.9)	35.7 (33.8)	–	57.7 (54.7)
Stack Gas	–	1,418 (1,344)	–	1,418 (1,344)
Sulfur	48.9 (46.3)	0.6 (0.6)	–	49.5 (46.9)
Motor Losses and Design Allowances	–	–	55.5 (52.6)	55.5 (52.6)
Condenser	–	1,335 (1,265)	–	1,335 (1,265)
Non-Condenser Cooling Tower Loads	–	724 (686)	–	724 (686)
CO ₂	–	-71.5 (-67.8)	–	-71.5 (-67.8)
Blowdown	–	41.3 (39.1)	–	41.3 (39.1)
CO ₂ Capture Losses	–	171.0 (162.1)	–	171.0 (162.1)
<i>Ambient Losses^A</i>	–	<i>115.3 (109.3)</i>	–	<i>115.3 (109.3)</i>
Power	–	–	2,424 (2,298)	2,424 (2,298)
TOTAL	71.0 (67.2)	3,769 (3,573)	2,480 (2,350)	6,320 (5,990)
Unaccounted Energy ^B	–	246 (233)	–	246 (233)

^AAmbient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers.

^BBy difference

3.2.10 Case B1B - Major Equipment List

Major equipment items for the Shell gasifier with CO₂ capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost

estimates in Section 3.2.11. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

Case B1B – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	40 tonne (50 ton)	2	1
7	Feeder	Vibratory	170 tonne/hr (190 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	350 tonne/hr (380 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	170 tonne (190 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/ tripper	350 tonne/hr (380 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	350 tonne/hr (380 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	800 tonne (900 ton)	3	0

Case B1B – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/hr (90 tph)	3	0
2	Conveyor No. 6	Belt w/ tripper	230 tonne/hr (260 tph)	1	0
3	Roller Mill Feed Hopper	Dual Outlet	460 tonne (510 ton)	1	0
4	Weigh Feeder	Belt	120 tonne/hr (130 tph)	2	0
5	Coal Dryer and Pulverizer	Rotary	120 tonne/hr (130 tph)	2	0
6	Coal Dryer Feed Hopper	Vertical Hopper	230 tonne (260 ton)	2	0

Case B1B – Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	987,000 liters (261,000 gal)	2	0
2	Condensate Pumps	Vertical canned	8,000 lpm @ 90 m H ₂ O (2,110 gpm @ 300 ft H ₂ O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	551,000 kg/hr (1,215,000 lb/hr)	2	0
4	Intermediate-Pressure Feedwater Pump	Horizontal centrifugal, single stage	7,970 lpm @ 30 m H ₂ O (2,110 gpm @ 90 ft H ₂ O)	2	1
5	High-Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 5,060 lpm @ 1,800 m H ₂ O (1,340 gpm @ 6,100 ft H ₂ O)	2	1
6	High-Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,220 lpm @ 220 m H ₂ O (320 gpm @ 730 ft H ₂ O)	2	1
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded Screw	28 m ³ /min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m ³ /min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	398 GJ/hr (377 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	142,800 lpm @ 20 m H ₂ O (37,700 gpm @ 70 ft H ₂ O)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H ₂ O (1,000 gpm @ 350 ft H ₂ O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H ₂ O (700 gpm @ 250 ft H ₂ O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	5,950 lpm @ 20 m H ₂ O (1,570 gpm @ 60 ft H ₂ O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	2,980 lpm @ 270 m H ₂ O (790 gpm @ 880 ft H ₂ O)	4	1
16	Filtered Water Pumps	Stainless steel, single suction	4,640 lpm @ 50 m H ₂ O (1,230 gpm @ 160 ft H ₂ O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	2,226,000 liter (588,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	2,090 lpm (550 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0

Case B1B – Account 4: Gasifier, ASU, and Accessories Including Low Temperature Heat Recovery and Fuel Gas Saturation

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized dry-feed, entrained bed	2,800 tonne/day, 4.2 MPa (3,100 tpd, 615 psia)	2	0
2	Synthesis Gas Cooler	Convective spiral-wound tube boiler	397,000 kg/hr (875,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	313,000 kg/hr (691,000 lb/hr) Design efficiency 90%	2	0
4	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0
5	Syngas Scrubber Including Sour Water Stripper	Vertical upflow	313,000 kg/hr (691,000 lb/hr)	2	0
6	Raw Gas Coolers	Shell and tube with condensate drain	398,000 kg/hr (877,000 lb/hr)	8	0
7	Raw Gas Knockout Drum	Vertical with mist eliminator	310,000 kg/hr, 35°C, 3.7 MPa (683,000 lb/hr, 95°F, 530 psia)	2	0
8	Saturation Water Economizers	Shell and tube	78 GJ/hr (74 MMBtu/hr)	2	0
9	Fuel Gas Saturator	Vertical tray tower	83,000 kg/hr, 137°C, 3.6 MPa (183,000 lb/hr, 278°F, 520 psia)	2	0
10	Saturator Water Pump	Centrifugal	1,700 lpm @ 13 m H ₂ O (500 gpm @ 42 ft H ₂ O)	2	2
11	Synthesis Gas Reheater	Shell and tube	83,000 kg/hr (183,000 lb/hr)	2	0
12	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	313,000 kg/hr (691,000 lb/hr) syngas	2	0
13	ASU Main Air Compressor	Centrifugal, multi-stage	5,000 m ³ /min @ 1.3 MPa (186,000 scfm @ 190 psia)	2	0
14	Cold Box	Vendor design	2,100 tonne/day (2,300 tpd) of 95% purity oxygen	2	0
15	Oxygen Compressor	Centrifugal, multi-stage	1,000 m ³ /min (37,000 scfm) Suction - 1.0 MPa (130 psia) Discharge - 6.5 MPa (940 psia)	2	0
16	Primary Nitrogen Compressor	Centrifugal, multi-stage	3,000 m ³ /min (123,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
17	Secondary Nitrogen Compressor	Centrifugal, single-stage	490 m ³ /min (17,000 scfm) Suction - 1.3 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0
18	Gasifier Purge Nitrogen Boost Compressor	Centrifugal, single-stage	1,000 m ³ /min (48,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 3.2 MPa (470 psia)	2	0

Case B1B – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	309,000 kg/hr (682,000 lb/hr), 35°C (95°F), 3.6 MPa (525 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	309,000 kg/hr (682,000 lb/hr), 35°C (95°F), 3.6 MPa (520 psia)	2	0
3	Sulfur Plant	Claus type	139 tonne/day (154 tpd)	1	0
4	Water Gas Shift Reactors	Fixed bed, catalytic	199,000 kg/hr (439,000 lb/hr), 216°C (420°F), 4.0 MPa (580 psia)	4	0
5	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 121 GJ/hr (114 MMBtu/hr) Exchanger 2: 3 GJ/hr (3 MMBtu/hr)	4	0
6	Acid Gas Removal Plant	Two-stage selexol	316,000 kg/hr (697,000 lb/hr), 35°C (94°F), 3.6 MPa (520 psia)	2	0
7	Hydrogenation Reactor	Fixed bed, catalytic	18,000 kg/hr (40,000 lb/hr) 232°C (450°F) 0.1 MPa (12.2959488 psia)	1	0
8	Tail Gas Recycle Compressor	Centrifugal	14,000 kg/hr (30,000 lb/hr) each	1	0
9	CO ₂ Compressor	Integrally geared, multi-stage centrifugal	1,100 m ³ /min @ 15.3 MPa (38,700 scfm @ 2,215 psia)	4	0

Case B1B – Account 6: Combustion Turbine and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	State-of-the-art 2008 F-class	230 MW	2	0
2	Combustion Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0

Case B1B – Account 7: HRSG, Ducting, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.5 m (28 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 285,910 kg/hr, 12.4 MPa/534°C (630,323 lb/hr, 1,800 psig/993°F)	2	0

Case B1B – Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	220 MW 12.4 MPa/534°C/534°C (1,800 psig/ 993°F/993°F)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
2	Steam Turbine Generator	Hydrogen cooled, static excitation	240 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,470GJ/hr (1,390 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0

Case B1B – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	440,000 lpm @ 30 m (116,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 2,450 GJ/hr (2,320 MMBtu/hr) heat duty	1	0

Case B1B – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	222,000 liters (59,000 gal)	2	0
2	Slag Crusher	Roll	12 tonne/hr (13 tph)	2	0
3	Slag Depressurizer	Lock hopper	12 tonne/hr (13 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	133,000 liters (35,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	60,000 liters (16,000 gal)	2	0
6	Slag Conveyor	Drag chain	12 tonne/hr (13 tph)	2	0
7	Slag Separation Screen	Vibrating	12 tonne/hr (13 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	12 tonne/hr (13 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	190,000 liters (50,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	50 lpm @ 14 m H ₂ O (10 gpm @ 46 ft H ₂ O)	2	2
11	Grey Water Storage Tank	Field erected	60,000 liters (16,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	210 lpm @ 430 m H ₂ O (60 gpm @ 1,420 ft H ₂ O)	2	2
13	Slag Storage Bin	Vertical, field erected	800 tonne (900 tons)	2	0
14	Unloading Equipment	Telescoping chute	100 tonne/hr (110 tph)	1	0

Case B1B – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	24 kV/345 kV, 200 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 74 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 46 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 7 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0

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

Case B1B – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.2.11 Case B1B - Cost Estimating

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-38 shows a detailed breakdown of the capital costs; Exhibit 3-39 shows the owner's costs, TOC, and TASC; Exhibit 3-40 shows the initial and annual O&M costs; and Exhibit 3-41 shows the COE breakdown.

The estimated TPC of the Shell gasifier with CO₂ capture is \$3,981/kW. Process contingency represents 4.0 percent of the TPC, and project contingency represents 14.1 percent. The COE, including CO₂ T&S costs of \$9.8/MWh, is \$162.4/MWh.

Exhibit 3-38 Case B1B total plant cost details

Case:		B1B – Shell IGCC w/ CO ₂				Estimate Type:			Conceptual		
Plant Size (MW,net):		497				Cost Base:			Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
1		Coal & Sorbent Handling									
1.1	Coal Receive & Unload	\$4,425	\$0	\$2,133	\$0	\$6,558	\$656	\$0	\$1,443	\$8,656	\$17
1.2	Coal Stackout & Reclaim	\$5,718	\$0	\$1,367	\$0	\$7,086	\$709	\$0	\$1,559	\$9,353	\$19
1.3	Coal Conveyors & Yd Crush	\$5,316	\$0	\$1,353	\$0	\$6,669	\$667	\$0	\$1,467	\$8,803	\$18
1.4	Other Coal Handling	\$1,391	\$0	\$313	\$0	\$1,704	\$170	\$0	\$375	\$2,249	\$5
1.9	Coal & Sorbent Hnd. Foundations	\$0	\$2,955	\$7,723	\$0	\$10,677	\$1,068	\$0	\$2,349	\$14,094	\$28
Subtotal		\$16,851	\$2,955	\$12,889	\$0	\$32,694	\$3,269	\$0	\$7,193	\$43,156	\$87
2		Coal & Sorbent Prep & Feed									
2.1	Coal Crushing & Drying	\$50,700	\$3,058	\$7,286	\$0	\$61,043	\$6,104	\$0	\$13,430	\$80,577	\$162
2.2	Prepared Coal Storage & Feed	\$2,401	\$577	\$371	\$0	\$3,350	\$335	\$0	\$737	\$4,422	\$9
2.3	Dry Coal Injection System	\$79,030	\$911	\$7,239	\$0	\$87,180	\$8,718	\$0	\$19,180	\$115,078	\$232
2.4	Misc. Coal Prep & Feed	\$1,321	\$965	\$2,842	\$0	\$5,127	\$513	\$0	\$1,128	\$6,767	\$14
2.9	Coal & Sorbent Feed Foundation	\$0	\$4,844	\$4,156	\$0	\$9,000	\$900	\$0	\$1,980	\$11,880	\$24
Subtotal		\$133,452	\$10,354	\$21,894	\$0	\$165,700	\$16,570	\$0	\$36,454	\$218,724	\$440
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	\$2,970	\$5,126	\$2,687	\$0	\$10,784	\$1,078	\$0	\$2,372	\$14,234	\$29
3.2	Water Makeup & Pretreating	\$857	\$89	\$471	\$0	\$1,416	\$142	\$0	\$467	\$2,024	\$4
3.3	Other Feedwater Subsystems	\$1,670	\$552	\$493	\$0	\$2,715	\$271	\$0	\$597	\$3,584	\$7
3.4	Service Water Systems	\$501	\$999	\$3,443	\$0	\$4,942	\$494	\$0	\$1,631	\$7,068	\$14
3.5	Other Boiler Plant Systems	\$2,699	\$1,009	\$2,483	\$0	\$6,191	\$619	\$0	\$1,362	\$8,172	\$16
3.6	FO Supply Sys & Nat Gas	\$17,717	\$675	\$625	\$0	\$19,017	\$1,902	\$0	\$4,184	\$25,103	\$51
3.7	Waste Treatment Equipment	\$1,159	\$0	\$718	\$0	\$1,877	\$188	\$0	\$619	\$2,683	\$5
3.8	Misc. Power Plant Equipment	\$1,261	\$168	\$655	\$0	\$2,085	\$208	\$0	\$688	\$2,981	\$6
Subtotal		\$28,834	\$8,617	\$11,575	\$0	\$49,026	\$4,903	\$0	\$11,921	\$65,849	\$133
4		Gasifier & Accessories									
4.1	Syngas Cooler Gasifier System	\$171,383	\$0	\$73,493	\$0	\$244,876	\$24,488	\$33,487	\$46,617	\$349,467	\$704
4.2	Syngas Cooler	w/4.1	\$0	w/ 4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.3	ASU & Oxidant Compression	\$207,843	\$0	w/equip.	\$0	\$207,843	\$20,784	\$0	\$22,863	\$251,490	\$506
4.4	LT Heat Recovery & FG Saturation	\$30,655	\$0	\$11,647	\$0	\$42,303	\$4,230	\$0	\$9,307	\$55,840	\$112
4.5	Misc. Gasification Equipment	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.6	Flare Stack System	\$0	\$1,694	\$685	\$0	\$2,379	\$238	\$0	\$523	\$3,141	\$6
4.8	Major Component Rigging	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.9	Gasification Foundations	\$0	\$9,672	\$5,768	\$0	\$15,440	\$1,544	\$0	\$4,246	\$21,230	\$43
Subtotal		\$409,881	\$11,366	\$91,593	\$0	\$512,841	\$51,284	\$33,487	\$83,556	\$681,168	\$1,371
5A		Gas Cleanup & Piping									
5A.1	Double Stage Selexol	\$162,818	\$0	w/equip.	\$0	\$162,818	\$16,282	\$32,564	\$42,333	\$253,996	\$511
5A.2	Elemental Sulfur Plant	\$12,076	\$2,354	\$15,473	\$0	\$29,902	\$2,990	\$0	\$6,579	\$39,471	\$79
5A.3	Mercury Removal	\$2,022	\$0	\$1,528	\$0	\$3,550	\$355	\$177	\$816	\$4,899	\$10
5A.4	Shift Reactors	\$8,880	\$0	\$3,550	\$0	\$12,429	\$1,243	\$0	\$2,734	\$16,406	\$33
5A.5	Particulate Removal	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0

Case:		B1B – Shell IGCC w/ CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		497					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
5A.6	Blowback Gas Systems	\$2,562	\$431	\$241	\$0	\$3,234	\$323	\$0	\$711	\$4,269	\$9
5A.7	Fuel Gas Piping	\$0	\$1,043	\$683	\$0	\$1,726	\$173	\$0	\$380	\$2,278	\$5
5A.9	HGCU Foundations	\$0	\$944	\$636	\$0	\$1,581	\$158	\$0	\$522	\$2,261	\$5
Subtotal		\$188,356	\$4,773	\$22,111	\$0	\$215,240	\$21,524	\$32,741	\$54,075	\$323,580	\$651
5B		CO₂ Compression									
5B.2	CO ₂ Compression & Drying	\$39,421	\$5,913	\$16,551	\$0	\$61,885	\$6,188	\$0	\$13,615	\$81,688	\$164
Subtotal		\$39,421	\$5,913	\$16,551	\$0	\$61,885	\$6,188	\$0	\$13,615	\$81,688	\$164
6		Combustion Turbine & Accessories									
6.1	Combustion Turbine Generator	\$111,210	\$0	\$7,881	\$0	\$119,091	\$11,909	\$11,909	\$14,291	\$157,201	\$316
6.9	Combustion Turbine Foundations	\$0	\$923	\$1,068	\$0	\$1,992	\$199	\$0	\$657	\$2,848	\$6
Subtotal		\$111,210	\$923	\$8,950	\$0	\$121,083	\$12,108	\$11,909	\$14,948	\$160,049	\$322
7		HRSG, Ducting, & Stack									
7.1	Heat Recovery Steam Generator	\$29,630	\$0	\$5,739	\$0	\$35,369	\$3,537	\$0	\$3,891	\$42,796	\$86
7.3	Ductwork	\$0	\$2,111	\$1,480	\$0	\$3,591	\$359	\$0	\$790	\$4,740	\$10
7.4	Stack	\$4,074	\$0	\$1,520	\$0	\$5,594	\$559	\$0	\$615	\$6,769	\$14
7.9	HRSG, Duct & Stack Foundations	\$0	\$776	\$779	\$0	\$1,554	\$155	\$0	\$513	\$2,222	\$4
Subtotal		\$33,704	\$2,887	\$9,517	\$0	\$46,108	\$4,611	\$0	\$5,809	\$56,527	\$114
8		Steam Turbine Generator									
8.1	Steam TG & Accessories	\$35,245	\$0	\$4,790	\$0	\$40,035	\$4,003	\$0	\$4,404	\$48,442	\$98
8.2	Turbine Plant Auxiliaries	\$199	\$0	\$453	\$0	\$651	\$65	\$0	\$72	\$788	\$2
8.3	Condenser & Auxiliaries	\$3,020	\$0	\$1,697	\$0	\$4,717	\$472	\$0	\$519	\$5,708	\$11
8.4	Steam Piping	\$13,533	\$0	\$5,869	\$0	\$19,402	\$1,940	\$0	\$5,335	\$26,677	\$54
8.9	TG Foundations	\$0	\$937	\$1,655	\$0	\$2,592	\$259	\$0	\$855	\$3,707	\$7
Subtotal		\$51,997	\$937	\$14,463	\$0	\$67,397	\$6,740	\$0	\$11,185	\$85,322	\$172
9		Cooling Water System									
9.1	Cooling Towers	\$4,330	\$0	\$1,310	\$0	\$5,640	\$564	\$0	\$931	\$7,135	\$14
9.2	Circulating Water Pumps	\$2,088	\$0	\$154	\$0	\$2,242	\$224	\$0	\$370	\$2,836	\$6
9.3	Circ. Water System Auxiliaries	\$183	\$0	\$26	\$0	\$208	\$21	\$0	\$34	\$264	\$1
9.4	Circ. Water Piping	\$0	\$8,098	\$1,962	\$0	\$10,060	\$1,006	\$0	\$2,213	\$13,280	\$27
9.5	Make-up Water System	\$475	\$0	\$653	\$0	\$1,128	\$113	\$0	\$248	\$1,489	\$3
9.6	Component Cooling Water Sys	\$926	\$1,107	\$760	\$0	\$2,794	\$279	\$0	\$615	\$3,687	\$7
9.9	Circ. Water System Foundations	\$0	\$2,651	\$4,711	\$0	\$7,362	\$736	\$0	\$2,429	\$10,527	\$21
Subtotal		\$8,001	\$11,857	\$9,575	\$0	\$29,434	\$2,943	\$0	\$6,840	\$39,217	\$79
10		Ash & Spent Sorbent Handling Systems									
10.1	Slag Dewatering & Cooling	\$19,503	\$0	\$9,552	\$0	\$29,055	\$2,905	\$0	\$3,196	\$35,156	\$71
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.6	Ash Storage Silos	\$658	\$0	\$711	\$0	\$1,368	\$137	\$0	\$226	\$1,731	\$3
10.7	Ash Transport & Feed Equipment	\$905	\$0	\$211	\$0	\$1,117	\$112	\$0	\$184	\$1,413	\$3
10.8	Misc. Ash Handling Equipment	\$1,362	\$1,670	\$495	\$0	\$3,527	\$353	\$0	\$582	\$4,462	\$9
10.9	Ash/Spent Sorbent Foundation	\$0	\$55	\$73	\$0	\$128	\$13	\$0	\$42	\$183	\$0
Subtotal		\$22,428	\$1,725	\$11,042	\$0	\$35,195	\$3,519	\$0	\$4,230	\$42,945	\$86

Case:		B1B – Shell IGCC w/ CO ₂					Estimate Type:			Conceptual	
Plant Size (MW,net):		497					Cost Base:			Jun 2011	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
11		Accessory Electric Plant									
11.1	Generator Equipment	\$1,096	\$0	\$1,067	\$0	\$2,162	\$216	\$0	\$238	\$2,616	\$5
11.2	Station Service Equipment	\$5,293	\$0	\$487	\$0	\$5,780	\$578	\$0	\$636	\$6,994	\$14
11.3	Switchgear & Motor Control	\$9,770	\$0	\$1,816	\$0	\$11,586	\$1,159	\$0	\$1,912	\$14,657	\$30
11.4	Conduit & Cable Tray	\$0	\$4,964	\$15,302	\$0	\$20,266	\$2,027	\$0	\$5,573	\$27,866	\$56
11.5	Wire & Cable	\$0	\$9,571	\$5,823	\$0	\$15,394	\$1,539	\$0	\$4,233	\$21,167	\$43
11.6	Protective Equipment	\$0	\$798	\$2,963	\$0	\$3,761	\$376	\$0	\$621	\$4,758	\$10
11.7	Standby Equipment	\$263	\$0	\$262	\$0	\$525	\$53	\$0	\$87	\$664	\$1
11.8	Main Power Transformers	\$18,593	\$0	\$162	\$0	\$18,756	\$1,876	\$0	\$3,095	\$23,726	\$48
11.9	Electrical Foundations	\$0	\$170	\$462	\$0	\$632	\$63	\$0	\$209	\$904	\$2
Subtotal		\$35,016	\$15,503	\$28,345	\$0	\$78,864	\$7,886	\$0	\$16,603	\$103,353	\$208
12		Instrumentation & Control									
12.1	IGCC Control Equipment	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.2	Combustion Turbine Control	w/6.1	\$0	w/6.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.3	Steam Turbine Control	w/8.1	\$0	w/8.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.4	Other Major Component Control	\$1,274	\$0	\$868	\$0	\$2,143	\$214	\$107	\$370	\$2,834	\$6
12.5	Signal Processing Equipment	w/12.7	\$0	w/12.7	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.6	Control Boards, Panels & Racks	\$293	\$0	\$192	\$0	\$484	\$48	\$24	\$111	\$669	\$1
12.7	Computer & Accessories	\$6,798	\$0	\$222	\$0	\$7,020	\$702	\$351	\$807	\$8,880	\$18
12.8	Instrument Wiring & Tubing	\$0	\$2,617	\$4,953	\$0	\$7,570	\$757	\$379	\$2,176	\$10,882	\$22
12.9	Other I & C Equipment	\$4,544	\$0	\$2,251	\$0	\$6,796	\$680	\$340	\$1,172	\$8,987	\$18
Subtotal		\$12,909	\$2,617	\$8,487	\$0	\$24,013	\$2,401	\$1,201	\$4,637	\$32,252	\$65
13		Improvements to Site									
13.1	Site Preparation	\$0	\$118	\$2,686	\$0	\$2,804	\$280	\$0	\$925	\$4,010	\$8
13.2	Site Improvements	\$0	\$2,101	\$2,970	\$0	\$5,072	\$507	\$0	\$1,674	\$7,252	\$15
13.3	Site Facilities	\$3,765	\$0	\$4,227	\$0	\$7,992	\$799	\$0	\$2,637	\$11,429	\$23
Subtotal		\$3,765	\$2,219	\$9,883	\$0	\$15,868	\$1,587	\$0	\$5,236	\$22,691	\$46
14		Buildings & Structures									
14.1	Combustion Turbine Area	\$0	\$317	\$179	\$0	\$496	\$50	\$0	\$109	\$655	\$1
14.2	Steam Turbine Building	\$0	\$2,481	\$3,532	\$0	\$6,014	\$601	\$0	\$992	\$7,607	\$15
14.3	Administration Building	\$0	\$1,027	\$745	\$0	\$1,772	\$177	\$0	\$292	\$2,242	\$5
14.4	Circulation Water Pumphouse	\$0	\$193	\$102	\$0	\$295	\$29	\$0	\$49	\$373	\$1
14.5	Water Treatment Buildings	\$0	\$704	\$686	\$0	\$1,391	\$139	\$0	\$229	\$1,759	\$4
14.6	Machine Shop	\$0	\$526	\$360	\$0	\$885	\$89	\$0	\$146	\$1,120	\$2
14.7	Warehouse	\$0	\$849	\$548	\$0	\$1,397	\$140	\$0	\$230	\$1,767	\$4
14.8	Other Buildings & Structures	\$0	\$508	\$396	\$0	\$904	\$90	\$0	\$199	\$1,193	\$2
14.9	Waste Treating Building & Str.	\$0	\$1,137	\$2,171	\$0	\$3,307	\$331	\$0	\$728	\$4,366	\$9
Subtotal		\$0	\$7,742	\$8,718	\$0	\$16,460	\$1,646	\$0	\$2,975	\$21,081	\$42
Total		\$1,095,825	\$90,388	\$285,594	\$0	\$1,471,807	\$147,181	\$79,338	\$279,277	\$1,977,603	\$3,981

Exhibit 3-39 Case B1B owner's costs

Description	\$/1,000	\$/kW
Pre-Production Costs		
6 Months All Labor	\$15,919	\$32
1 Month Maintenance Materials	\$3,587	\$7
1 Month Non-fuel Consumables	\$682	\$1
1 Month Waste Disposal	\$434	\$1
25% of 1 Months Fuel Cost at 100% CF	\$2,910	\$6
2% of TPC	\$39,552	\$80
Total	\$63,083	\$127
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	\$24,228	\$49
0.5% of TPC (spare parts)	\$9,888	\$20
Total	\$34,116	\$69
Other Costs		
Initial Cost for Catalyst and Chemicals	\$16,739	\$34
Land	\$900	\$2
Other Owner's Costs	\$296,640	\$597
Financing Costs	\$53,395	\$107
Total Overnight Costs (TOC)	\$2,442,476	\$4,917
TASC Multiplier (IOU, high-risk, 35 year)	1.140	
Total As-Spent Cost (TASC)	\$2,784,423	\$5,605

Exhibit 3-40 Case B1B initial and annual operating and maintenance costs

Case:	B1B – Shell IGCC w/ CO ₂			Cost Base:	Jun 2011
Plant Size (MW _{net}):	497	Heat Rate-net (Btu/kWh):	10,927	Capacity Factor (%):	80
Operating & Maintenance Labor					
Operating Labor			Operating Labor Requirements per Shift		
Operating Labor Rate (base):	39.70	\$/hour	Skilled Operator:	2.0	
Operating Labor Burden:	30.00	% of base	Operator:	10.0	
Labor O-H Charge Rate:	25.00	% of labor	Foreman:	1.0	
			Lab Tech's, etc.:	3.0	
			Total:	16.0	
Fixed Operating Costs					
			Annual Cost		
			(\$)	(\$/kW-net)	
Annual Operating Labor:				\$7,233,658	\$14.562
Maintenance Labor:				\$18,236,067	\$36.711
Administrative & Support Labor:				\$6,367,431	\$12.818
Property Taxes and Insurance:				\$39,552,053	\$79.623
Total:				\$71,389,208	\$143.715
Variable Operating Costs					
			(\$)	(\$/MWh-net)	
Maintenance Material:				\$34,437,835	\$9.89265
Consumables					
	Consumption				Cost (\$)
	Initial Fill	Per Day	Per Unit	Initial Fill	
Water (/1000 gallons):	0	4,070	\$1.67	\$0	\$1,989,343
Makeup and Waste Water Treatment Chemicals (lbs):	0	24,247	\$0.27	\$0	\$1,896,308
Carbon (Mercury Removal) (lb):	201,639	345	\$5.50	\$1,109,016	\$554,508
Shift Catalyst (ft³):	6,470	4.43	\$771.99	\$4,994,685	\$998,937
Selexol Solution (gal):	289,068	92	\$36.79	\$10,635,172	\$989,212
Claus Catalyst (ft³):	w/equip.	1.93	\$203.15	\$0	\$114,583
Subtotal:				\$16,738,872	\$6,542,891
Waste Disposal					
Spent Mercury Catalyst (lb.):	0	345	\$0.65	\$0	\$65,542
Flyash (ton):	0	0	\$0.00	\$0	\$0
Slag (ton):	0	559	\$25.11	\$0	\$4,099,990
Subtotal:				\$0	\$4,165,532
By-Products					
Sulfur (tons):	0	140	\$0.00	\$0	\$0
Subtotal:				\$0	\$0
Variable Operating Costs Total:			\$16,738,872	\$45,146,258	\$12.96876
Fuel Cost					
Illinois Number 6 (ton):	0	5,583	\$68.54	\$0	\$111,739,739
Total:				\$0	\$111,739,739

Exhibit 3-41 Case B1B COE breakdown

Component	Value, \$/MWh	Percentage
Capital	87.0	57%
Fixed	20.5	13%
Variable	13.0	8%
Fuel	32.1	21%
Total (Excluding T&S)	152.6	N/A
CO ₂ T&S	9.8	6%
Total (Including T&S)	162.4	N/A

3.3 CB&I E-GasTM IGCC Cases

This section contains an evaluation of plant designs for cases B4A and B4B, which are based on the CB&I E-GasTM gasifier. Cases B4A and B4B are very similar in terms of process, equipment, scope and arrangement, except that Case B4B includes SGS reactors, CO₂ absorption/regeneration and compression/transport systems. There are no provisions for CO₂ removal in Case B4A.

The balance of this section is organized in an analogous manner to Section 3.2:

- Gasifier Background
- Process System Description for Case B4A
- Key Assumptions for Cases B4A and B4B
- Sparing Philosophy for Cases B4A and B4B
- Performance Results for Case B4A
- Equipment List for Case B4A
- Cost Estimates for Case B4A
- Process and System Description, Performance Results, Equipment List, and Cost Estimate for Case B4B

3.3.1 Gasifier Background

As mentioned in Section 3.2.1, the “Coal Gasification Guidebook: Status, Application, and Technologies” report published by EPRI provides a detailed history of the development of several types of gasifier technology, including the Shell gasifier, as well as gasifier capacity, distinguishing characteristics, and important coal characteristics. (46)

In January of 2000, CB&I completed construction of an E-GASTM demonstration facility, located at the Wabash River Generating Station in West Terre Haute, Indiana. It is still operational as a baseload unit. (39)

The Wabash River unit has a coal handling capacity of 1,678 tonnes/day (1,850 tons/day) – on a moisture and ash-free (MAF) basis – for bituminous coal with a high sulfur content. The unit produces dry gas at a rate of 189,724 Nm³/hr (6.7 million scf/hr) with an energy content of about 1,950 GJ/hr (1,850 MMBtu/hr) (HHV). This size matches the CT, which is a GE 7FA. (46)

The E-GasTM gasifier has significant operating experience with bituminous coal at full commercial scale via the Wabash plant. (46)

Compared to a single stage slurry-fed gasifier, the E-GasTM technology demonstrates superior efficiency and lower oxygen requirements. (46)

Notable characteristics of the E-GasTM gasifier are the relatively short refractory life and the high waste heat recovery rate, resulting from the high operating temperature. The E-GasTM gasifier produces a syngas with a higher CH₄ content than other single-stage slurry fed gasifiers, due to the use of quenching in the second stage. However, in CO₂ capture cases the CH₄ passes through the SGS reactors without change, and is also not separated by the AGR thus limiting the amount of carbon that can be captured. (46)

Bituminous coals with low moisture content are desired for use with the E-GasTM gasifier as it benefits the slurry concentration and lowers the oxygen requirement. As with all slagging gasifiers, low concentrations of ash with low to moderate ash fusion temperatures are preferred. (46)

3.3.2 Process Description

In this section the overall E-GasTM gasification process is described. The system description follows the BFD in Exhibit 3-42 and stream numbers reference the same exhibit. The tables in Exhibit 3-43 provide process data for the numbered streams in the BFD.

3.3.2.1 Coal Grinding and Slurry Preparation

Coal receiving and handling is common to all cases and was covered in Section 3.1.1. The receiving and handling subsystem ends at the coal silo. Coal grinding and slurry preparation is similar to the GEE cases but repeated here for completeness.

Coal from the coal silo is fed onto a conveyor by vibratory feeders located below each silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. Each hopper outlet discharges onto a weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 55 percent of the coal feed requirements of the gasifier. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is discharged through a trommel screen into the rod mill discharge tank, and then the slurry is pumped to the slurry storage tanks. The dry solids concentration of the final slurry is 63 percent. The Polk Power Station operates at a slurry concentration of 62-68 percent using bituminous coal, and ConocoPhillips presented a paper showing the slurry concentration of Illinois No. 6 coal as 63 percent. (49)

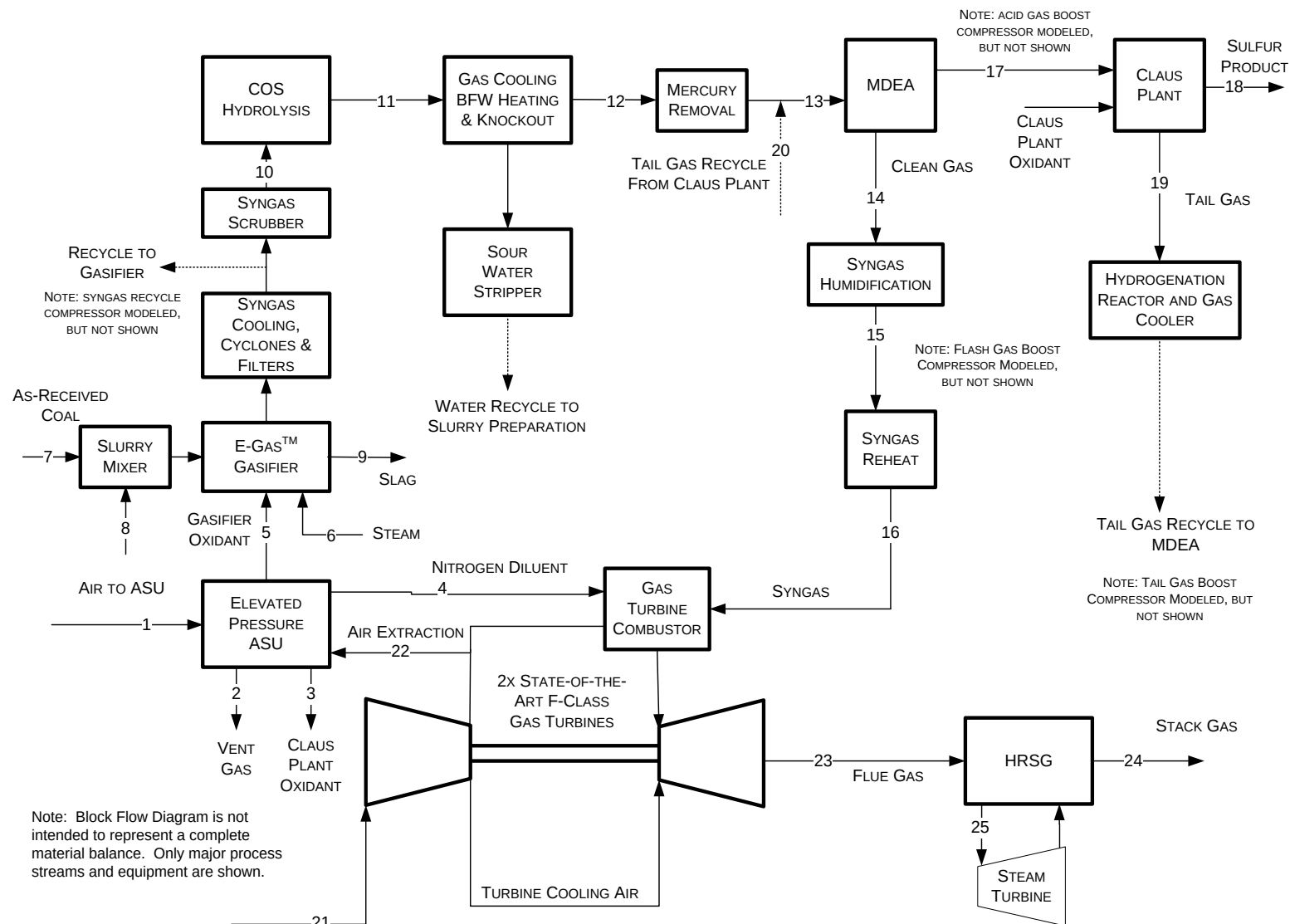
The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required depends on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended.

The equipment in the coal grinding and slurry preparation system is fabricated of materials appropriate for the abrasive environment present in the system. The tanks and agitators are rubber lined. The pumps are either rubber-lined or hardened metal to minimize erosion. Piping is fabricated of HDPE.

3.3.2.2 Gasification

This plant utilizes two gasification trains to process a total of 5,007 tonnes/day (5,519 tpd) of Illinois No. 6 coal. Each of the 2 x 50 percent gasifiers operate at maximum capacity. The E-GasTM two-stage coal gasification technology features an oxygen-blown, entrained-flow, refractory-lined gasifier with continuous slag removal. About 78 percent of the total slurry feed is fed to the first (or bottom) stage of the gasifier. The air separation plant supplies 3,711 tonnes/day (4,090 tpd) of 95 percent oxygen to the gasifiers (stream 5) and the Claus plant (stream 3). All oxygen for gasification is fed to this stage of the gasifier at a pressure of 4.2 MPa (615 psia). This stage is best described as a horizontal cylinder with two horizontally opposed burners. The highly exothermic gasification/oxidation reactions take place rapidly at temperatures of 1,316 to 1,427°C (2,400 to 2,600°F). The hot raw gas from the first stage enters the second (top) stage, which is a vertical cylinder perpendicular to the first stage. The remaining 22 percent of coal slurry is injected into this hot raw gas. The endothermic gasification/devolatilization reaction in this stage reduces the gasifier exit temperature to about 1,038°C (1,900°F). Total slurry to both stages is shown as stream 7 in Exhibit 3-42.

Exhibit 3-42 Case B4A block flow diagram, E-Gas™ IGCC without CO₂ capture



Source: NETL

Exhibit 3-43 Case B4A stream table, E-Gas™ IGCC without capture

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0241	0.0318	0.0023	0.0318	0.0000	0.0000	0.0000	0.0000	0.0078	0.0078	0.0096	0.0095	0.0098
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0465	0.0465	0.0575	0.0555	0.0576
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3011	0.3011	0.3727	0.3597	0.3734
CO ₂	0.0003	0.0083	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1556	0.1559	0.1929	0.2137	0.1931
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	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	0.2725	0.2725	0.3372	0.3299	0.3425
H ₂ O	0.0099	0.2164	0.0000	0.0003	0.0000	1.0000	0.0000	0.9968	0.0000	0.1888	0.1885	0.0015	0.0015	0.0015
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0072	0.0075	0.0092	0.0090	0.0000
N ₂	0.7732	0.5456	0.0178	0.9919	0.0178	0.0000	0.0000	0.0000	0.0000	0.0157	0.0157	0.0194	0.0213	0.0221
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0032	0.0000	0.0045	0.0045	0.0000	0.0000	0.0000
O ₂	0.2074	0.2055	0.9504	0.0054	0.9504	0.0000	0.0000	0.0000	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	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	18,975	915	111	17,428	4,694	3,766	0	4,758	0	21,815	21,815	17,626	18,275	17,601
V-L Flowrate (kg/hr)	547,573	24,798	3,567	489,025	151,055	67,847	0	85,706	0	459,698	459,698	384,238	408,328	380,348
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	208,633	0	21,295	0	0	0	0
Temperature (°C)	15	20	32	93	32	343	15	149	1,038	186	186	35	34	45
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	5.10	0.10	5.79	4.24	4.07	4.00	3.83	3.79	3.76
Steam Table Enthalpy (kJ/kg) ^A	30.23	36.92	26.67	92.52	26.67	3,063.97	---	567.36	---	673.89	673.68	39.55	36.82	55.08
AspenPlus Enthalpy (kJ/kg) ^B	-97.58	-2,404.24	4.13	63.38	4.13	-12,907.38	-2,116.95	-15,363.86	1,005.43	-6,602.13	-6,602.13	-5,592.59	-5,753.40	-5,625.00
Density (kg/m ³)	1.2	1.6	11.0	24.4	11.0	20.1	---	862.4	---	22.7	22.3	33.0	33.7	31.0
V-L Molecular Weight	28.857	27.090	32.181	28.060	32.181	18.015	---	18.012	---	21.073	21.073	21.799	22.344	21.610
V-L Flowrate (lb _{mole} /hr)	41,834	2,018	244	38,422	10,348	8,303	0	10,490	0	48,093	48,093	38,859	40,289	38,803
V-L Flowrate (lb/hr)	1,207,192	54,670	7,864	1,078,115	333,020	149,578	0	188,950	0	1,013,462	1,013,462	847,100	900,209	838,525
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	459,956	0	46,947	0	0	0	0
Temperature (°F)	59	69	90	199	90	650	59	300	1,900	367	367	95	94	112
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	740.0	14.7	840.0	615.0	590.0	580.0	555.0	550.0	545.0
Steam Table Enthalpy (Btu/lb) ^A	13.0	15.9	11.5	39.8	11.5	1,317.3	---	243.9	---	289.7	289.6	17.0	15.8	23.7
AspenPlus Enthalpy (Btu/lb) ^B	-42.0	-1,033.6	1.8	27.2	1.8	-5,549.2	-910.1	-6,605.3	432.3	-2,838.4	-2,838.4	-2,404.4	-2,473.5	-2,418.3
Density (lb/ft ³)	0.076	0.098	0.687	1.521	0.687	1.257	---	53.837	---	1.416	1.391	2.061	2.105	1.938

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 3-43 Case B4A stream table, E-Gas™ IGCC without capture (continued)

	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction											
Ar	0.0085	0.0085	0.0000	0.0000	0.0041	0.0055	0.0092	0.0092	0.0087	0.0087	0.0000
CH ₄	0.0500	0.0500	0.0002	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.3241	0.3241	0.0010	0.0000	0.0848	0.0061	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.1676	0.1676	0.7537	0.0000	0.4983	0.7790	0.0003	0.0003	0.0803	0.0803	0.0000
COS	0.0000	0.0000	0.0001	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.2973	0.2973	0.0009	0.0000	0.0209	0.1311	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.1333	0.1333	0.0000	0.0000	0.3350	0.0018	0.0099	0.0099	0.0860	0.0860	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.2443	0.0000	0.0014	0.0043	0.0000	0.0000	0.0000	0.0000	0.0000
N ₂	0.0192	0.0192	0.0000	0.0000	0.0537	0.0723	0.7732	0.7732	0.7250	0.7250	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
O ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.2074	0.1000	0.1000	0.0000
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0013	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	20,276	20,276	674	0	873	649	110,253	4,410	137,247	137,247	40,181
V-L Flowrate (kg/hr)	428,547	428,547	27,980	0	28,108	24,090	3,181,557	127,262	3,971,867	3,971,867	723,870
Solids Flowrate (kg/hr)	0	0	0	5,215	0	0	0	0	0	0	0
Temperature (°C)	143	193	45	175	232	38	15	432	588	132	561
Pressure (MPa, abs)	3.21	3.17	3.76	0.41	0.41	5.51	0.10	1.62	0.10	0.10	12.51
Steam Table Enthalpy (kJ/kg) ^A	489.40	569.86	-0.67	---	731.47	-6.63	30.23	463.79	783.86	272.30	3,500.61
AspenPlus Enthalpy (kJ/kg) ^B	-6,348.13	-6,267.67	-7,288.11	140.10	-8,672.98	-8,318.28	-97.58	335.98	-1,186.76	-1,698.31	-12,469.70
Density (kg/m ³)	19.7	17.3	73.5	5,285.4	3.1	96.3	1.2	7.9	0.4	0.9	35.2
V-L Molecular Weight	21.135	21.135	41.529	---	32.209	37.141	28.857	28.857	28.940	28.940	18.015
V-L Flowrate (lb _{mole} /hr)	44,702	44,702	1,485	0	1,924	1,430	243,066	9,723	302,577	302,577	88,584
V-L Flowrate (lb/hr)	944,784	944,784	61,685	0	61,967	53,109	7,014,133	280,565	8,756,467	8,756,467	1,595,860
Solids Flowrate (lb/hr)	0	0	0	11,497	0	0	0	0	0	0	0
Temperature (°F)	290	380	112	347	450	100	59	810	1,091	270	1,041
Pressure (psia)	465.0	460.0	545.0	59.3	58.9	799.5	14.7	234.9	15.2	15.2	1,814.7
Steam Table Enthalpy (Btu/lb) ^A	210.4	245.0	-0.3	---	314.5	-2.8	13.0	199.4	337.0	117.1	1,505.0
AspenPlus Enthalpy (Btu/lb) ^B	-2,729.2	-2,694.6	-3,133.3	60.2	-3,728.7	-3,576.2	-42.0	144.4	-510.2	-730.1	-5,361.0
Density (lb/ft ³)	1.230	1.081	4.586	329.954	0.195	6.012	0.076	0.495	0.026	0.056	2.194

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The syngas produced by the E-Gas™ gasifier is higher in methane content than either the GEE or Shell gasifier. The two-stage design allows for improved cold gas efficiency (CGE) and lower oxygen consumption, but the quenched second stage allows some CH₄ to remain. The syngas CH₄ concentration exiting the gasifier in Case B4A is 4.3 vol% (compared to 0.10 vol% in Case B5A [GEE] and 0.001 vol% in Case B1A [Shell]). The relatively high CH₄ concentration impacts carbon capture efficiency as discussed in Section 3.3.8.

3.3.2.3 Raw Gas Cooling/Particulate Removal

The raw syngas, less than 1,038°C (1,900°F), from the second stage of the gasifier is cooled to 316°C (600°F) in the waste heat recovery (SGC) unit, which consists of a fire-tube boiler and convective superheating and economizing sections. 554,830 kg/hr (1,223,171 lb/hr) of HP saturated steam is raised as a result of the raw gas cooling. Fire-tube boilers cost markedly less than comparable duty water-tube boilers. This is because of the large savings in high-grade steel associated with containing the hot HP syngas in relatively small tubes.

The coal ash is converted to molten slag, which flows down through a tap hole. The molten slag is quenched in water and removed through a proprietary continuous-pressure letdown/dewatering system (stream 9). Char is produced in the second gasifier stage and is captured and recycled to the hotter first stage to be gasified.

The cooled gas from the SGC is cleaned of remaining particulate via a cyclone collector followed by a ceramic candle filter. Recycled syngas is used as the pulse gas to clean the candle filters. The recovered fines are pneumatically returned to the first stage of the gasifier. The combination of recycled char and recycled particulate results in high overall carbon conversion (99.2 percent used in this report).

Following particulate removal, additional heat is removed from the syngas to raise saturated IP steam at 0.4 MPa (65 psia). In this manner the syngas is cooled to 232°C (450°F) prior to the syngas scrubber.

3.3.2.4 Syngas Scrubber/Sour Water Stripper

Syngas exiting the second of the two low-temperature heat exchangers passes to a syngas scrubber where a water wash is used to remove chlorides, SO₂, NH₃, and particulate. The syngas exits the scrubber saturated at 169°C (337°F).

The sour water stripper removes NH₃, SO₂, and impurities from the scrubber and other waste streams. The stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from SGCs. Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the SRU. Remaining water is sent to wastewater treatment.

3.3.2.5 COS Hydrolysis, Mercury Removal and Acid Gas Removal

Syngas exiting the scrubber is reheated to 186°C (367°F) by using HP steam from the HRSG evaporator prior to entering a COS hydrolysis reactor (stream 10). About 99.5 percent of the COS is converted to CO₂ and H₂O (Section 3.1.5). The gas exiting the COS reactor (stream 11) passes through a series of heat exchangers and KO drums to lower the syngas temperature to 35°C (95°F) and to separate entrained water. The cooled syngas (stream 12) then passes through a carbon bed to remove approximately 97 percent of the Hg (Section 3.1.4).

Cool, particulate-free syngas (stream 13) enters the absorber unit at approximately 3.8 MPa (550 psia) and 34°C (94°F). In the absorber, H₂S is preferentially removed from the fuel gas stream by contact with MDEA. The absorber column is operated at 44°C (112°F) by refrigerating the lean MDEA solvent. The lower temperature is required to achieve an outlet H₂S concentration of less than 30 ppmv in the sweet syngas. The stripper acid gas stream (stream 17), consisting of 24 percent H₂S and 75 percent CO₂, is sent to the Claus unit. The acid gas is combined with the sour water stripper off gas and introduced into the Claus plant burner section.

3.3.2.6 Claus Unit

Acid gas from the MDEA unit is preheated to 232°C (450°F). A portion of the acid gas along with all of the sour gas from the stripper and oxygen from the ASU are fed to the Claus furnace. In the furnace, H₂S is catalytically oxidized to SO₂ at a furnace temperature of 1,316°C (2,400°F), which must be maintained in order to thermally decompose all of the NH₃ present in the sour gas stream.

Following the thermal stage and condensation of sulfur, two reheaters and two sulfur converters are used to obtain a per-pass H₂S conversion of approximately 99.5 percent. The Claus Plant tail gas is hydrogenated and recycled back to the gasifier (stream 20). In the furnace waste heat boiler, 13,866 kg/hr (30,568 lb/hr) of 3.0 MPa (430 psia) steam is generated. This steam is used to satisfy all Claus process preheating and reheating requirements as well as to provide some steam to the medium-pressure steam header. The sulfur condensers produce 0.34 MPa (50 psig) steam for the LP steam header and IP steam at 2.9 MPa (415 psig).

A flow rate of 5,215 kg/hr (11,497 lb/hr) of elemental sulfur (stream 18) is recovered from the fuel gas stream. This value represents an overall sulfur recovery efficiency of 99.7 percent.

3.3.2.7 Power Block

Clean syngas exiting the MDEA absorber (stream 14) is partially humidified (stream 15) because there is not sufficient nitrogen from the ASU to provide the level of dilution required to reach the target syngas heating value. The moisturized syngas stream is reheated (stream 16), further diluted with nitrogen from the ASU (stream 4) and enters the state-of-the-art 2008 F-class CT burner. The CT compressor provides combustion air to the burner and also 19 percent of the total ASU air requirement (stream 22). The exhaust gas exits the CT at 588°C (1,091°F) (stream 23) and enters the HRSG where additional heat is recovered until the flue gas exits the HRSG at 132°C (270°F) (stream 24) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced, commercially available steam turbine using a 12.4 MPa/561°C/561°C (1,800 psig/1,041°F/1,041°F) steam cycle.

3.3.2.8 Air Separation Unit

The elevated pressure ASU was described in Section 3.1.2. In Case B4A, the ASU is designed to produce a nominal output of 3,711 tonnes/day (4,091 tpd) of 95 mol% O₂ for use in the gasifier (stream 5) and Claus plant (stream 3). The plant is designed with two production trains. The air compressor is powered by an electric motor. Approximately 11,736 tonnes/day (12,937 tpd) of nitrogen is also recovered, compressed, and used as dilution in the CT combustor (stream 4). About 4 percent of the CT air is used to supply approximately 19 percent of the ASU air requirements (stream 22) in the non-capture case.

3.3.2.9 Balance of Plant

Balance of plant items were covered in Sections 3.1.9, 3.1.10, and 3.1.12.

3.3.3 Key System Assumptions

System assumptions for cases B4A and B4B, E-GasTM IGCC with and without CO₂ capture, are compiled in Exhibit 3-44.

3.3.3.1 Balance of Plant – Cases B4A and B4B

The balance of plant assumptions are common to all cases and were presented previously in Exhibit 3-11.

3.3.4 Sparing Philosophy

The sparing philosophy for cases B4A and B4B is provided below. Dual trains are used to accommodate the size of commercial CTs. There is no redundancy other than normal sparing of rotating equipment. The plant design consists of the following major subsystems:

- Two ASUs (2 x 50%)
- Two trains of slurry preparation and slurry pumps (2 x 50%)
- Two trains of gasification, including gasifier, SGC, cyclone, and candle filter (2 x 50%)
- Two trains of syngas clean-up process (2 x 50%)
- Two trains of refrigerated MDEA AGR in Case B4A and two-stage Selexol in Case B4B (2 x 50%)
- Two CO₂ compression systems (2 x 50%) in Case B4B
- One train of Claus-based sulfur recovery (1 x 100%)
- Two CT/HRSG tandems (2 x 50%)
- One steam turbine (1 x 100%)

Exhibit 3-44 E-Gas™ IGCC plant study configuration matrix

Case	B4A	B4B
Gasifier Pressure, MPa (psia)	4.2 (615)	4.2 (615)
O ₂ :Coal Ratio, kg O ₂ /kg dry coal	0.81	0.88
Carbon Conversion, %	99.2	99.2
Syngas HHV at Gasifier Outlet, kJ/Nm ³ (Btu/scf)	8,319 (223)	7,021 (189)
Steam Cycle, MPa/°C/°C (psig/°F/°F)	12.4/561/561 (1,800/1,041/1,041)	12.4/534/534 (1,800/994/994)
Condenser Pressure, mm Hg (in Hg)	51 (2.0)	51 (2.0)
CT	2x State-of-the-Art 2008 F-Class (232 MW output each)	2x State-of-the-Art 2008 F-Class (232 MW output each)
Gasifier Technology	CB&I E-Gas™	CB&I E-Gas™
Oxidant	95 vol% Oxygen	95 vol% Oxygen
Coal	Illinois No. 6	Illinois No. 6
Coal Slurry Solids Content, %	63	63
COS Hydrolysis	Yes	Occurs in SGS
SGS	No	Yes
H ₂ S Separation	Refrigerated MDEA	Selexol 1 st Stage
Sulfur Removal, %	99.7	99.9
Sulfur Recovery	Claus Plant with Tail Gas Recycle to Gasifier/ Elemental Sulfur	Claus Plant with Tail Gas Recycle to Gasifier/ Elemental Sulfur
Particulate Control	Cyclone, Candle Filter, Scrubber, and AGR Absorber	Cyclone, Candle Filter, Scrubber, and AGR Absorber
Mercury Control	Carbon Bed	Carbon Bed
NO _x Control	MNQC (LNB), N ₂ Dilution and Humidification	MNQC (LNB), N ₂ Dilution and Humidification
CO ₂ Separation	N/A	Selexol 2 nd Stage
Overall Carbon Capture	N/A	89.9%
CO ₂ Sequestration	N/A	Off-site Saline Formation

3.3.5 Case B4A - Performance Results

The plant produces a net output of 625 MWe at a net plant efficiency of 39.7 percent (HHV basis).

Overall performance for the entire plant is summarized in Exhibit 3-45; Exhibit 3-46 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for approximately 76 percent of the total auxiliary load distributed between the main air compressor, oxygen compressor, nitrogen compressor, and ASU auxiliaries. The cooling water system, including the CWP and cooling tower fan, accounts for approximately 5 percent of the auxiliary load, and the

BFW pumps account for an additional 3.9 percent. All other individual auxiliary loads are less than 3.5 percent of the total.

Exhibit 3-45 Case B4A plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	274
Total Gross Power, MWe	738
Air Separation Unit Main Air Compressor, kWe	46,700
Oxygen Compressor, kWe	7,920
Nitrogen Compressors, kWe	29,910
CO ₂ Compression, kWe	0
Acid Gas Removal, kWe	3,150
Balance of Plant, kWe	25,530
Total Auxiliaries, MWe	113
Net Power, MWe	625
HHV Net Plant Efficiency (%)	39.7%
HHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	9,058 (8,585)
HHV Cold Gas Efficiency, %	82.0%
HHV Combustion Turbine Efficiency, %	37.2%
LHV Net Plant Efficiency (%)	41.2%
LHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	8,737 (8,281)
LHV Cold Gas Efficiency, %	78.2%
LHV Combustion Turbine Efficiency, %	40.4%
Steam Turbine Cycle Efficiency, %	45.3%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	7,951 (7,536)
Condenser Duty, GJ/hr (MMBtu/hr)	1,415 (1,341)
As-Received Coal Feed, kg/hr (lb/hr)	208,633 (459,956)
HHV Thermal Input, kWt	1,572,575
LHV Thermal Input, kWt	1,516,768
Raw Water Withdrawal, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.027 (7.0)
Raw Water Consumption, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.021 (5.6)
O ₂ :Coal	0.684

Exhibit 3-46 Case B4A plant power summary

Power Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	274
Total Gross Power, MWe	738
Auxiliary Load Summary	
Coal Handling, kWe	460
Coal Milling, kWe	2,150
Sour Water Recycle Slurry Pump, kWe	180
Slag Handling, kWe	1,100
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	46,700
Oxygen Compressor, kWe	7,920
Nitrogen Compressors, kWe	29,910
Feedwater Pumps, kWe	4,410
Quench Water Pump, kWe	0
Syngas Recycle Compressor, kWe	810
CO ₂ Compression, kWe	0
Scrubber Pumps, kWe	320
Acid Gas Removal, kWe	3,150
Combustion Turbine Auxiliaries, kWe	1,000
Steam Turbine Auxiliaries, kWe	100
Condensate Pumps, kWe	240
Circulating Water Pumps, kWe	3,890
Ground Water Pumps, kWe	400
Cooling Tower Fans, kWe	2,010
Claus Plant/TGTU Auxiliaries, kWe	250
Claus Plant TG Recycle Compressor, kWe	1,610
Miscellaneous Balance of Plant ^A , kWe	3,000
Transformer Losses, kWe	2,600
Total Auxiliaries, MWe	113
Net Power, MWe	625

^AIncludes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.3.5.1 Environmental Performance

The environmental targets for emissions of Hg, NO_x, SO₂, and PM were presented in Section 2.3. A summary of the plant air emissions for Case B4A is presented in Exhibit 3-47. All HCl is assumed to be removed and is, therefore, not reported.

Exhibit 3-47 Case B4A air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year) ^A	kg/MWh (lb/MWh) ^B
SO ₂	0.005 (0.012)	200 (220)	0.039 (0.085)
NO _x	0.026 (0.060)	1,017 (1,122)	0.197 (0.434)
Particulates	0.003 (0.0071)	121 (133)	0.023 (0.052)
Hg	1.77E-7 (4.13E-7)	0.007 (0.008)	1.36E-6 (3.00E-6)
CO ₂	86 (199)	3,398,347 (3,746,036)	657 (1,448)
CO ₂ ^C	-	-	776 (1,711)

^ACalculations based on an 80 percent capacity factor

^BEmissions are based on gross power except where otherwise noted

^CCO₂ emissions based on net power instead of gross power

The low level of SO₂ in the plant emissions is achieved by capturing the sulfur in the gas by the refrigerated MDEA AGR process. The AGR process removes 99.7 percent of the sulfur compounds in the fuel gas down to a level of less than 30 ppmv. This results in a concentration in the flue gas of less than 4 ppmv. The H₂S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is hydrogenated to convert all sulfur species to H₂S and then recycled back to the gasifier, thereby eliminating the need for a tail gas treatment unit.

NO_x emissions are limited by the use of nitrogen dilution (primarily) and humidification (to a lesser extent) to 15 ppmvd (as NO₂ @ 15 percent O₂). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process and destroyed in the Claus plant burner. This helps lower NO_x levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a cyclone and a barrier filter in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of the mercury is captured from the syngas by dual activated carbon beds.

CO₂ emissions represent the uncontrolled discharge from the process.

The carbon balance for the plant is shown in Exhibit 3-48. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not neglected here since the Aspen model accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag and as CO₂ in the stack gas and ASU vent gas.

Exhibit 3-48 Case B4A carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	132,992 (293,198)	Stack Gas	132,344 (291,768)
Air (CO ₂)	507 (1,118)	CO ₂ Product	0 (0)
		ASU Vent	92 (202)
		Slag	1,064 (2,346)
Total	133,499 (294,316)	Total	133,499 (294,316)

Exhibit 3-49 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant and sulfur emitted in the stack gas. Sulfur in the slag is considered to be negligible.

Exhibit 3-49 Case B4A sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,229 (11,528)	Stack Gas	14 (31)
		CO ₂ Product	0 (0)
		Elemental Sulfur	5,215 (11,497)
Total	5,229 (11,528)	Total	5,229 (11,528)

Exhibit 3-50 shows the overall water balance for the plant. The water balance was explained in Case B1A [Shell], but is also presented here for completeness.

Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is defined as the water removed from the ground or diverted from a surface-water source for use in the plant and was assumed to be provided 50 percent by a POTW and 50 percent from groundwater. Raw water withdrawal can be represented by the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, quench system makeup, and slag handling makeup. 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 3-50 Case B4A water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)
Slag Handling	0.46 (122)	0.46 (122)	—	—	—
Slurry Water	1.43 (378)	0.99 (262)	0.44 (115)	—	0.44 (115)
Quench/Wash	—	—	—	—	—
Humidifier	0.84 (222)	0.84 (222)	—	—	—
SWS Blowdown	—	—	—	0.01 (3.84)	-0.01 (-3.84)
Condenser Makeup	1.34 (354)	—	1.34 (354)	—	1.34 (354)
BFW Makeup	0.21 (55)	—	0.21 (55)	—	0.21 (55)
Gasifier Steam	1.13 (299)	—	1.13 (299)	—	1.13 (299)
Shift Steam	—	—	—	—	—
CT Steam Dilution	—	—	—	—	—
Cooling Tower	15.16 (4,006)	0.35 (93)	14.81 (3,913)	3.41 (901)	11.40 (3,012)
BFW Blowdown	—	0.21 (55)	-0.21 (-55)	—	-0.21 (-55)
SWS Blowdown	—	0.15 (38.44)	-0.15 (-38.44)	—	-0.15 (-38.44)
Total	19.24 (5,082)	2.65 (700)	16.59 (4,382)	3.42 (905)	13.16 (3,477)

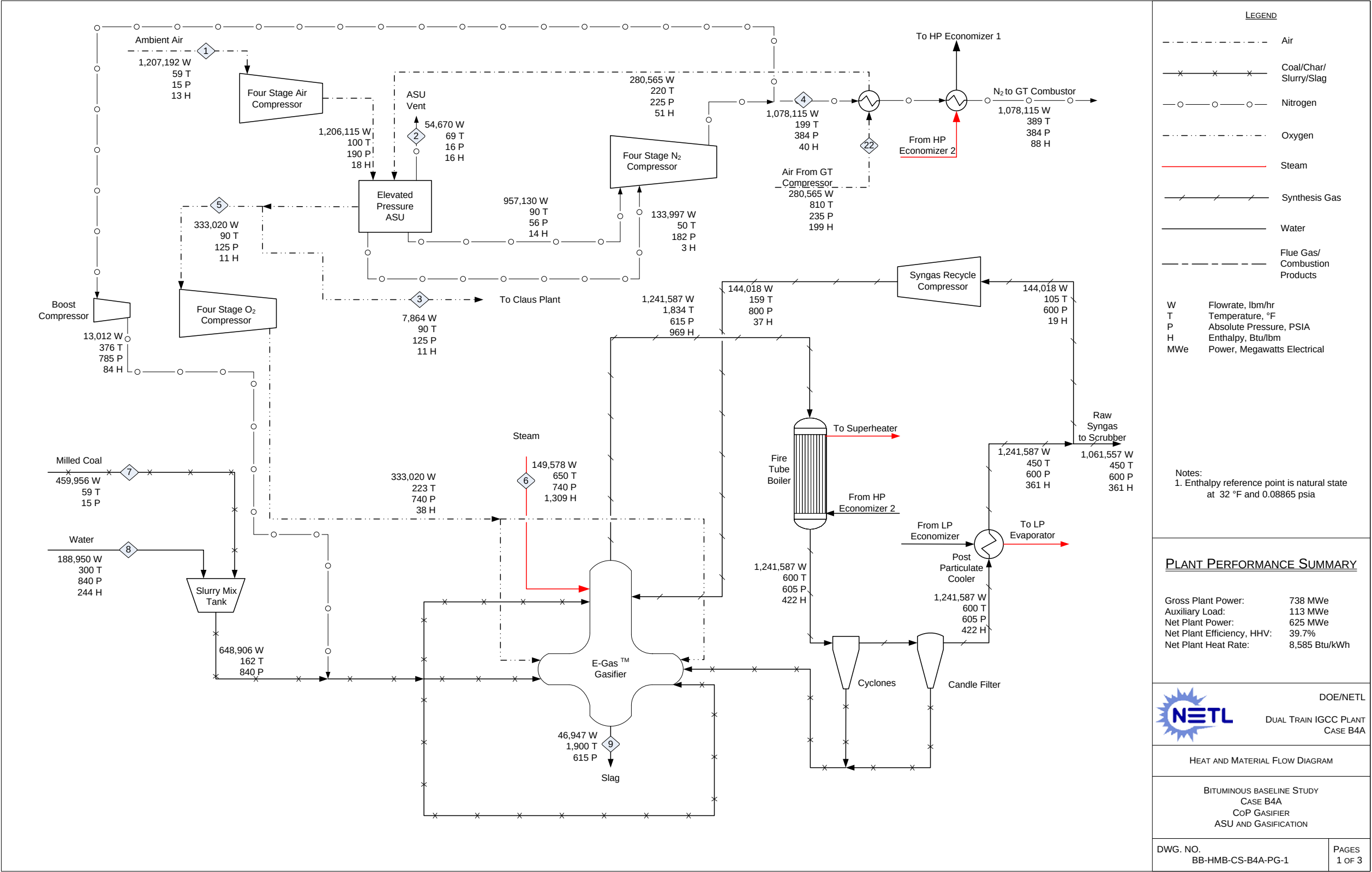
3.3.5.2 Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-51 through Exhibit 3-53:

- Coal gasification and ASU
- Syngas cleanup, sulfur recovery, and tail gas recycle
- Combined cycle power generation, steam, and FW

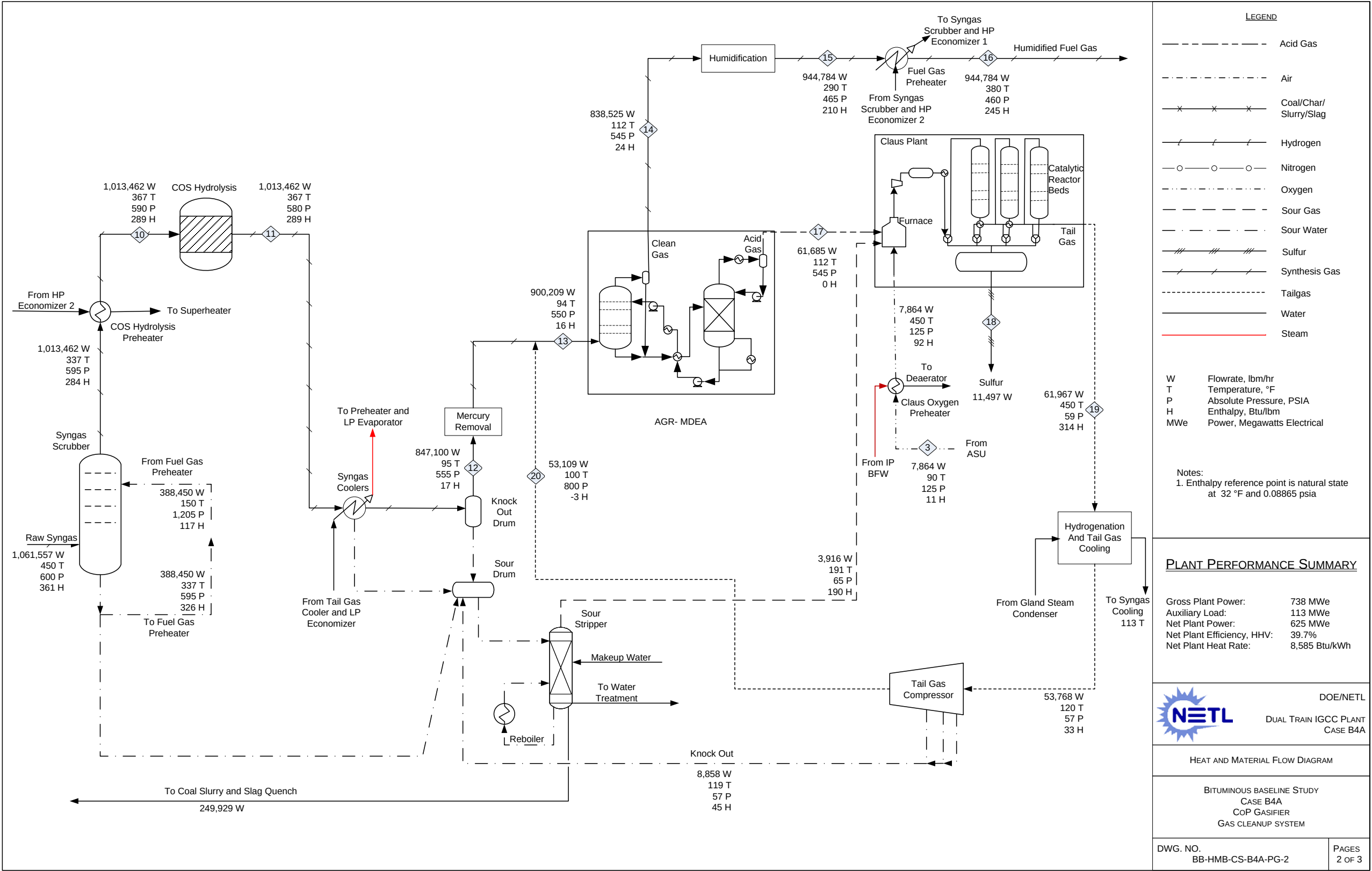
An overall plant energy balance is provided in tabular form in Exhibit 3-54. The power out is the combined CT and steam turbine power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-45) is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

Exhibit 3-51 Case B4A coal gasification and ASU heat and mass balance schematic



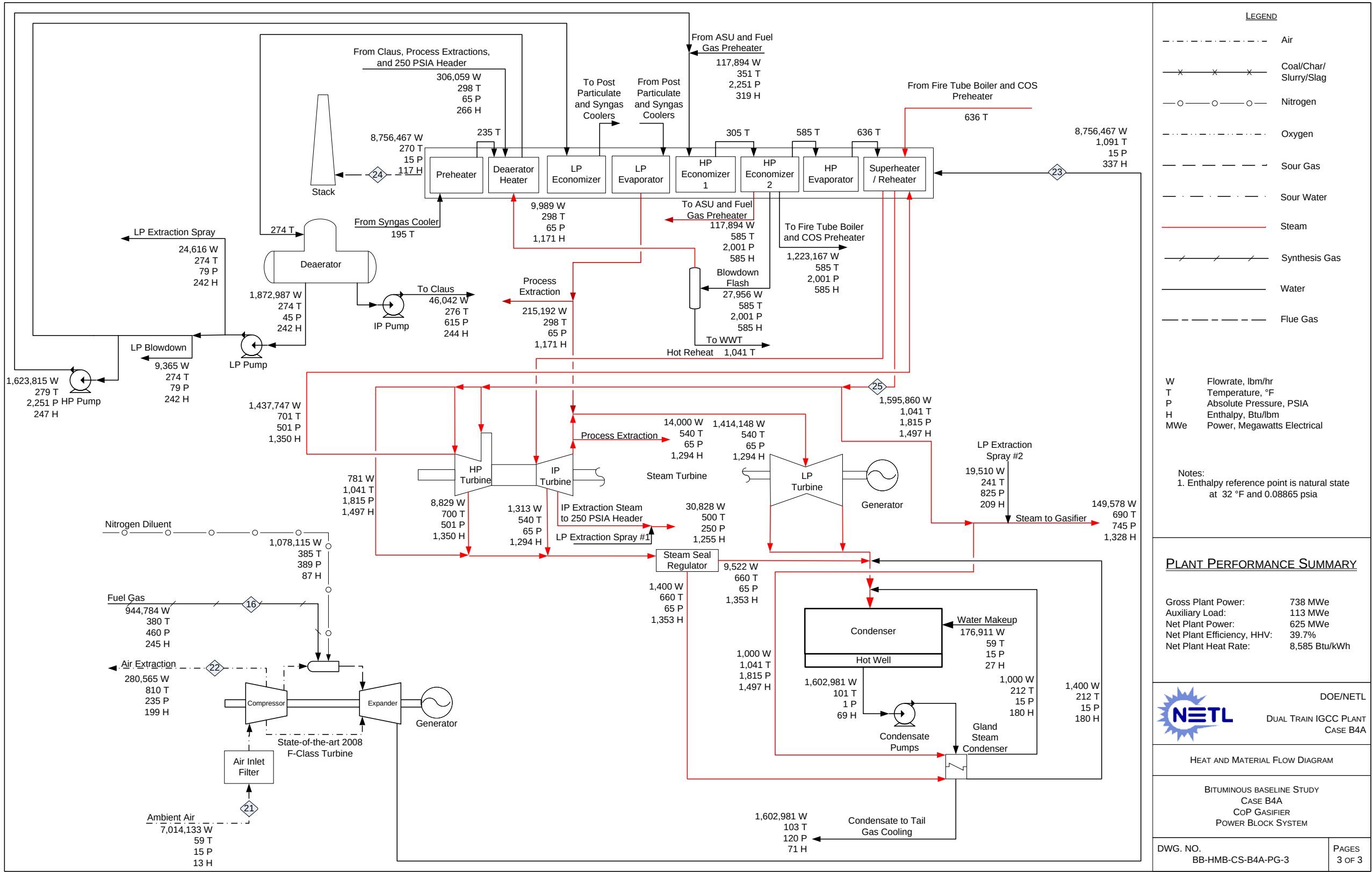
Source: NETL

Exhibit 3-52 Case B4A syngas cleanup heat and mass balance schematic



Source: NETL

Exhibit 3-53 Case B4A combined cycle power generation heat and mass balance schematic



Source: NETL

This page intentionally left blank

Exhibit 3-54 Case B4A overall energy balance (0°C [32°F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In (GJ/hr)				
Coal	5,661 (5,366)	4.7 (4.5)	–	5,666 (5,370)
Air	–	112.7 (106.8)	–	112.7 (106.8)
Raw Water Makeup	–	62.4 (59.1)	–	62.4 (59.1)
Auxiliary Power	–	–	408 (386)	408 (386)
TOTAL	5,661 (5,366)	179.8 (170.4)	408 (386)	6,249 (5,923)
Heat Out (GJ/hr)				
ASU Vent	–	0.9 (0.9)	–	0.9 (0.9)
Slag	34.9 (33.1)	23.9 (22.7)	–	58.8 (55.7)
Stack Gas	–	1,082 (1,025)	–	1,082 (1,025)
Sulfur	48.3 (45.8)	0.6 (0.6)	–	48.9 (46.4)
Motor Losses and Design Allowances	–	–	52.3 (49.6)	52.3 (49.6)
Condenser	–	1,415 (1,341)	–	1,415 (1,341)
Non-Condenser Cooling Tower Loads	–	445 (421)	–	445 (421)
CO ₂	–	–	–	–
Blowdown	–	32.2 (30.5)	–	32.2 (30.5)
CO ₂ Capture Losses	–	–	–	–
<i>Ambient Losses^A</i>	–	157.4 (149.2)	–	157.4 (149.2)
Power	–	–	2,658 (2,519)	2,658 (2,519)
TOTAL	83.2 (78.9)	3,157 (2,992)	2,710 (2,568)	5,950 (5,639)
Unaccounted Energy ^B	–	299 (283)	–	299 (283)

^AAmbient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers.

^BBy difference

3.3.6 Case B4A - Major Equipment List

Major equipment items for the E-GasTM gasifier with no CO₂ capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 3.3.7. In general, the design conditions include a 10 percent contingency for flows and heat duties, and a 21 percent contingency for heads on pumps and fans.

Case B4A – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	40 tonne (50 ton)	2	1
7	Feeder	Vibratory	170 tonne/hr (190 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	340 tonne/hr (380 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	170 tonne (190 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/ tripper	340 tonne/hr (380 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	340 tonne/hr (380 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	760 tonne (840 ton)	3	0

Case B4A – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/h (80 tph)	3	0
2	Conveyor No. 6	Belt w/ tripper	230 tonne/h (250 tph)	1	0
3	Rod Mill Feed Hopper	Dual Outlet	460 tonne (510 ton)	1	0
4	Weigh Feeder	Belt	110 tonne/h (130 tph)	2	0
5	Rod Mill	Rotary	110 tonne/h (130 tph)	2	0
6	Slurry Water Storage Tank with Agitator	Field erected	283,240 liters (74,820 gal)	2	0
7	Slurry Water Pumps	Centrifugal	790 lpm (210 gpm)	2	1
8	Trommel Screen	Coarse	160 tonne/h (180 tph)	2	0
9	Rod Mill Discharge Tank with Agitator	Field erected	370,530 liters (97,880 gal)	2	0
10	Rod Mill Product Pumps	Centrifugal	3,100 lpm (800 gpm)	2	2
11	Slurry Storage Tank with Agitator	Field erected	1,111,600 liters (293,600 gal)	2	0
12	Slurry Recycle Pumps	Centrifugal	6,200 lpm (1,600 gpm)	2	2
13	Slurry Product Pumps	Positive displacement	3,100 lpm (800 gpm)	2	2

Case B4A – Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1,083,000 liters (286,000 gal)	2	0
2	Condensate Pumps	Vertical canned	6,710 lpm @ 90 m H ₂ O (1,770 gpm @ 300 ft H ₂ O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	479,000 kg/hr (1,055,000 lb/hr)	2	0
4	Intermediate-Pressure Feedwater Pump	Horizontal centrifugal, single stage	8,100 lpm @ 30 m H ₂ O (2,140 gpm @ 90 ft H ₂ O)	2	1
5	High-Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 7,020 lpm @ 1,800 m H ₂ O (1,860 gpm @ 6,100 ft H ₂ O)	2	1
6	High-Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 510 lpm @ 220 m H ₂ O (130 gpm @ 730 ft H ₂ O)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded Screw	28 m ³ /min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m ³ /min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	245 GJ/hr (232 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	87,700 lpm @ 20 m H ₂ O (23,200 gpm @ 70 ft H ₂ O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H ₂ O (1,000 gpm @ 350 ft H ₂ O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H ₂ O (700 gpm @ 250 ft H ₂ O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	4,240 lpm @ 20 m H ₂ O (1,120 gpm @ 60 ft H ₂ O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	2,830 lpm @ 270 m H ₂ O (750 gpm @ 880 ft H ₂ O)	3	1
16	Filtered Water Pumps	Stainless steel, single suction	1,640 lpm @ 50 m H ₂ O (430 gpm @ 160 ft H ₂ O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	787,000 liter (208,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	300 lpm (80 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0

Case B4A – Account 4: Gasifier, ASU, and Accessories Including Low Temperature Heat Recovery and Fuel Gas Saturation

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized two-stage, slurry-feed entrained bed	2,800 tonne/day, 4.2 MPa (3,000 tpd, 615 psia)	2	0
2	Synthesis Gas Cooler	Fire-tube boiler	310,000 kg/hr (683,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	310,000 kg/hr (683,000 lb/hr) Design efficiency 90%	2	0
4	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0
5	Syngas Scrubber Including Sour Water Stripper	Vertical upflow	265,000 kg/hr (584,000 lb/hr)	2	0
6	Raw Gas Coolers	Shell and tube with condensate drain	213,000 kg/hr (469,000 lb/hr)	8	0
7	Raw Gas Knockout Drum	Vertical with mist eliminator	212,000 kg/hr, 35°C, 3.9 MPa (468,000 lb/hr, 95°F, 560 psia)	2	0
8	Saturation Water Economizers	Shell and tube	91 GJ/hr (86 MMBtu/hr)	2	0
9	Fuel Gas Saturator	Vertical tray tower	236,000 kg/hr, 143°C, 3.3 MPa (520,000 lb/hr, 290°F, 480 psia)	2	0
10	Saturator Water Pump	Centrifugal	2,100 lpm @ 13 m H ₂ O (600 gpm @ 42 ft H ₂ O)	2	2
11	Synthesis Gas Reheater	Shell and tube	209,000 kg/hr (461,000 lb/hr)	2	0
12	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	265,000 kg/hr (584,000 lb/hr) syngas	2	0
13	ASU Main Air Compressor	Centrifugal, multi-stage	4,000 m ³ /min @ 1.3 MPa (146,000 scfm @ 190 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
14	Cold Box	Vendor design	2,000 tonne/day (2,200 tpd) of 95% purity oxygen	2	0
15	Oxygen Compressor	Centrifugal, multi-stage	1,000 m ³ /min (36,000 scfm) Suction - 1.0 MPa (130 psia) Discharge - 5.1 MPa (740 psia)	2	0
16	Primary Nitrogen Compressor	Centrifugal, multi-stage	3,000 m ³ /min (119,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0
17	Secondary Nitrogen Compressor	Centrifugal, single-stage	470 m ³ /min (17,000 scfm) Suction - 1.3 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0
18	Transport Nitrogen Boost Compressor	Centrifugal, single-stage	50 m ³ /min (2,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 5.4 MPa (790 psia)	2	0
19	Extraction Air Heat Exchanger	Gas-to-gas, vendor design	70,000 kg/hr, 432°C, 1.6 MPa (154,000 lb/hr, 810°F, 235 psia)	2	0

Case B4A – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	211,000 kg/hr (466,000 lb/hr), 35°C (95°F), 3.8 MPa (555 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	211,000 kg/hr (466,000 lb/hr), 35°C (95°F), 3.8 MPa (550 psia)	2	0
3	Sulfur Plant	Claus type	138 tonne/day (152 tpd)	1	0
4	COS Hydrolysis Reactor	Fixed bed, catalytic	253,000 kg/hr (557,000 lb/hr), 188°C (370°F), 4.1 MPa (590 psia)	2	0
5	Acid Gas Removal Plant	MDEA	225,000 kg/hr (495,000 lb/hr), 34°C (94°F), 3.8 MPa (550 psia)	2	0
6	Hydrogenation Reactor	Fixed bed, catalytic	31,000 kg/hr (68,000 lb/hr), 232°C (450°F), 0.4 MPa (58.9 psia)	1	0
7	Tail Gas Recycle Compressor	Centrifugal	26,000 kg/hr (58,000 lb/hr) each	1	0

Case B4A – Account 6: Combustion Turbine and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	State-of-the-art 2008 F-class	230 MW	2	0
2	Combustion Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0

Case B4A – Account 7: HRSG, Ducting, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.4 m (28 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 398,129 kg/hr, 12.4 MPa/561°C (877,723 lb/hr, 1,800 psig/1,041°F)	2	0

Case B4A – Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	289 MW 12.4 MPa/561°C/561°C(1,800psig/ 1,041°F/1,041°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	320 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,560GJ/hr (1,480 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0

Case B4A – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	391,000 lpm @ 30 m (103,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 2,180 GJ/hr (2,060 MMBtu/hr) heat duty	1	0

Case B4A – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	223,000 liters (59,000 gal)	2	0
2	Slag Crusher	Roll	12 tonne/hr (13 tph)	2	0
3	Slag Depressurizer	Proprietary	12 tonne/hr (13 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	134,000 liters (36,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	61,000 liters (16,000 gal)	2	0
6	Slag Conveyor	Drag chain	12 tonne/hr (13 tph)	2	0
7	Slag Separation Screen	Vibrating	12 tonne/hr (13 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	12 tonne/hr (13 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	191,000 liters (50,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	50 lpm @ 14 m H ₂ O (10 gpm @ 46 ft H ₂ O)	2	2
11	Grey Water Storage Tank	Field erected	61,000 liters (16,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	220 lpm @ 430 m H ₂ O (60 gpm @ 1,420 ft H ₂ O)	2	2
13	Slag Storage Bin	Vertical, field erected	800 tonne (900 tons)	2	0
14	Unloading Equipment	Telescoping chute	100 tonne/hr (110 tph)	1	0

Case B4A – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	24 kV/345 kV, 290 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 47 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 29 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 4 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case B4A – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.3.7 Case B4A - Costs Estimating Results

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-55 shows a detailed breakdown of the capital costs; Exhibit 3-56 shows the owner's costs, TOC, and TASC; Exhibit 3-57 shows the initial and annual O&M costs; and Exhibit 3-58 shows the COE breakdown.

The estimated TPC of the E-GasTM gasifier with no CO₂ capture is \$2,372/kW. Process contingency represents 2.6 percent of the TPC, and project contingency is 13.6 percent. The COE is \$99.8/MWh.

Exhibit 3-55 Case B4A total plant cost details

Case:		B4A – E-Gas™ IGCC w/o CO ₂				Estimate Type:			Conceptual		
Plant Size (MW,net):		625				Cost Base:			Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
1		Coal & Sorbent Handling									
1.1	Coal Receive & Unload	\$4,394	\$0	\$2,118	\$0	\$6,511	\$651	\$0	\$1,433	\$8,595	\$14
1.2	Coal Stackout & Reclaim	\$5,678	\$0	\$1,358	\$0	\$7,035	\$704	\$0	\$1,548	\$9,287	\$15
1.3	Coal Conveyors & Yd Crush	\$5,279	\$0	\$1,343	\$0	\$6,622	\$662	\$0	\$1,457	\$8,741	\$14
1.4	Other Coal Handling	\$1,381	\$0	\$311	\$0	\$1,692	\$169	\$0	\$372	\$2,233	\$4
1.9	Coal & Sorbent Hnd. Foundations	\$0	\$2,934	\$7,668	\$0	\$10,602	\$1,060	\$0	\$2,332	\$13,994	\$22
Subtotal		\$16,731	\$2,934	\$12,797	\$0	\$32,462	\$3,246	\$0	\$7,142	\$42,850	\$69
2		Coal & Sorbent Prep & Feed									
2.1	Coal Crushing & Drying	w/2.3	\$0	w/2.3	\$0	\$0	\$0	\$0	\$0	\$0	\$0
2.2	Prepared Coal Storage & Feed	\$1,872	\$450	\$290	\$0	\$2,612	\$261	\$0	\$575	\$3,447	\$6
2.3	Slurry Prep & Feed	\$25,543	\$0	\$11,204	\$0	\$36,748	\$3,675	\$0	\$8,084	\$48,507	\$78
2.4	Misc. Coal Prep & Feed	\$1,030	\$752	\$2,216	\$0	\$3,997	\$400	\$0	\$879	\$5,277	\$8
2.9	Coal & Sorbent Feed Foundation	\$0	\$3,777	\$3,241	\$0	\$7,017	\$702	\$0	\$1,544	\$9,263	\$15
Subtotal		\$28,445	\$4,979	\$16,950	\$0	\$50,374	\$5,037	\$0	\$11,082	\$66,494	\$106
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	\$3,767	\$6,501	\$3,408	\$0	\$13,676	\$1,368	\$0	\$3,009	\$18,052	\$29
3.2	Water Makeup & Pretreating	\$715	\$74	\$393	\$0	\$1,182	\$118	\$0	\$390	\$1,690	\$3
3.3	Other Feedwater Subsystems	\$2,118	\$700	\$626	\$0	\$3,443	\$344	\$0	\$757	\$4,545	\$7
3.4	Service Water Systems	\$418	\$834	\$2,874	\$0	\$4,125	\$413	\$0	\$1,361	\$5,899	\$9
3.5	Other Boiler Plant Systems	\$2,253	\$842	\$2,072	\$0	\$5,167	\$517	\$0	\$1,137	\$6,821	\$11
3.6	FO Supply Sys & Nat Gas	\$17,737	\$713	\$660	\$0	\$19,110	\$1,911	\$0	\$4,204	\$25,225	\$40
3.7	Waste Treatment Equipment	\$967	\$0	\$599	\$0	\$1,566	\$157	\$0	\$517	\$2,240	\$4
3.8	Misc. Power Plant Equipment	\$1,268	\$169	\$659	\$0	\$2,096	\$210	\$0	\$692	\$2,997	\$5
Subtotal		\$29,243	\$9,832	\$11,290	\$0	\$50,365	\$5,036	\$0	\$12,067	\$67,468	\$108
4		Gasifier & Accessories									
4.1	Syngas Cooler Gasifier System	\$144,797	\$0	\$80,864	\$0	\$225,661	\$22,566	\$31,351	\$42,853	\$322,431	\$516
4.2	Syngas Cooler	w/4.1	\$0	w/ 4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.3	ASU & Oxidant Compression	\$174,614	\$0	w/equip.	\$0	\$174,614	\$17,461	\$0	\$19,208	\$211,284	\$338
4.4	LT Heat Recovery & FG Saturation	\$20,488	\$0	\$7,784	\$0	\$28,273	\$2,827	\$0	\$6,220	\$37,320	\$60
4.5	Misc. Gasification Equipment	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.6	Flare Stack System	\$0	\$1,778	\$719	\$0	\$2,496	\$250	\$0	\$549	\$3,295	\$5
4.8	Major Component Rigging	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.9	Gasification Foundations	\$0	\$9,517	\$5,676	\$0	\$15,193	\$1,519	\$0	\$4,178	\$20,891	\$33
Subtotal		\$339,900	\$11,295	\$95,043	\$0	\$446,238	\$44,624	\$31,351	\$73,008	\$595,220	\$952
5A		Gas Cleanup & Piping									
5A.1	MDEA-LT AGR System	\$42,958	\$0	\$36,202	\$0	\$79,161	\$7,916	\$0	\$17,415	\$104,492	\$167
5A.2	Elemental Sulfur Plant	\$11,980	\$2,335	\$15,351	\$0	\$29,667	\$2,967	\$0	\$6,527	\$39,160	\$63
5A.3	Mercury Removal	\$1,390	\$0	\$1,051	\$0	\$2,441	\$244	\$122	\$562	\$3,369	\$5
5A.4	COS Hydrolysis	\$3,852	\$0	\$4,996	\$0	\$8,849	\$885	\$0	\$1,947	\$11,680	\$19
5A.5	Particulate Removal	w/4.1	\$0	w/ 4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0

Case:		B4A – E-Gas™ IGCC w/o CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		625					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
5A.6	Blowback Gas Systems	\$601	\$337	\$188	\$0	\$1,127	\$113	\$0	\$248	\$1,487	\$2
5A.7	Fuel Gas Piping	\$0	\$984	\$644	\$0	\$1,628	\$163	\$0	\$358	\$2,149	\$3
5A.9	HGCU Foundations	\$0	\$891	\$600	\$0	\$1,491	\$149	\$0	\$492	\$2,132	\$3
	Subtotal	\$60,783	\$4,547	\$59,033	\$0	\$124,363	\$12,436	\$122	\$27,548	\$164,470	\$263
6	Combustion Turbine & Accessories										
6.1	Combustion Turbine Generator	\$104,200	\$0	\$7,506	\$0	\$111,706	\$11,171	\$5,585	\$12,846	\$141,308	\$226
6.9	Combustion Turbine Foundations	\$0	\$923	\$1,068	\$0	\$1,992	\$199	\$0	\$657	\$2,848	\$5
	Subtotal	\$104,200	\$923	\$8,575	\$0	\$113,698	\$11,370	\$5,585	\$13,504	\$144,157	\$231
7	HRSG, Ducting, & Stack										
7.1	Heat Recovery Steam Generator	\$31,180	\$0	\$6,038	\$0	\$37,218	\$3,722	\$0	\$4,094	\$45,034	\$72
7.3	Ductwork	\$0	\$2,084	\$1,461	\$0	\$3,544	\$354	\$0	\$780	\$4,679	\$7
7.4	Stack	\$4,021	\$0	\$1,500	\$0	\$5,522	\$552	\$0	\$607	\$6,681	\$11
7.9	HRSG, Duct & Stack Foundations	\$0	\$766	\$768	\$0	\$1,534	\$153	\$0	\$506	\$2,194	\$4
	Subtotal	\$35,201	\$2,849	\$9,768	\$0	\$47,818	\$4,782	\$0	\$5,987	\$58,588	\$94
8	Steam Turbine Generator										
8.1	Steam TG & Accessories	\$38,955	\$0	\$5,943	\$0	\$44,898	\$4,490	\$0	\$4,939	\$54,326	\$87
8.2	Turbine Plant Auxiliaries	\$242	\$0	\$551	\$0	\$793	\$79	\$0	\$87	\$960	\$2
8.3	Condenser & Auxiliaries	\$3,140	\$0	\$1,774	\$0	\$4,914	\$491	\$0	\$541	\$5,946	\$10
8.4	Steam Piping	\$16,507	\$0	\$7,158	\$0	\$23,665	\$2,367	\$0	\$6,508	\$32,540	\$52
8.9	TG Foundations	\$0	\$1,141	\$2,015	\$0	\$3,156	\$316	\$0	\$1,042	\$4,513	\$7
	Subtotal	\$58,844	\$1,141	\$17,441	\$0	\$77,426	\$7,743	\$0	\$13,116	\$98,285	\$157
9	Cooling Water System										
9.1	Cooling Towers	\$3,980	\$0	\$1,210	\$0	\$5,190	\$519	\$0	\$856	\$6,565	\$11
9.2	Circulating Water Pumps	\$1,923	\$0	\$137	\$0	\$2,060	\$206	\$0	\$340	\$2,605	\$4
9.3	Circ. Water System Auxiliaries	\$170	\$0	\$24	\$0	\$194	\$19	\$0	\$32	\$246	\$0
9.4	Circ. Water Piping	\$0	\$7,546	\$1,828	\$0	\$9,374	\$937	\$0	\$2,062	\$12,373	\$20
9.5	Make-up Water System	\$408	\$0	\$560	\$0	\$968	\$97	\$0	\$213	\$1,278	\$2
9.6	Component Cooling Water Sys	\$863	\$1,032	\$708	\$0	\$2,603	\$260	\$0	\$573	\$3,436	\$5
9.9	Circ. Water System Foundations	\$0	\$2,470	\$4,389	\$0	\$6,859	\$686	\$0	\$2,264	\$9,809	\$16
	Subtotal	\$7,343	\$11,048	\$8,857	\$0	\$27,248	\$2,725	\$0	\$6,340	\$36,313	\$58
10	Ash & Spent Sorbent Handling Systems										
10.1	Slag Dewatering & Cooling	\$20,022	\$0	\$9,806	\$0	\$29,828	\$2,983	\$0	\$3,281	\$36,092	\$58
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.6	Ash Storage Silos	\$660	\$0	\$714	\$0	\$1,374	\$137	\$0	\$227	\$1,738	\$3
10.7	Ash Transport & Feed Equipment	\$909	\$0	\$212	\$0	\$1,121	\$112	\$0	\$185	\$1,418	\$2
10.8	Misc. Ash Handling Equipment	\$1,368	\$1,676	\$497	\$0	\$3,542	\$354	\$0	\$584	\$4,480	\$7
10.9	Ash/Spent Sorbent Foundation	\$0	\$55	\$73	\$0	\$128	\$13	\$0	\$42	\$184	\$0
	Subtotal	\$22,959	\$1,732	\$11,302	\$0	\$35,994	\$3,599	\$0	\$4,320	\$43,913	\$70
11	Accessory Electric Plant										
11.1	Generator Equipment	\$1,156	\$0	\$1,126	\$0	\$2,282	\$228	\$0	\$251	\$2,761	\$4
11.2	Station Service Equipment	\$4,391	\$0	\$404	\$0	\$4,795	\$479	\$0	\$527	\$5,802	\$9

Case:		B4A – E-Gas™ IGCC w/o CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		625					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
11.3	Switchgear & Motor Control	\$8,105	\$0	\$1,506	\$0	\$9,611	\$961	\$0	\$1,586	\$12,158	\$19
11.4	Conduit & Cable Tray	\$0	\$4,118	\$12,694	\$0	\$16,812	\$1,681	\$0	\$4,623	\$23,116	\$37
11.5	Wire & Cable	\$0	\$7,940	\$4,831	\$0	\$12,770	\$1,277	\$0	\$3,512	\$17,559	\$28
11.6	Protective Equipment	\$0	\$798	\$2,961	\$0	\$3,759	\$376	\$0	\$620	\$4,755	\$8
11.7	Standby Equipment	\$275	\$0	\$274	\$0	\$548	\$55	\$0	\$90	\$694	\$1
11.8	Main Power Transformers	\$17,611	\$0	\$173	\$0	\$17,785	\$1,778	\$0	\$2,934	\$22,497	\$36
11.9	Electrical Foundations	\$0	\$181	\$493	\$0	\$674	\$67	\$0	\$223	\$964	\$2
Subtotal		\$31,538	\$13,036	\$24,462	\$0	\$69,036	\$6,904	\$0	\$14,367	\$90,306	\$144
12		Instrumentation & Control									
12.1	IGCC Control Equipment	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.2	Combustion Turbine Control	w/6.1	\$0	w/6.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.3	Steam Turbine Control	w/8.1	\$0	w/8.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.4	Other Major Component Control	\$1,186	\$0	\$808	\$0	\$1,994	\$199	\$100	\$344	\$2,637	\$4
12.5	Signal Processing Equipment	w/12.7	\$0	w/12.7	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.6	Control Boards, Panels & Racks	\$273	\$0	\$178	\$0	\$451	\$45	\$23	\$104	\$622	\$1
12.7	Computer & Accessories	\$6,326	\$0	\$207	\$0	\$6,533	\$653	\$327	\$751	\$8,264	\$13
12.8	Instrument Wiring & Tubing	\$0	\$2,435	\$4,610	\$0	\$7,045	\$705	\$352	\$2,025	\$10,127	\$16
12.9	Other I & C Equipment	\$4,229	\$0	\$2,095	\$0	\$6,324	\$632	\$316	\$1,091	\$8,364	\$13
Subtotal		\$12,014	\$2,435	\$7,898	\$0	\$22,347	\$2,235	\$1,117	\$4,315	\$30,014	\$48
13		Improvements to Site									
13.1	Site Preparation	\$0	\$118	\$2,672	\$0	\$2,790	\$279	\$0	\$921	\$3,990	\$6
13.2	Site Improvements	\$0	\$2,090	\$2,955	\$0	\$5,046	\$505	\$0	\$1,665	\$7,215	\$12
13.3	Site Facilities	\$3,746	\$0	\$4,205	\$0	\$7,951	\$795	\$0	\$2,624	\$11,370	\$18
Subtotal		\$3,746	\$2,208	\$9,833	\$0	\$15,787	\$1,579	\$0	\$5,210	\$22,575	\$36
14		Buildings & Structures									
14.1	Combustion Turbine Area	\$0	\$317	\$179	\$0	\$496	\$50	\$0	\$109	\$655	\$1
14.2	Steam Turbine Building	\$0	\$2,917	\$4,153	\$0	\$7,069	\$707	\$0	\$1,166	\$8,943	\$14
14.3	Administration Building	\$0	\$1,008	\$731	\$0	\$1,740	\$174	\$0	\$287	\$2,201	\$4
14.4	Circulation Water Pump House	\$0	\$199	\$105	\$0	\$304	\$30	\$0	\$50	\$384	\$1
14.5	Water Treatment Buildings	\$0	\$588	\$573	\$0	\$1,161	\$116	\$0	\$192	\$1,468	\$2
14.6	Machine Shop	\$0	\$516	\$353	\$0	\$869	\$87	\$0	\$143	\$1,099	\$2
14.7	Warehouse	\$0	\$833	\$538	\$0	\$1,371	\$137	\$0	\$226	\$1,734	\$3
14.8	Other Buildings & Structures	\$0	\$499	\$388	\$0	\$888	\$89	\$0	\$195	\$1,172	\$2
14.9	Waste Treating Building & Str.	\$0	\$1,116	\$2,131	\$0	\$3,247	\$325	\$0	\$714	\$4,286	\$7
Subtotal		\$0	\$7,993	\$9,151	\$0	\$17,144	\$1,714	\$0	\$3,083	\$21,942	\$35
Total		\$750,947	\$76,953	\$302,400	\$0	\$1,130,300	\$113,030	\$38,175	\$201,088	\$1,482,594	\$2,372

Exhibit 3-56 Case B4A owner's costs

Description	\$/1,000	\$/kW
Pre-Production Costs		
6 Months All Labor	\$14,793	\$24
1 Month Maintenance Materials	\$3,377	\$5
1 Month Non-fuel Consumables	\$399	\$1
1 Month Waste Disposal	\$434	\$1
25% of 1 Months Fuel Cost at 100% CF	\$2,877	\$5
2% of TPC	\$29,652	\$47
Total	\$51,532	\$82
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	\$23,491	\$38
0.5% of TPC (spare parts)	\$7,413	\$12
Total	\$30,904	\$49
Other Costs		
Initial Cost for Catalyst and Chemicals	\$2,218	\$4
Land	\$900	\$1
Other Owner's Costs	\$222,389	\$356
Financing Costs	\$40,030	\$64
Total Overnight Costs (TOC)	\$1,830,567	\$2,929
TASC Multiplier (IOU, high-risk, 35 year)	1.140	
Total As-Spent Cost (TASC)	\$2,086,847	\$3,339

Exhibit 3-57 Case B4A initial and annual operating and maintenance costs

Case:	B4A – E-Gas™ IGCC w/o CO ₂			Cost Base:	Jun 2011
Plant Size (MW _{net}):	625	Heat Rate-net (Btu/kWh):	8,585	Capacity Factor (%):	80
Operating & Maintenance Labor					
Operating Labor			Operating Labor Requirements per Shift		
Operating Labor Rate (base):	39.70	\$/hour	Skilled Operator:	2.0	
Operating Labor Burden:	30.00	% of base	Operator:	9.0	
Labor O-H Charge Rate:	25.00	% of labor	Foreman:	1.0	
			Lab Tech's, etc.:	3.0	
			Total:	15.0	
Fixed Operating Costs					
			Annual Cost		
			(\$)	(\$/kW-net)	
Annual Operating Labor:				\$6,781,554	\$10.851
Maintenance Labor:				\$16,886,458	\$27.019
Administrative & Support Labor:				\$5,917,003	\$9.467
Property Taxes and Insurance:				\$29,651,871	\$47.444
Total:				\$59,236,887	\$94.781
Variable Operating Costs					
			(\$)	(\$/MWh-net)	
Maintenance Material:				\$32,421,135	\$7.40220
Consumables					
	Consumption			Cost (\$)	
	Initial Fill	Per Day	Per Unit	Initial Fill	
Water (/1000 gallons):	0	3,155	\$1.67	\$0	\$1,542,265 \$0.35212
Makeup and Waste Water Treatment Chemicals (lbs):	0	18,798	\$0.27	\$0	\$1,470,139 \$0.33565
Carbon (Mercury Removal) (lb):	115,489	198	\$5.50	\$635,191	\$317,595 \$0.07251
COS Catalyst (m³):	327	0.22	\$3,751.70	\$1,227,074	\$245,415 \$0.05603
MDEA Solution (gal):	26,146	37	\$13.62	\$356,095	\$146,323 \$0.03341
Claus Catalyst (ft³):	w/equip.	1.91	\$203.15	\$0	\$113,350 \$0.02588
Subtotal:				\$2,218,360	\$3,835,086 \$0.87560
Waste Disposal					
Spent Mercury Catalyst (lb.):	0	198	\$0.65	\$0	\$37,539 \$0.00857
Flyash (ton):	0	0	\$0.00	\$0	\$0 \$0.00000
Slag (ton):	0	563	\$25.11	\$0	\$4,130,607 \$0.94308
Subtotal:				\$0	\$4,168,147 \$0.95165
By-Products					
Sulfur (tons):	0	138	\$0.00	\$0	\$0 \$0.00000
Subtotal:				\$0	\$0 \$0.00000
Variable Operating Costs Total:				\$2,218,360	\$40,424,368 \$9.22946
Fuel Cost					
Illinois Number 6 (ton):	0	5,519	\$68.54	\$0	\$110,464,984 \$25.22072
Total:				\$0	\$110,464,984 \$25.22072

Exhibit 3-58 Case B4A COE breakdown

Component	Value, \$/MWh	Percentage
Capital	51.8	52%
Fixed	13.5	14%
Variable	9.2	9%
Fuel	25.2	25%
Total (Excluding T&S)	99.8	N/A
CO ₂ T&S	0.0	0%
Total (Including T&S)	99.8	N/A

3.3.8 Case B4B - E-Gas™ IGCC Power Plant with CO₂ Capture

This case is configured to produce electric power with CO₂ capture. The plant configuration is the same as Case B4A, namely two gasifier trains, two state-of-the-art 2008 F-class turbines, two HRSGs, and one steam turbine. The gross power output from the plant is constrained by the capacity of the two CTs, and since the CO₂ capture and compression process increases the auxiliary load on the plant, the net output is significantly reduced relative to Case B4A.

The process description for Case B4B is similar to Case B4A with several notable exceptions to accommodate CO₂ capture. A BFD and stream tables for Case B4B are shown in Exhibit 3-59 and Exhibit 3-60, respectively. Instead of repeating the entire process description, only differences from Case B4A are reported here.

3.3.8.1 Coal Preparation and Feed Systems

No differences from Case B4A.

3.3.8.2 Gasification

The gasification process is the same as Case B4A with the exception that total coal feed to the two gasifiers is 5,271 tonnes/day (5,811 tpd) (stream 8) and the ASU provides 4,234 tonnes/day (4,668 tpd) of 95 mol% oxygen to the gasifier and Claus plant (streams 5 and 3).

3.3.8.3 Raw Gas Cooling/Particulate Removal

Raw gas cooling and particulate removal are the same as Case B4A with the exception that approximately 418,710 kg/hr (923,082 lb/hr) of saturated steam at 13.8 MPa (2,000 psia) is generated in the SGC.

3.3.8.4 Syngas Scrubber/Sour Water Stripper

No differences from Case B4A.

3.3.8.5 Sour Gas Shift

The SGS process was described in Section 3.1.3. In Case B4B steam (stream 11) is added to the syngas exiting the scrubber to adjust the H₂O:CO molar ratio to approximately 2.25:1 prior to the first SGS reactor. The hot syngas exiting the first stage of SGS is used to preheat a portion of the water used to humidify the clean syngas leaving the AGR. The final stage of SGS brings the overall conversion of the CO to CO₂ to 98.5 percent. The syngas exiting the final stage of SGS still contains 1.2 vol% CH₄, which is subsequently oxidized to CO₂ in the CT and results in a carbon capture of 89.9 percent. The warm syngas exiting the second stage of the SGS at 204°C (400°F) (stream 12) is cooled to 201°C (393°F) by preheating the syngas entering the first SGS reactor. The SGS catalyst also serves to hydrolyze COS thus eliminating the need for a separate COS hydrolysis reactor. The syngas is further dehydrated and cooled to 35°C (95°F) in syngas coolers prior to the mercury removal beds.

3.3.8.6 Mercury Removal and Acid Gas Removal

Mercury removal is the same as in Case B4A.

The AGR process in Case B4B is a two-stage Selexol process where H₂S is removed in the first stage and CO₂ in the second stage of absorption, as previously described in Section 3.1.5. The process results in three product streams: the clean syngas, a CO₂-rich stream, and an acid gas feed to the Claus plant. The acid gas (stream 19) contains 21.5 percent H₂S and 70 percent CO₂

with the balance primarily H₂. The CO₂-rich stream is discussed further in the CO₂ compression section.

3.3.8.7 CO₂ Compression and Dehydration

CO₂ from the AGR process is flashed at three pressure levels to separate CO₂ and decrease H₂ losses to the CO₂ product pipeline. The HP CO₂ stream is flashed at 2.0 MPa (289.7 psia), compressed, and recycled back to the CO₂ absorber. The MP CO₂ stream is flashed at 1.0 MPa (149.7 psia). The LP CO₂ stream is flashed at 0.1 MPa (16.7 psia), compressed to 1.0 MPa (149.5 psia), and combined with the MP CO₂ stream. The combined stream is compressed from 1.0 MPa (149.5 psia) to an SC condition at 15.3 MPa (2215 psia) using a multiple-stage, intercooled compressor. During compression, the CO₂ stream is dehydrated to a dew point of -40°C (-40°F) with triethylene glycol. The raw CO₂ stream from the Selexol process contains over 99 percent CO₂. The CO₂ (stream 18) is transported to the plant fence line and is sequestration ready. CO₂ T&S costs were estimated using the methodology described in Section 2.6.

3.3.8.8 Claus Unit

The Claus plant is the same as Case B4A with the following exceptions:

- 5,494 kg/hr (12,112 lb/hr) of sulfur (stream 20) are produced
- The waste heat boiler generates 12,679 kg/hr (27,953 lb/hr) of 3.0 MPa (430 psia) steam, which provides all of the Claus plant process needs and provides some additional steam to the medium pressure steam header

3.3.8.9 Power Block

Clean syngas from the AGR plant is partially humidified to 4 percent because the nitrogen available from the ASU is insufficient to provide adequate dilution. The moisturized syngas is reheated (stream 17) to 193°C (380°F) using HP BFW, diluted with nitrogen (stream 4), and then enters the CT burner. There is no integration between the CT and the ASU in this case. The exhaust gas (stream 23) exits the CT at 562°C (1044°F) and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 132°C (270°F) (stream 24) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available steam turbine using a 12.4 MPa/534°C/534°C (1,800 psig/994°F/994°F) steam cycle.

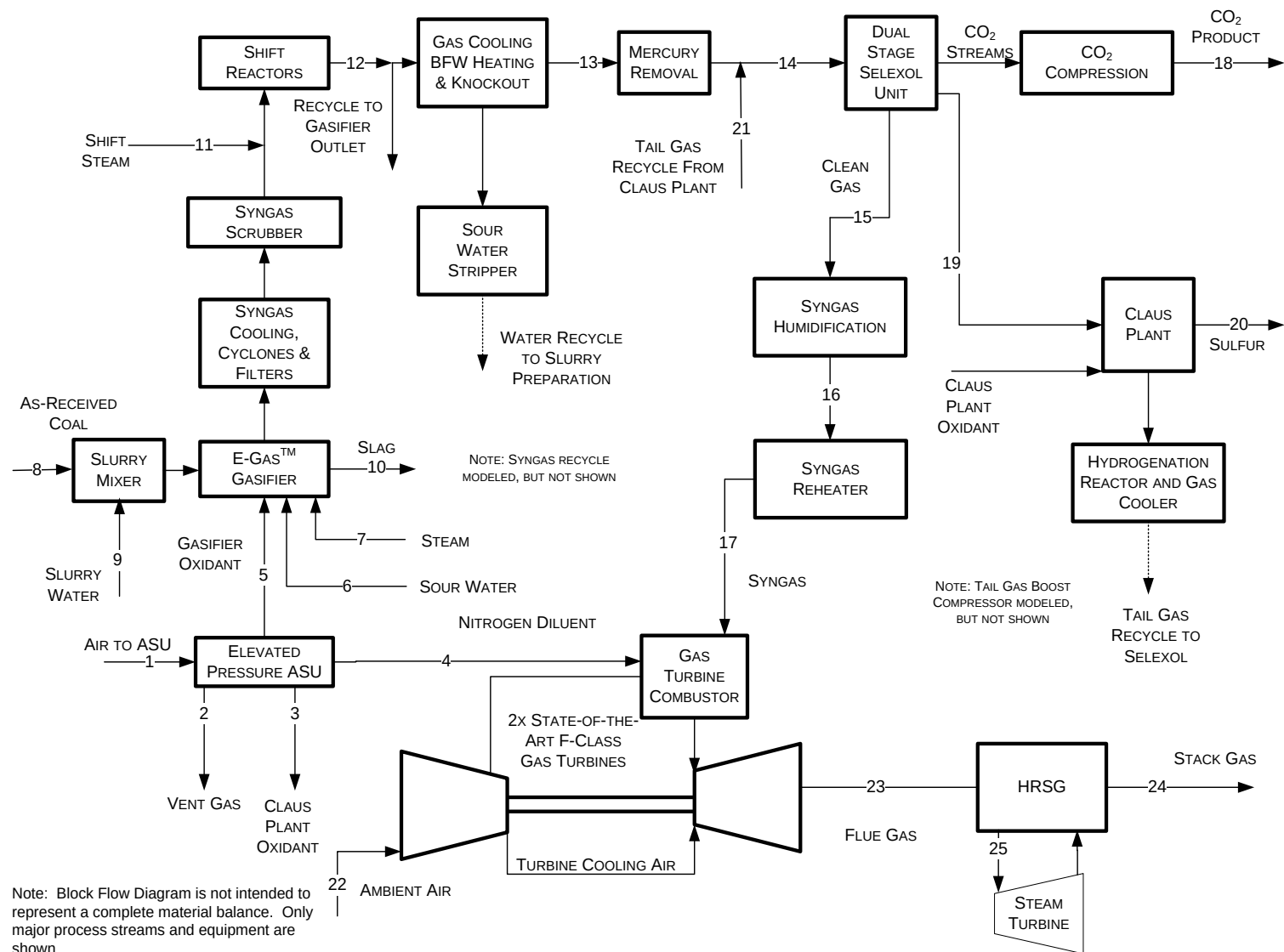
3.3.8.10 Air Separation Unit (ASU)

The elevated pressure ASU is the same as in other cases and produces 4,234 tonnes/day (4,668 tpd) of 95 mol% oxygen and 14,230 tonnes/day (15,686 tpd) of nitrogen. There is no integration between the ASU and the CT for the CO₂ capture case.

3.3.8.11 Balance of Plant

Balance of plant items were covered in Sections 3.1.9, 3.1.10, and 3.1.12.

Exhibit 3-59 Case B4B block flow diagram, E-Gas™ IGCC with CO₂ capture



Source: NETL

Exhibit 3-60 Case B4B stream table, E-Gas™ IGCC with capture

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0209	0.0318	0.0023	0.0318	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0054	0.0071	0.0071
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0125	0.0164	0.0159
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0028	0.0037	0.0038
CO ₂	0.0003	0.0071	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3103	0.4090	0.4172
COS	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
H ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4128	0.5441	0.5341
H ₂ O	0.0099	0.1780	0.0000	0.0003	0.0000	1.0000	1.0000	0.0000	0.9963	0.0000	1.0000	0.2376	0.0015	0.0015
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0048	0.0063	0.0062
N ₂	0.7732	0.6187	0.0178	0.9919	0.0178	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0090	0.0119	0.0142
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0037	0.0000	0.0000	0.0046	0.0000	0.0000
O ₂	0.2074	0.1754	0.9504	0.0054	0.9504	0.0000	0.0000	0.0000	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	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	26,685	1,231	145	19,704	5,338	1,287	4,969	0	5,009	0	9,357	37,866	26,948	27,761
V-L Flowrate (kg/hr)	770,042	33,603	4,654	552,893	171,782	23,193	89,523	0	90,226	0	168,565	748,369	547,649	575,208
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	219,635	0	22,418	0	0	0	0
Temperature (°C)	15	19	32	93	32	154	343	15	171	1,038	288	204	35	34
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	5.79	5.10	0.10	5.79	4.24	5.52	4.07	3.79	3.76
Steam Table Enthalpy (kJ/kg) ^A	30.23	36.49	26.67	92.50	26.67	646.63	3,062.93	---	673.50	---	2,917.15	874.40	40.91	39.04
AspenPlus Enthalpy (kJ/kg) ^B	-97.58	-1,964.01	4.13	63.47	4.13	-15,372.01	-12,907.38	-2,116.95	-15,251.86	1,005.43	-13,053.16	-8,872.57	-8,024.73	-8,027.16
Density (kg/m ³)	1.2	1.5	11.0	24.4	11.0	857.7	20.1	---	836.0	---	25.6	20.6	30.9	31.3
V-L Molecular Weight	28.857	27.295	32.181	28.060	32.181	18.015	18.015	---	18.012	---	18.015	19.764	20.322	20.720
V-L Flowrate (lb _{mole} /hr)	58,830	2,714	319	43,440	11,768	2,838	10,955	0	11,044	0	20,628	83,479	59,411	61,202
V-L Flowrate (lb/hr)	1,697,652	74,082	10,260	1,218,920	378,715	51,133	197,365	0	198,914	0	371,622	1,649,872	1,207,359	1,268,117
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	484,212	0	49,422	0	0	0	0
Temperature (°F)	59	67	90	199	90	310	650	59	340	1,900	550	400	95	94
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	840.0	740.0	14.7	840.0	615.0	800.0	590.0	550.0	545.0
Steam Table Enthalpy (Btu/lb) ^A	13.0	15.7	11.5	39.8	11.5	278.0	1,316.8	---	289.6	---	1,254.1	375.9	17.6	16.8
AspenPlus Enthalpy (Btu/lb) ^B	-42.0	-844.4	1.8	27.3	1.8	-6,608.8	-5,549.2	-910.1	-6,557.1	432.3	-5,611.9	-3,814.5	-3,450.0	-3,451.1
Density (lb/ft ³)	0.076	0.095	0.687	1.521	0.687	53.543	1.257	---	52.192	---	1.597	1.285	1.928	1.953

^ASteam table reference conditions are 32.02°F & 0.089 psia

^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 3-60 Case B4B stream table, E-Gas™ IGCC with capture (continued)

	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction											
Ar	0.0119	0.0114	0.0114	0.0002	0.0011	0.0000	0.0068	0.0092	0.0090	0.0090	0.0000
CH ₄	0.0262	0.0251	0.0251	0.0008	0.0045	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0063	0.0060	0.0060	0.0001	0.0007	0.0000	0.0076	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0352	0.0338	0.0338	0.9945	0.7021	0.0000	0.6907	0.0003	0.0082	0.0082	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.8964	0.8588	0.8588	0.0044	0.0591	0.0000	0.2007	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0001	0.0421	0.0421	0.0000	0.0162	0.0000	0.0017	0.0099	0.1246	0.1246	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.2153	0.0000	0.0026	0.0000	0.0000	0.0000	0.0000
N ₂	0.0239	0.0229	0.0229	0.0001	0.0010	0.0000	0.0901	0.7732	0.7529	0.7529	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
O ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.1052	0.1052	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	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	16,435	17,155	17,155	10,499	800	0	812	110,253	139,694	139,694	34,515
V-L Flowrate (kg/hr)	83,813	96,780	96,780	459,855	31,069	0	27,560	3,181,557	3,831,230	3,831,230	621,792
Solids Flowrate (kg/hr)	0	0	0	0	0	5,494	0	0	0	0	0
Temperature (°C)	34	108	193	51	48	176	38	15	562	132	534
Pressure (MPa, abs)	3.76	3.21	3.17	15.27	0.16	0.12	5.51	0.10	0.10	0.10	12.51
Steam Table Enthalpy (kJ/kg) ^A	196.11	903.06	1,360.50	-162.35	62.13	---	1.77	30.23	839.77	348.19	3,432.70
AspenPlus Enthalpy (kJ/kg) ^B	-3,190.25	-4,169.46	-3,712.02	-9,119.38	-7,323.59	140.70	-8,075.42	-97.58	-608.99	-1,100.57	-12,537.61
Density (kg/m ³)	7.4	5.6	4.6	641.8	2.4	5,283.7	83.6	1.2	0.4	0.9	36.7
V-L Molecular Weight	5.100	5.642	5.642	43.800	38.814	---	33.925	28.857	27.426	27.426	18.015
V-L Flowrate (lb _{mole} /hr)	36,233	37,820	37,820	23,146	1,765	0	1,791	243,066	307,972	307,972	76,092
V-L Flowrate (lb/hr)	184,776	213,363	213,363	1,013,808	68,496	0	60,759	7,014,133	8,446,417	8,446,417	1,370,817
Solids Flowrate (lb/hr)	0	0	0	0	0	12,112	0	0	0	0	0
Temperature (°F)	94	227	380	124	119	349	100	59	1,044	270	994
Pressure (psia)	545.0	465.0	460.0	2,214.7	23.7	17.3	799.5	14.7	15.2	15.2	1,814.7
Steam Table Enthalpy (Btu/lb) ^A	84.3	388.2	584.9	-69.8	26.7	---	0.8	13.0	361.0	149.7	1,475.8
AspenPlus Enthalpy (Btu/lb) ^B	-1,371.6	-1,792.5	-1,595.9	-3,920.6	-3,148.6	60.5	-3,471.8	-42.0	-261.8	-473.2	-5,390.2
Density (lb/ft ³)	0.461	0.352	0.285	40.067	0.149	329.851	5.220	0.076	0.026	0.053	2.293

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

3.3.9 Case B4B - Performance Results

The Case B4B modeling assumptions were presented previously in Section 3.3.3.

The plant produces a net output of 513 MWe at a net plant efficiency of 31.0 percent (HHV basis). Overall performance for the entire plant is summarized in Exhibit 3-61; Exhibit 3-62 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for nearly 58 percent of the auxiliary load between the main air compressor, nitrogen compressor, oxygen compressor, and ASU auxiliaries. The two-stage Selexol process and CO₂ compression account for an additional 27 percent of the auxiliary power load. The BFW pumps and cooling water system (CWPs and cooling tower fan) compose nearly 6 percent of the load, leaving 9 percent of the auxiliary load for all other systems.

Exhibit 3-61 Case B4B plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	240
Total Gross Power, MWe	704
Air Separation Unit Main Air Compressor, kWe	65,670
Oxygen Compressor, kWe	9,010
Nitrogen Compressors, kWe	35,340
CO ₂ Compression, kWe	31,380
Acid Gas Removal, kWe	19,900
Balance of Plant, kWe	28,950
Total Auxiliaries, MWe	190
Net Power, MWe	513
HHV Net Plant Efficiency (%)	31.0%
HHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	11,607 (11,002)
HHV Cold Gas Efficiency, %	81.0%
HHV Combustion Turbine Efficiency, %	36.1%
LHV Net Plant Efficiency (%)	32.2%
LHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	11,195 (10,611)
LHV Cold Gas Efficiency, %	76.9%
LHV Combustion Turbine Efficiency, %	42.4%
Steam Turbine Cycle Efficiency, %	41.5%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	8,668 (8,216)
Condenser Duty, GJ/hr (MMBtu/hr)	1,408 (1,334)
As-Received Coal Feed, kg/hr (lb/hr)	219,635 (484,212)
HHV Thermal Input, kWt	1,655,503
LHV Thermal Input, kWt	1,596,753
Raw Water Withdrawal, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.042 (11.2)
Raw Water Consumption, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.034 (9.1)
O ₂ :Coal	0.739

Exhibit 3-62 Case B4B plant power summary

Power Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	0
Steam Turbine Power, MWe	240
Total Gross Power, MWe	704
Auxiliary Load Summary	
Coal Handling, kWe	470
Coal Milling, kWe	2,260
Sour Water Recycle Slurry Pump, kWe	200
Slag Handling, kWe	1,160
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	65,670
Oxygen Compressor, kWe	9,010
Nitrogen Compressors, kWe	35,340
Feedwater Pumps, kWe	4,160
Quench Water Pump, kWe	0
Syngas Recycle Compressor, kWe	520
CO ₂ Compression, kWe	31,380
Scrubber Pumps, kWe	400
Acid Gas Removal, kWe	19,900
Combustion Turbine Auxiliaries, kWe	1,000
Steam Turbine Auxiliaries, kWe	100
Condensate Pumps, kWe	310
Circulating Water Pumps, kWe	4,700
Ground Water Pumps, kWe	520
Cooling Tower Fans, kWe	2,430
Claus Plant/TGTU Auxiliaries, kWe	250
Claus Plant TG Recycle Compressor, kWe	3,700
Miscellaneous Balance of Plant ^A , kWe	3,000
Transformer Losses, kWe	2,770
Total Auxiliaries, MWe	190
Net Power, MWe	513

^AIncludes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.3.9.1 Environmental Performance

The environmental targets for emissions of Hg, NO_x, SO₂, CO₂, and PM were presented in Section 2.3. A summary of the plant air emissions for Case B4B is presented in Exhibit 3-63. All HCl is assumed to be removed and is, therefore, not reported.

Exhibit 3-63 Case B4B air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year) ^A	kg/MWh (lb/MWh) ^B
SO ₂	0.001 (0.002)	39 (43)	0.008 (0.017)
NO _x	0.021 (0.049)	885 (976)	0.180 (0.396)
Particulates	0.003 (0.0071)	127 (141)	0.026 (0.057)
Hg	1.61E-7 (3.74E-7)	0.007 (0.007)	1.36E-6 (3.00E-6)
CO ₂	8 (20)	354,267 (390,512)	72 (158)
CO ₂ ^C	-	-	98 (217)

^ACalculations based on an 80 percent capacity factor

^BEmissions are based on gross power except where otherwise noted

^CCO₂ emissions based on net power instead of gross power

The low level of SO₂ emissions is achieved by capturing the sulfur in the gas by the two-stage Selexol AGR process. The CO₂ capture target results in the sulfur compounds being removed to a greater extent than required in the environmental targets of Section 2.3. The clean syngas exiting the AGR process has a sulfur concentration of approximately 5 ppmv. This results in a concentration in the flue gas of less than 1 ppmv. The H₂S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is hydrogenated to convert all sulfur species to H₂S, and then recycled back to the Selexol, thereby eliminating the need for a tail gas treatment unit.

NO_x emissions are limited by the use of humidification and nitrogen dilution to 15 ppmvd (NO₂ @ 15 percent O₂). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process and ultimately destroyed in the Claus plant burner. This helps lower NO_x levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a cyclone and a barrier filter in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of the mercury is captured from the syngas by dual activated carbon beds.

Ninety five percent of the CO₂ from the syngas is captured in the AGR system and compressed for sequestration. Because not all CO is converted to CO₂ in the shift reactors, the overall carbon removal is 89.9 percent.

The carbon balance for the plant is shown in Exhibit 3-64. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon leaves the plant as unburned carbon in the slag and as CO₂ in the stack gas, ASU vent gas, and the captured CO₂ product. The carbon capture efficiency is defined as one minus the amount of carbon in the stack gas relative to the total carbon in less carbon contained in the slag, represented by the following fraction:

$$\frac{\text{Carbon in Stack}}{(\text{Total Carbon In}) - (\text{Carbon in Slag})} = \left(1 - \left(\frac{30,416}{309,844 - 2,469}\right) * \right) 100 = 89.9\%$$

The high methane content of the syngas, relative to the GEE and Shell cases, prevented reaching the nominal 90 percent carbon capture. In order to achieve 90 percent capture, the two-stage Selexol CO₂ removal efficiency was increased from 92 to 95 percent.

Exhibit 3-64 Case B4B carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	140,006 (308,659)	Stack Gas	13,796 (30,416)
Air (CO ₂)	537 (1,185)	CO ₂ Product	125,522 (276,728)
		ASU Vent	105 (231)
		Slag	1,120 (2,469)
	140,543 (309,844)	Total	140,543 (309,844)

Exhibit 3-65 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, and sulfur in the CO₂ product. Sulfur in the slag is considered to be negligible.

Exhibit 3-65 Case B4B sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,505 (12,136)	Stack Gas	3 (6)
		CO ₂ Product	8 (18)
		Elemental Sulfur	5,494 (12,112)
Total	5,505 (12,136)	Total	5,505 (12,136)

Exhibit 3-66 shows the overall water balance for the plant. The exhibit is presented in an identical manner for cases B1A through B4A.

Exhibit 3-66 Case B4B water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)
Slag Handling	0.49 (128)	0.49 (128)	–	–	–
Slurry Water	1.51 (398)	1.51 (398)	–	–	–
Quench/Wash	–	0.00 (0)	–	–	–
Humidifier	0.23 (61)	0.23 (61)	–	–	–
SWS Blowdown	–	–	–	0.03 (6.78)	-0.03 (-6.78)
Condenser Makeup	4.52 (1,193)	–	4.52 (1,193)	–	4.52 (1,193)
BFW Makeup	0.21 (55)	–	0.21 (55)	–	0.21 (55)
Gasifier Steam	1.49 (395)	–	1.49 (395)	–	1.49 (395)
Shift Steam	2.81 (743)	–	2.81 (743)	–	2.81 (743)
CT Steam Dilution	–	–	–	–	–
Cooling Tower	18.32 (4,839)	1.04 (274)	17.28 (4,565)	4.12 (1,088)	13.16 (3,476)
BFW Blowdown	–	0.21 (55)	-0.21 (-55)	–	-0.21 (-55)
SWS Blowdown	–	0.26 (67.78)	-0.26 (-67.78)	–	-0.26 (-67.78)
Total	25.05 (6,618)	3.26 (861)	21.79 (5,757)	4.14 (1,095)	17.65 (4,662)

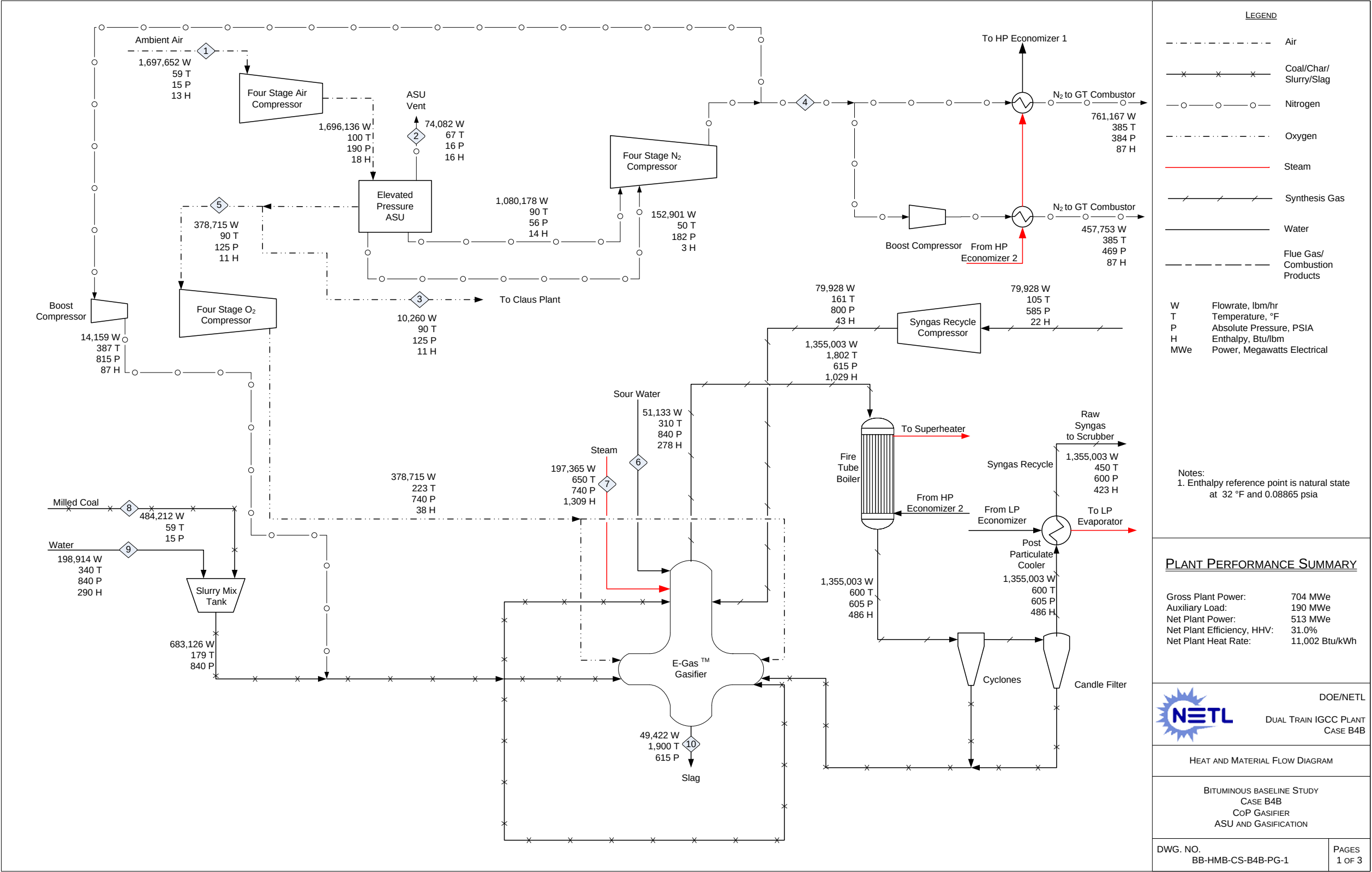
3.3.9.2 Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-67 through Exhibit 3-69:

- Coal gasification and ASU
- Syngas cleanup including sulfur recovery and tail gas recycle
- Combined cycle power generation, steam, and FW

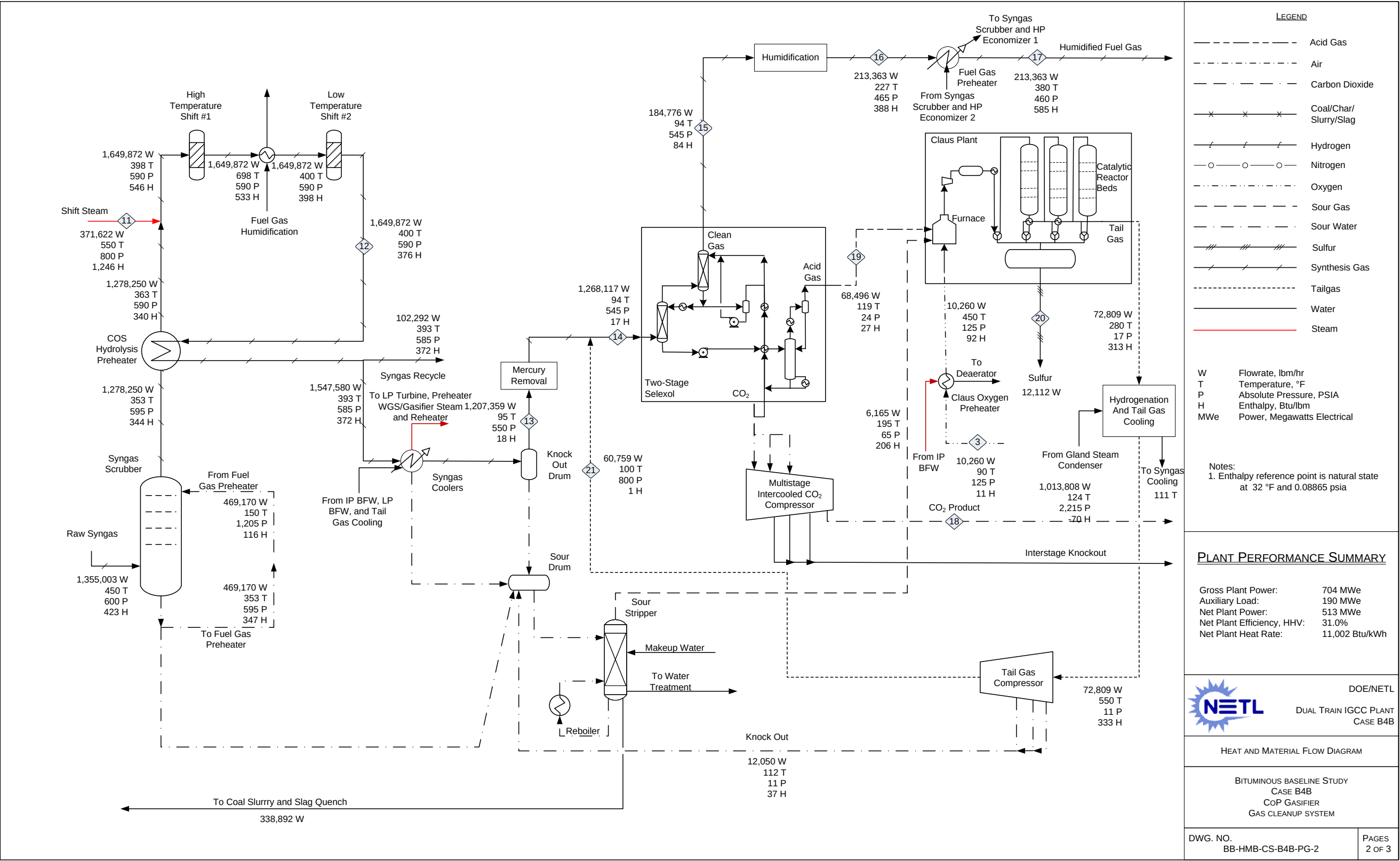
An overall plant energy balance is provided in tabular form in Exhibit 3-70. The power out is the combined CT and steam turbine power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-61) is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

Exhibit 3-67 Case B4B coal gasification and ASU heat and mass balance schematic



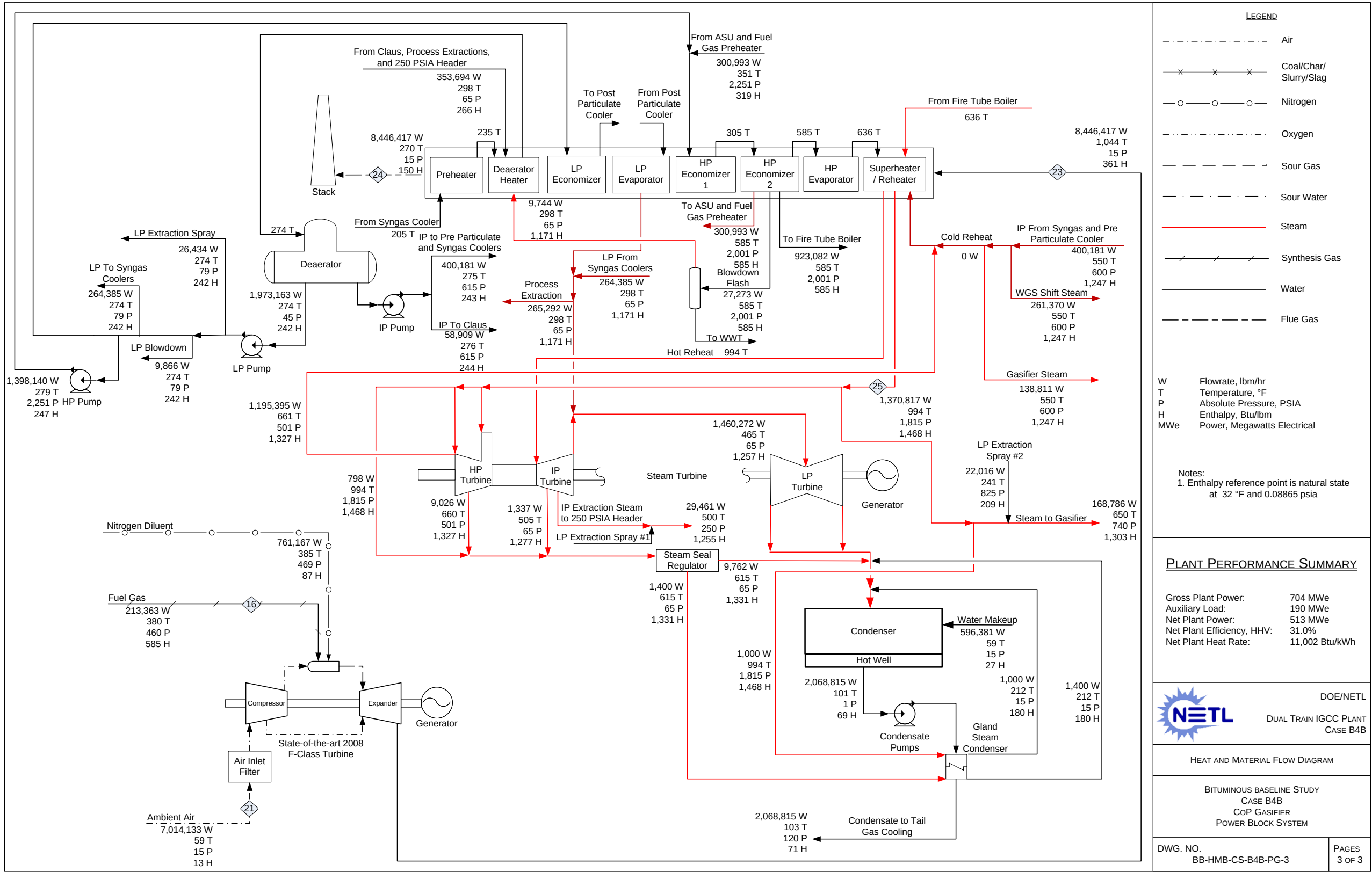
Source: NETL

Exhibit 3-68 Case B4B syngas cleanup heat and mass balance schematic



Source: NETL

Exhibit 3-69 Case B4B combined cycle power generation heat and mass balance schematic



Source: NETL

This page intentionally left blank

Exhibit 3-70 Case B4B overall energy balance (0°C [32°F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In (GJ/hr)				
Coal	5,960 (5,649)	5.0 (4.7)	–	5,965 (5,654)
Air	–	119.4 (113.2)	–	119.4 (113.2)
Raw Water Makeup	–	81.9 (77.7)	–	81.9 (77.7)
Auxiliary Power	–	–	685 (649)	685 (649)
TOTAL	5,960 (5,649)	206.4 (195.6)	685 (649)	6,851 (6,494)
Heat Out (GJ/hr)				
ASU Vent	–	1.2 (1.2)	–	1.2 (1.2)
Slag	36.7 (34.8)	25.2 (23.9)	–	61.9 (58.7)
Stack Gas	–	1,334 (1,264)	–	1,334 (1,264)
Sulfur	50.9 (48.3)	0.6 (0.6)	–	51.5 (48.8)
Motor Losses and Design Allowances	–	–	58.6 (55.6)	58.6 (55.6)
Condenser	–	1,408 (1,334)	–	1,408 (1,334)
Non-Condenser Cooling Tower Loads	–	801 (760)	–	801 (760)
CO ₂	–	-74.7 (-70.8)	–	-74.7 (-70.8)
Blowdown	–	42.7 (40.5)	–	42.7 (40.5)
CO ₂ Capture Losses	–	182.6 (173.0)	–	182.6 (173.0)
<i>Ambient Losses^A</i>	–	168.2 (159.5)	–	168.2 (159.5)
Power	–	–	2,533 (2,401)	2,533 (2,401)
TOTAL	87.6 (83.1)	3,889 (3,686)	2,592 (2,457)	6,569 (6,226)
Unaccounted Energy ^B	–	282 (268)	–	282 (268)

^AAmbient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers.

^BBy difference

3.3.10 Case B4B - Major Equipment List

Major equipment items for the E-GasTM gasifier with CO₂ capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 3.3.11. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

Case B4B – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	50 tonne (50 ton)	2	1
7	Feeder	Vibratory	180 tonne/hr (200 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	180 tonne (200 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	800 tonne (900 ton)	3	0

Case B4B – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/h (90 tph)	3	0
2	Conveyor No. 6	Belt w/ tripper	240 tonne/h (270 tph)	1	0
3	Rod Mill Feed Hopper	Dual Outlet	480 tonne (530 ton)	1	0
4	Weigh Feeder	Belt	120 tonne/h (130 tph)	2	0
5	Rod Mill	Rotary	120 tonne/h (130 tph)	2	0
6	Slurry Water Storage Tank with Agitator	Field erected	298,180 liters (78,770 gal)	2	0
7	Slurry Water Pumps	Centrifugal	830 lpm (220 gpm)	2	1
8	Trommel Screen	Coarse	170 tonne/h (190 tph)	2	0
9	Rod Mill Discharge Tank with Agitator	Field erected	390,070 liters (103,040 gal)	2	0
10	Rod Mill Product Pumps	Centrifugal	3,300 lpm (900 gpm)	2	2
11	Slurry Storage Tank with Agitator	Field erected	1,170,200 liters (309,100 gal)	2	0
12	Slurry Recycle Pumps	Centrifugal	6,500 lpm (1,700 gpm)	2	2
13	Slurry Product Pumps	Positive displacement	3,300 lpm (900 gpm)	2	2

Case B4B – Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1,100,000 liters (291,000 gal)	2	0
2	Condensate Pumps	Vertical canned	8,660 lpm @ 90 m H ₂ O (2,290 gpm @ 300 ft H ₂ O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	607,000 kg/hr (1,338,000 lb/hr)	2	0
4	Intermediate-Pressure Feedwater Pump	Horizontal centrifugal, single stage	8,540 lpm @ 30 m H ₂ O (2,260 gpm @ 90 ft H ₂ O)	2	1
5	High-Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 6,050 lpm @ 1,800 m H ₂ O (1,600 gpm @ 6,100 ft H ₂ O)	2	1
6	High-Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,300 lpm @ 220 m H ₂ O (340 gpm @ 730 ft H ₂ O)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded Screw	28 m ³ /min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m ³ /min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	441 GJ/hr (418 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	158,100 lpm @ 20 m H ₂ O (41,800 gpm @ 70 ft H ₂ O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H ₂ O (1,000 gpm @ 350 ft H ₂ O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H ₂ O (700 gpm @ 250 ft H ₂ O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	5,660 lpm @ 20 m H ₂ O (1,490 gpm @ 60 ft H ₂ O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	2,830 lpm @ 270 m H ₂ O (750 gpm @ 880 ft H ₂ O)	4	1
16	Filtered Water Pumps	Stainless steel, single suction	3,070 lpm @ 50 m H ₂ O (810 gpm @ 160 ft H ₂ O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	1,471,000 liter (389,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	1,850 lpm (490 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0

Case B4B – Account 4: Gasifier, ASU, and Accessories Including Low Temperature Heat Recovery and Fuel Gas Saturation

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized two-stage, slurry-feed entrained bed	2,900 tonne/day, 4.2 MPa (3,200 tpd, 615 psia)	2	0
2	Synthesis Gas Cooler	Fire-tube boiler	338,000 kg/hr (745,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	338,000 kg/hr (745,000 lb/hr) Design efficiency 90%	2	0
4	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0
5	Syngas Scrubber Including Sour Water Stripper	Vertical upflow	338,000 kg/hr (745,000 lb/hr)	2	0
6	Raw Gas Coolers	Shell and tube with condensate drain	386,000 kg/hr (851,000 lb/hr)	8	0
7	Raw Gas Knockout Drum	Vertical with mist eliminator	303,000 kg/hr, 35°C, 3.8 MPa (668,000 lb/hr, 95°F, 555 psia)	2	0
8	Saturation Water Economizers	Shell and tube	36 GJ/hr (34 MMBtu/hr)	2	0
9	Fuel Gas Saturator	Vertical tray tower	53,000 kg/hr, 108°C, 3.8 MPa (117,000 lb/hr, 227°F, 545 psia)	2	0
10	Saturator Water Pump	Centrifugal	700 lpm @ 13 m H ₂ O (200 gpm @ 42 ft H ₂ O)	2	2
11	Synthesis Gas Reheater	Shell and tube	53,000 kg/hr (117,000 lb/hr)	2	0
12	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	338,000 kg/hr (745,000 lb/hr) syngas	2	0
13	ASU Main Air Compressor	Centrifugal, multi-stage	6,000 m ³ /min @ 1.3 MPa (205,000 scfm @ 190 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
14	Cold Box	Vendor design	2,300 tonne/day (2,600 tpd) of 95% purity oxygen	2	0
15	Oxygen Compressor	Centrifugal, multi-stage	1,000 m ³ /min (41,000 scfm) Suction - 1.0 MPa (130 psia) Discharge - 5.1 MPa (740 psia)	2	0
16	Primary Nitrogen Compressor	Centrifugal, multi-stage	4,000 m ³ /min (134,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0
17	Secondary Nitrogen Compressor	Centrifugal, single-stage	540 m ³ /min (19,000 scfm) Suction - 1.3 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0
18	Gasifier Purge Nitrogen Boost Compressor	Centrifugal, single-stage	2,000 m ³ /min (57,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 3.2 MPa (470 psia)	2	0

Case B4B – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	301,000 kg/hr (664,000 lb/hr), 35°C (95°F), 3.8 MPa (550 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	301,000 kg/hr (664,000 lb/hr), 35°C (95°F), 3.8 MPa (550 psia)	2	0
3	Sulfur Plant	Claus type	145 tonne/day (160 tpd)	1	0
4	Water Gas Shift Reactors	Fixed bed, catalytic	206,000 kg/hr (454,000 lb/hr), 204°C (400°F), 4.1 MPa (590 psia)	4	0
5	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 94 GJ/hr (89 MMBtu/hr) Exchanger 2: 3 GJ/hr (3 MMBtu/hr)	4	0
6	Acid Gas Removal Plant	Two-stage Selexol	316,000 kg/hr (697,000 lb/hr), 34°C (94°F), 3.8 MPa (545 psia)	2	0
7	Hydrogenation Reactor	Fixed bed, catalytic	36,000 kg/hr (80,000 lb/hr), 232°C (450°F), 0.1 MPa (12.3 psia)	1	0
8	Tail Gas Recycle Compressor	Centrifugal	30,000 kg/hr (67,000 lb/hr) each	1	0
9	CO ₂ Compressor	Integrally geared, multi-stage centrifugal	1,100 m ³ /min @ 15.3 MPa (40,300 scfm @ 2,215 psia)	4	0

Case B4B – Account 6: Combustion Turbine and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	State-of-the-art 2008 F-class	230 MW	2	0
2	Combustion Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0

Case B4B – Account 7: HRSG, Ducting, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.5 m (28 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 341,986 kg/hr, 12.4 MPa/534°C (753,949 lb/hr, 1,800 psig/994°F)	2	0

Case B4B – Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	252 MW 12.4 MPa/534°C/534°C (1,800 psig/994°F/994°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	280 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,550GJ/hr (1,470 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0

Case B4B – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	472,000 lpm @ 30 m (125,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 2,630 GJ/hr (2,490 MMBtu/hr) heat duty	1	0

Case B4B – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	235,000 liters (62,000 gal)	2	0
2	Slag Crusher	Roll	12 tonne/hr (14 tph)	2	0
3	Slag Depressurizer	Proprietary	12 tonne/hr (14 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	142,000 liters (37,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	64,000 liters (17,000 gal)	2	0
6	Slag Conveyor	Drag chain	12 tonne/hr (14 tph)	2	0
7	Slag Separation Screen	Vibrating	12 tonne/hr (14 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	12 tonne/hr (14 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	201,000 liters (53,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	50 lpm @ 14 m H ₂ O (10 gpm @ 46 ft H ₂ O)	2	2
11	Grey Water Storage Tank	Field erected	64,000 liters (17,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	230 lpm @ 430 m H ₂ O (60 gpm @ 1,420 ft H ₂ O)	2	2
13	Grey Water Recycle Heat Exchanger	Shell and tube	52,200 kg/hr (30,000 lb/hr)	2	0
14	Slag Storage Bin	Vertical, field erected	890 tonne (980 tons)	2	0
15	Unloading Equipment	Telescoping chute	100 tonne/hr (110 tph)	1	0

Case B4B – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	24 kV/345 kV, 230 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 79 MVA, 3-ph, 60 Hz	2	0

4	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 51 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 8 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case B4B – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.3.11 Case B4B - Cost Estimating Results

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-71 shows a detailed breakdown of the capital costs; Exhibit 3-72 shows the owner's costs, TOC, and TASC; Exhibit 3-73 shows the initial and annual O&M costs; and Exhibit 3-74 shows the COE breakdown.

The estimated TPC of the E-GasTM gasifier with CO₂ capture is \$3,540/kW. Process contingency represents 4.1 percent of the TPC, and project contingency represents 13.9 percent. The COE, including CO₂ T&S costs of \$9.8/MWh, is \$151.8/MWh.

Exhibit 3-71 Case B4B total plant cost details

Case:		B4B – E-Gas™ IGCC w/ CO ₂				Estimate Type:			Conceptual		
Plant Size (MW,net):		514				Cost Base:			Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
1		Coal & Sorbent Handling									
1.1	Coal Receive & Unload	\$4,536	\$0	\$2,186	\$0	\$6,722	\$672	\$0	\$1,479	\$8,873	\$17
1.2	Coal Stackout & Reclaim	\$5,862	\$0	\$1,402	\$0	\$7,263	\$726	\$0	\$1,598	\$9,587	\$19
1.3	Coal Conveyors & Yd Crush	\$5,450	\$0	\$1,387	\$0	\$6,836	\$684	\$0	\$1,504	\$9,024	\$18
1.4	Other Coal Handling	\$1,426	\$0	\$321	\$0	\$1,747	\$175	\$0	\$384	\$2,306	\$4
1.9	Coal & Sorbent Hnd. Foundations	\$0	\$3,029	\$7,916	\$0	\$10,945	\$1,095	\$0	\$2,408	\$14,447	\$28
Subtotal		\$17,273	\$3,029	\$13,211	\$0	\$33,513	\$3,351	\$0	\$7,373	\$44,238	\$86
2		Coal & Sorbent Prep & Feed									
2.1	Coal Crushing & Drying	w/2.3	\$0	w/2.3	\$0	\$0	\$0	\$0	\$0	\$0	\$0
2.2	Prepared Coal Storage & Feed	\$1,937	\$465	\$300	\$0	\$2,702	\$270	\$0	\$594	\$3,566	\$7
2.3	Slurry Prep & Feed	\$26,425	\$0	\$11,591	\$0	\$38,015	\$3,802	\$0	\$8,363	\$50,180	\$98
2.4	Misc. Coal Prep & Feed	\$1,065	\$778	\$2,292	\$0	\$4,135	\$414	\$0	\$910	\$5,459	\$11
2.9	Coal & Sorbent Feed Foundation	\$0	\$3,907	\$3,352	\$0	\$7,259	\$726	\$0	\$1,597	\$9,582	\$19
Subtotal		\$29,427	\$5,150	\$17,535	\$0	\$52,112	\$5,211	\$0	\$11,465	\$68,788	\$134
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	\$3,381	\$5,834	\$3,059	\$0	\$12,274	\$1,227	\$0	\$2,700	\$16,202	\$32
3.2	Water Makeup & Pretreating	\$868	\$90	\$477	\$0	\$1,434	\$143	\$0	\$473	\$2,051	\$4
3.3	Other Feedwater Subsystems	\$1,901	\$628	\$561	\$0	\$3,090	\$309	\$0	\$680	\$4,079	\$8
3.4	Service Water Systems	\$507	\$1,012	\$3,488	\$0	\$5,007	\$501	\$0	\$1,652	\$7,160	\$14
3.5	Other Boiler Plant Systems	\$2,735	\$1,022	\$2,515	\$0	\$6,272	\$627	\$0	\$1,380	\$8,279	\$16
3.6	FO Supply Sys & Nat Gas	\$17,726	\$693	\$642	\$0	\$19,061	\$1,906	\$0	\$4,193	\$25,160	\$49
3.7	Waste Treatment Equipment	\$1,174	\$0	\$727	\$0	\$1,901	\$190	\$0	\$627	\$2,719	\$5
3.8	Misc. Power Plant Equipment	\$1,294	\$173	\$672	\$0	\$2,140	\$214	\$0	\$706	\$3,060	\$6
Subtotal		\$29,586	\$9,452	\$12,142	\$0	\$51,179	\$5,118	\$0	\$12,413	\$68,710	\$134
4		Gasifier & Accessories									
4.1	Syngas Cooler Gasifier System	\$137,262	\$0	\$76,506	\$0	\$213,769	\$21,377	\$29,564	\$40,623	\$305,333	\$594
4.2	Syngas Cooler	w/4.1	\$0	w/ 4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.3	ASU & Oxidant Compression	\$222,841	\$0	w/equip.	\$0	\$222,841	\$22,284	\$0	\$24,513	\$269,638	\$525
4.4	LT Heat Recovery & FG Saturation	\$28,817	\$0	\$10,949	\$0	\$39,766	\$3,977	\$0	\$8,749	\$52,492	\$102
4.5	Misc. Gasification Equipment	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.6	Flare Stack System	\$0	\$1,981	\$801	\$0	\$2,782	\$278	\$0	\$612	\$3,672	\$7
4.8	Major Component Rigging	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.9	Gasification Foundations	\$0	\$9,765	\$5,824	\$0	\$15,589	\$1,559	\$0	\$4,287	\$21,434	\$42
Subtotal		\$388,921	\$11,746	\$94,080	\$0	\$494,747	\$49,475	\$29,564	\$78,784	\$652,569	\$1,271
5A		Gas Cleanup & Piping									
5A.1	Double Stage Selexol	\$162,818	\$0	w/equip.	\$0	\$162,818	\$16,282	\$32,564	\$42,333	\$253,996	\$495
5A.2	Elemental Sulfur Plant	\$12,406	\$2,418	\$15,897	\$0	\$30,721	\$3,072	\$0	\$6,759	\$40,552	\$79
5A.3	Mercury Removal	\$1,867	\$0	\$1,411	\$0	\$3,278	\$328	\$164	\$754	\$4,524	\$9
5A.4	Shift Reactors	\$8,605	\$0	\$3,440	\$0	\$12,044	\$1,204	\$0	\$2,650	\$15,898	\$31
5A.5	Particulate Removal	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0

Case:		B4B – E-Gas™ IGCC w/ CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		514					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
5A.6	Blowback Gas Systems	\$657	\$369	\$206	\$0	\$1,232	\$123	\$0	\$271	\$1,626	\$3
5A.7	Fuel Gas Piping	\$0	\$926	\$606	\$0	\$1,532	\$153	\$0	\$337	\$2,022	\$4
5A.9	HGCU Foundations	\$0	\$838	\$565	\$0	\$1,403	\$140	\$0	\$463	\$2,007	\$4
	Subtotal	\$186,353	\$4,551	\$22,125	\$0	\$213,029	\$21,303	\$32,727	\$53,566	\$320,626	\$624
5B		CO₂ Compression									
5B.2	CO ₂ Compression & Drying	\$40,408	\$6,061	\$17,509	\$0	\$63,978	\$6,398	\$0	\$14,075	\$84,451	\$164
	Subtotal	\$40,408	\$6,061	\$17,509	\$0	\$63,978	\$6,398	\$0	\$14,075	\$84,451	\$164
6		Combustion Turbine & Accessories									
6.1	Combustion Turbine Generator	\$111,210	\$0	\$7,881	\$0	\$119,091	\$11,909	\$11,909	\$14,291	\$157,201	\$306
6.9	Combustion Turbine Foundations	\$0	\$923	\$1,068	\$0	\$1,992	\$199	\$0	\$657	\$2,848	\$6
	Subtotal	\$111,210	\$923	\$8,950	\$0	\$121,083	\$12,108	\$11,909	\$14,948	\$160,049	\$312
7		HRSG, Ducting, & Stack									
7.1	Heat Recovery Steam Generator	\$29,570	\$0	\$5,725	\$0	\$35,295	\$3,530	\$0	\$3,882	\$42,707	\$83
7.3	Ductwork	\$0	\$2,109	\$1,479	\$0	\$3,588	\$359	\$0	\$789	\$4,736	\$9
7.4	Stack	\$4,071	\$0	\$1,519	\$0	\$5,590	\$559	\$0	\$615	\$6,763	\$13
7.9	HRSG, Duct & Stack Foundations	\$0	\$775	\$778	\$0	\$1,553	\$155	\$0	\$512	\$2,221	\$4
	Subtotal	\$33,641	\$2,884	\$9,501	\$0	\$46,026	\$4,603	\$0	\$5,799	\$56,428	\$110
8		Steam Turbine Generator									
8.1	Steam TG & Accessories	\$36,358	\$0	\$5,336	\$0	\$41,695	\$4,169	\$0	\$4,586	\$50,450	\$98
8.2	Turbine Plant Auxiliaries	\$219	\$0	\$500	\$0	\$719	\$72	\$0	\$79	\$870	\$2
8.3	Condenser & Auxiliaries	\$3,130	\$0	\$1,764	\$0	\$4,894	\$489	\$0	\$538	\$5,922	\$12
8.4	Steam Piping	\$15,068	\$0	\$6,534	\$0	\$21,603	\$2,160	\$0	\$5,941	\$29,703	\$58
8.9	TG Foundations	\$0	\$1,034	\$1,827	\$0	\$2,861	\$286	\$0	\$944	\$4,091	\$8
	Subtotal	\$54,776	\$1,034	\$15,962	\$0	\$71,771	\$7,177	\$0	\$12,089	\$91,037	\$177
9		Cooling Water System									
9.1	Cooling Towers	\$4,540	\$0	\$1,466	\$0	\$6,006	\$601	\$0	\$991	\$7,597	\$15
9.2	Circulating Water Pumps	\$2,187	\$0	\$164	\$0	\$2,351	\$235	\$0	\$388	\$2,974	\$6
9.3	Circ. Water System Auxiliaries	\$190	\$0	\$27	\$0	\$217	\$22	\$0	\$36	\$274	\$1
9.4	Circ. Water Piping	\$0	\$8,426	\$2,041	\$0	\$10,468	\$1,047	\$0	\$2,303	\$13,817	\$27
9.5	Make-up Water System	\$480	\$0	\$660	\$0	\$1,140	\$114	\$0	\$251	\$1,505	\$3
9.6	Component Cooling Water Sys	\$963	\$1,152	\$791	\$0	\$2,907	\$291	\$0	\$639	\$3,837	\$7
9.9	Circ. Water System Foundations	\$0	\$2,765	\$4,913	\$0	\$7,678	\$768	\$0	\$2,534	\$10,980	\$21
	Subtotal	\$8,360	\$12,344	\$10,063	\$0	\$30,767	\$3,077	\$0	\$7,142	\$40,985	\$80
10		Ash & Spent Sorbent Handling Systems									
10.1	Slag Dewatering & Cooling	\$20,691	\$0	\$10,134	\$0	\$30,825	\$3,083	\$0	\$3,391	\$37,299	\$73
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.6	Ash Storage Silos	\$680	\$0	\$734	\$0	\$1,414	\$141	\$0	\$233	\$1,789	\$3
10.7	Ash Transport & Feed Equipment	\$935	\$0	\$218	\$0	\$1,154	\$115	\$0	\$190	\$1,459	\$3
10.8	Misc. Ash Handling Equipment	\$1,408	\$1,725	\$512	\$0	\$3,644	\$364	\$0	\$601	\$4,610	\$9
10.9	Ash/Spent Sorbent Foundation	\$0	\$57	\$75	\$0	\$132	\$13	\$0	\$44	\$189	\$0
	Subtotal	\$23,714	\$1,782	\$11,674	\$0	\$37,169	\$3,717	\$0	\$4,459	\$45,346	\$88

Case:		B4B – E-Gas™ IGCC w/ CO ₂				Estimate Type:			Conceptual		
Plant Size (MW,net):		514				Cost Base:			Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
11		Accessory Electric Plant									
11.1	Generator Equipment	\$1,124	\$0	\$1,095	\$0	\$2,219	\$222	\$0	\$244	\$2,685	\$5
11.2	Station Service Equipment	\$5,488	\$0	\$505	\$0	\$5,993	\$599	\$0	\$659	\$7,252	\$14
11.3	Switchgear & Motor Control	\$10,130	\$0	\$1,883	\$0	\$12,013	\$1,201	\$0	\$1,982	\$15,196	\$30
11.4	Conduit & Cable Tray	\$0	\$5,147	\$15,866	\$0	\$21,013	\$2,101	\$0	\$5,779	\$28,893	\$56
11.5	Wire & Cable	\$0	\$9,924	\$6,038	\$0	\$15,961	\$1,596	\$0	\$4,389	\$21,947	\$43
11.6	Protective Equipment	\$0	\$799	\$2,967	\$0	\$3,766	\$377	\$0	\$621	\$4,764	\$9
11.7	Standby Equipment	\$269	\$0	\$268	\$0	\$536	\$54	\$0	\$88	\$678	\$1
11.8	Main Power Transformers	\$19,358	\$0	\$167	\$0	\$19,525	\$1,952	\$0	\$3,222	\$24,699	\$48
11.9	Electrical Foundations	\$0	\$175	\$477	\$0	\$652	\$65	\$0	\$215	\$933	\$2
Subtotal		\$36,369	\$16,045	\$29,264	\$0	\$81,678	\$8,168	\$0	\$17,200	\$107,046	\$208
12		Instrumentation & Control									
12.1	IGCC Control Equipment	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.2	Combustion Turbine Control	w/6.1	\$0	w/6.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.3	Steam Turbine Control	w/8.1	\$0	w/8.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.4	Other Major Component Control	\$1,292	\$0	\$880	\$0	\$2,173	\$217	\$109	\$375	\$2,873	\$6
12.5	Signal Processing Equipment	w/12.7	\$0	w/12.7	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.6	Control Boards, Panels & Racks	\$297	\$0	\$194	\$0	\$491	\$49	\$25	\$113	\$678	\$1
12.7	Computer & Accessories	\$6,893	\$0	\$225	\$0	\$7,119	\$712	\$356	\$819	\$9,005	\$18
12.8	Instrument Wiring & Tubing	\$0	\$2,654	\$5,023	\$0	\$7,677	\$768	\$384	\$2,207	\$11,035	\$21
12.9	Other I & C Equipment	\$4,608	\$0	\$2,283	\$0	\$6,891	\$689	\$345	\$1,189	\$9,113	\$18
Subtotal		\$13,090	\$2,654	\$8,606	\$0	\$24,350	\$2,435	\$1,217	\$4,702	\$32,705	\$64
13		Improvements to Site									
13.1	Site Preparation	\$0	\$121	\$2,743	\$0	\$2,864	\$286	\$0	\$945	\$4,095	\$8
13.2	Site Improvements	\$0	\$2,146	\$3,033	\$0	\$5,179	\$518	\$0	\$1,709	\$7,406	\$14
13.3	Site Facilities	\$3,845	\$0	\$4,316	\$0	\$8,161	\$816	\$0	\$2,693	\$11,670	\$23
Subtotal		\$3,845	\$2,266	\$10,092	\$0	\$16,203	\$1,620	\$0	\$5,347	\$23,171	\$45
14		Buildings & Structures									
14.1	Combustion Turbine Area	\$0	\$317	\$179	\$0	\$496	\$50	\$0	\$109	\$655	\$1
14.2	Steam Turbine Building	\$0	\$2,691	\$3,831	\$0	\$6,521	\$652	\$0	\$1,076	\$8,250	\$16
14.3	Administration Building	\$0	\$1,042	\$756	\$0	\$1,798	\$180	\$0	\$297	\$2,274	\$4
14.4	Circulation Water Pumphouse	\$0	\$196	\$104	\$0	\$299	\$30	\$0	\$49	\$378	\$1
14.5	Water Treatment Buildings	\$0	\$713	\$696	\$0	\$1,409	\$141	\$0	\$232	\$1,782	\$3
14.6	Machine Shop	\$0	\$533	\$365	\$0	\$898	\$90	\$0	\$148	\$1,136	\$2
14.7	Warehouse	\$0	\$861	\$556	\$0	\$1,417	\$142	\$0	\$234	\$1,792	\$3
14.8	Other Buildings & Structures	\$0	\$516	\$401	\$0	\$917	\$92	\$0	\$202	\$1,211	\$2
14.9	Waste Treating Building & Str.	\$0	\$1,153	\$2,202	\$0	\$3,356	\$336	\$0	\$738	\$4,430	\$9
Subtotal		\$0	\$8,023	\$9,089	\$0	\$17,112	\$1,711	\$0	\$3,086	\$21,909	\$43
Total		\$976,972	\$87,945	\$289,801	\$0	\$1,354,718	\$135,472	\$75,418	\$252,447	\$1,818,056	\$3,540

Exhibit 3-72 Case B4B owner's costs

Description	\$/1,000	\$/kW
Pre-Production Costs		
6 Months All Labor	\$16,220	\$32
1 Month Maintenance Materials	\$3,663	\$7
1 Month Non-fuel Consumables	\$690	\$1
1 Month Waste Disposal	\$459	\$1
25% of 1 Months Fuel Cost at 100% CF	\$3,028	\$6
2% of TPC	\$36,361	\$71
Total	\$60,421	\$118
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	\$25,185	\$49
0.5% of TPC (spare parts)	\$9,090	\$18
Total	\$34,275	\$67
Other Costs		
Initial Cost for Catalyst and Chemicals	\$17,040	\$33
Land	\$900	\$2
Other Owner's Costs	\$272,708	\$531
Financing Costs	\$49,088	\$96
Total Overnight Costs (TOC)	\$2,252,488	\$4,386
TASC Multiplier (IOU, high-risk, 35 year)	1.140	
Total As-Spent Cost (TASC)	\$2,567,836	\$5,000

Exhibit 3-73 Case B4B initial and annual operating and maintenance costs

Case:	B4B – E-Gas™ IGCC w/ CO ₂			Cost Base:	Jun 2011	
Plant Size (MW _{net}):	514	Heat Rate-net (Btu/kWh):	11,002	Capacity Factor (%):	80	
Operating & Maintenance Labor						
Operating Labor			Operating Labor Requirements per Shift			
Operating Labor Rate (base):	39.70	\$/hour	Skilled Operator:	2.0		
Operating Labor Burden:	30.00	% of base	Operator:	10.0		
Labor O-H Charge Rate:	25.00	% of labor	Foreman:	1.0		
			Lab Tech's, etc.:	3.0		
			Total:	16.0		
Fixed Operating Costs						
			Annual Cost			
			(\$)	(\$/kW-net)		
Annual Operating Labor:				\$7,233,658	\$14.084	
Maintenance Labor:				\$18,718,137	\$36.444	
Administrative & Support Labor:				\$6,487,949	\$12.632	
Property Taxes and Insurance:				\$36,361,120	\$70.795	
Total:				\$68,800,863	\$133.955	
Variable Operating Costs						
			(\$)	(\$/MWh-net)		
Maintenance Material:				\$35,160,939	\$9.76861	
Consumables						
	Consumption				Cost (\$)	
	Initial Fill	Per Day	Per Unit	Initial Fill		
Water (/1000 gallons):	0	4,145	\$1.67	\$0	\$2,026,257	\$0.56295
Makeup and Waste Water Treatment Chemicals (lbs):	0	24,697	\$0.27	\$0	\$1,931,496	\$0.53662
Carbon (Mercury Removal) (lb):	177,789	304	\$5.50	\$977,839	\$488,920	\$0.13583
Shift Catalyst (ft³):	6,484	4.44	\$771.99	\$5,005,557	\$1,001,111	\$0.27813
Selexol Solution (gal):	300,533	98	\$36.79	\$11,056,980	\$1,055,946	\$0.29337
Claus Catalyst (ft³):	w/equip.	2.00	\$203.15	\$0	\$118,896	\$0.03303
Subtotal:				\$17,040,376	\$6,622,626	\$1.83994
Waste Disposal						
Spent Mercury Catalyst (lb.):	0	304	\$0.65	\$0	\$57,790	\$0.01606
Flyash (ton):	0	0	\$0.00	\$0	\$0	\$0.00000
Slag (ton):	0	593	\$25.11	\$0	\$4,348,431	\$1.20811
Subtotal:				\$0	\$4,406,221	\$1.22416
By-Products						
Sulfur (tons):	0	145	\$0.00	\$0	\$0	\$0.00000
Subtotal:				\$0	\$0	\$0.00000
Variable Operating Costs Total:				\$17,040,376	\$46,189,787	\$12.83271
Fuel Cost						
Illinois Number 6 (ton):	0	5,811	\$68.54	\$0	\$116,290,256	\$32.30842
Total:				\$0	\$116,290,256	\$32.30842

Exhibit 3-74 Case B4B COE breakdown

Component	Value, \$/MWh	Percentage
Capital	77.6	55%
Fixed	19.1	13%
Variable	12.8	9%
Fuel	32.3	23%
Total (Excluding T&S)	141.9	N/A
CO ₂ T&S	9.9	7%
Total (Including T&S)	151.8	N/A

3.4 General Electric Energy IGCC Cases

This section contains an evaluation of three GEE plant designs (radiant-only [with and without CO₂ capture] and quench-only [with CO₂ capture]). Cases B5A and B5B are based on the GEE gasifier in the “radiant-only” configuration and Case B5B-Q is based on the GEE gasifier in the “quench-only” configuration. GEE offers three design configurations (50):

- **Quench:** In this configuration, the hot syngas exiting the gasifier passes through a pool of water to quench the temperature to 289°C (550°F) before entering the syngas scrubber. It is the simplest and lowest capital cost design, but also the least efficient.
- **Radiant:** In this configuration, the hot syngas exiting the gasifier passes through a radiant syngas cooler where it is cooled from about 1,316°C (2,400°F) to 677°C (1,250°F), and then passes through a water quench where the syngas is further cooled to about 232°C (450°F) prior to entering the syngas scrubber. Relative to the quench configuration, the radiant design offers increased output, higher efficiency, improved reliability/availability, and results in the lowest COE. This configuration was chosen by GEE and Bechtel for the design of their reference plant.
- **Radiant-Convective:** In this configuration, the hot syngas exiting the gasifier passes through a radiant syngas cooler where it is cooled from about 1,316°C (2,400°F) to 760°C (1,400°F), then passes over a pool of water where particulate is removed but the syngas is not quenched, then through a convective syngas cooler where the syngas is further cooled to about 371°C (700°F) prior to entering additional heat exchangers or the scrubber. This configuration has the highest overall efficiency, but at the expense of highest capital cost and the lowest availability. This is the configuration used at Tampa Electric’s Polk Power Station.

Note that the radiant configuration includes a water quench and, based on functionality, would be more appropriately named radiant-quench. The term radiant is used to distinguish it from the radiant-convective configuration. Since radiant is the terminology used by GEE, it will be used throughout this report.

The balance of this section is organized in an analogous manner to Sections 3.2 and 3.3:

- Gasifier Background
- Process System Description for Case B5A
- Key Assumptions for Cases B5A, B5B, and B5B-Q
- Sparing Philosophy for Cases B5A, B5B, and B5B-Q
- Performance Results for Case B5A
- Equipment List for Case B5A
- Cost Estimates for Case B5A
- Process and System Description, Performance Results, Equipment List, and Cost Estimate for Cases B5B and B5B-Q

3.4.1 Gasifier Background

As mentioned in Sections 3.2.1 and 3.3.1, the “Coal Gasification Guidebook: Status, Application, and Technologies” report published by the EPRI provides a detailed history of the development of several types of gasifier technology, including the GEE gasifier, as well as gasifier capacity, distinguishing characteristics, and important coal characteristics. (46)

The largest GEE gasifier is the unit at Tampa Electric, which consists of the radiant-convective configuration. The daily coal-handling capacity of this unit is 2,268 tonnes (2,500 tons) of bituminous coal. The dry gas production rate is 0.19 million Nm³/hr (6.7 million scfh) with an energy content of about 1,897 million kJ/hr (HHV) (1,800 million Btu/hr). This size matches the F-Class CTs that are used at Tampa.

The GEE gasifier operates at the highest pressure of the three gasifiers in this report, 5.6 MPa (815 psia) compared to 4.2 MPa (615 psia) for E-GasTM and Shell.

The relatively high H₂/CO ratio and CO₂ content of GEE gasification fuel gas helps achieve low NO_x and CO emissions in even the higher-temperature advanced CTs. (46)

In contrast to the Shell and E-GasTM gasification technologies, the GEE gasifier has a relatively low refractory life and a high oxygen requirement. These characteristics are more prevalent in the single-stage quench design, with a lower efficiency. However, the quench design has a significantly lower capital cost for syngas cooling. (46)

Coals with low concentrations of ash and soluble salts are preferred for use with the GEE gasifiers, as high levels of ash increase the oxygen requirement and soluble salts may build-up in concentration with high levels of process condensate recycle is used. (46)

The slurry feeding also favors the use of high-rank coals, such as bituminous coal, since their low inherent moisture content increases the moisture-free solids content of the slurry and thereby reduces oxygen requirements. (46)

3.4.2 Process Description

In this section the overall GEE gasification process is described. The system description follows the BFD in Exhibit 3-75 and stream numbers reference the same exhibit. The tables in Exhibit 3-76 provide stream compositions, temperature, pressure, enthalpy, and flow rates for the numbered streams in the BFD.

3.4.2.1 Coal Grinding and Slurry Preparation

Coal receiving and handling is common to all cases and was covered in Section 3.1.1. The receiving and handling subsystem ends at the coal silo. Coal is then fed onto a conveyor by vibratory feeders located below each silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. Each hopper outlet discharges onto a weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 55 percent of the coal feed requirements of the gasifier. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is discharged through a trommel screen into the rod mill discharge tank, and then the slurry is pumped to the slurry storage tanks. The dry solids concentration of the final slurry is 63 percent. The Polk Power Station operates at a slurry concentration of 62-68 percent using bituminous coal, and ConocoPhillips presented a paper showing the slurry concentration of Illinois No. 6 coal as 63 percent. (49)

The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required depends on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended.

The equipment in the coal grinding and slurry preparation system is fabricated of materials appropriate for the abrasive environment present in the system. The tanks and agitators are rubber lined. The pumps are either rubber-lined or hardened metal to minimize erosion. Piping is fabricated of high-density polyethylene (HDPE).

3.4.2.2 Gasification

This plant utilizes two gasification trains to process a total of 5,083 tonnes/day (5,603 tpd) of Illinois No. 6 coal. Each of the 2 x 50 percent gasifiers operates at maximum capacity. The largest operating GEE gasifier is the 2,268 tonne/day (2,500 tpd) unit at Polk Power Station. However, that unit operates at about 2.8 MPa (400 psia). The gasifier in this report, which operates at 5.6 MPa (815 psia), will be able to process more coal and maintain the same gas residence time.

The slurry feed pump takes suction from the slurry run tank, and the discharge is sent to the feed injector of the GEE gasifier (stream 6). Oxygen from the ASU is vented during preparation for startup and is sent to the feed injector during normal operation. The air separation plant supplies 4,171 tonnes/day (4,597 tpd) of 95 mol% oxygen to the gasifiers (stream 5) and the Claus plant (stream 3). Carbon conversion in the gasifier is assumed to be 98 percent, including a fines recycle stream.

The gasifier vessel is a refractory-lined, HP combustion chamber. The coal slurry feedstock and oxygen are fed through a fuel injector at the top of the gasifier vessel. The coal slurry and the oxygen react in the gasifier at 5.6 MPa (815 psia) and 1,316°C (2,400°F) to produce syngas.

The syngas consists primarily of hydrogen and carbon monoxide, with lesser amounts of water vapor and carbon dioxide, and small amounts of hydrogen sulfide, COS, methane, argon, and nitrogen. The heat in the gasifier liquefies coal ash. Hot syngas and molten solids from the reactor flow downward into a radiant heat exchanger where the syngas is cooled.

3.4.2.3 Raw Gas Cooling/Particulate Removal

Syngas is cooled from 1,316°C (2,400°F) to 677°C (1,250°F) in the radiant SGC (stream 9), and the molten slag solidifies in the process. The solids collect in the water sump at the bottom of the gasifier and are removed periodically using a lock hopper system (stream 8). The waste heat from this cooling is used to generate HP steam. BFW in the tubes is saturated, and then steam and water are separated in a steam drum. Approximately 412,096 kg/hr (908,500 lb/hr) of saturated steam at 13.8 MPa (2,000 psia) is produced. This steam then forms part of the general heat recovery system that provides steam to the steam turbine.

The syngas exiting the radiant cooler is directed downwards by a dip tube into a water sump. Most of the entrained solids are separated from the syngas at the bottom of the dip tube as the syngas goes upwards through the water. The syngas exits the quench chamber saturated at a temperature of 232°C (450°F).

The slag handling system removes solids from the gasification process equipment. These solids consist of a small amount of unconverted carbon and essentially all of the ash contained in the feed coal. These solids are in the form of glass, which fully encapsulates any metals. Solids collected in the water sump below the radiant SGC are removed by gravity and forced circulation of water from the lock hopper circulating pump. The fine solids not removed from the bottom of the quench water sump remain entrained in the water circulating through the quench chamber. In

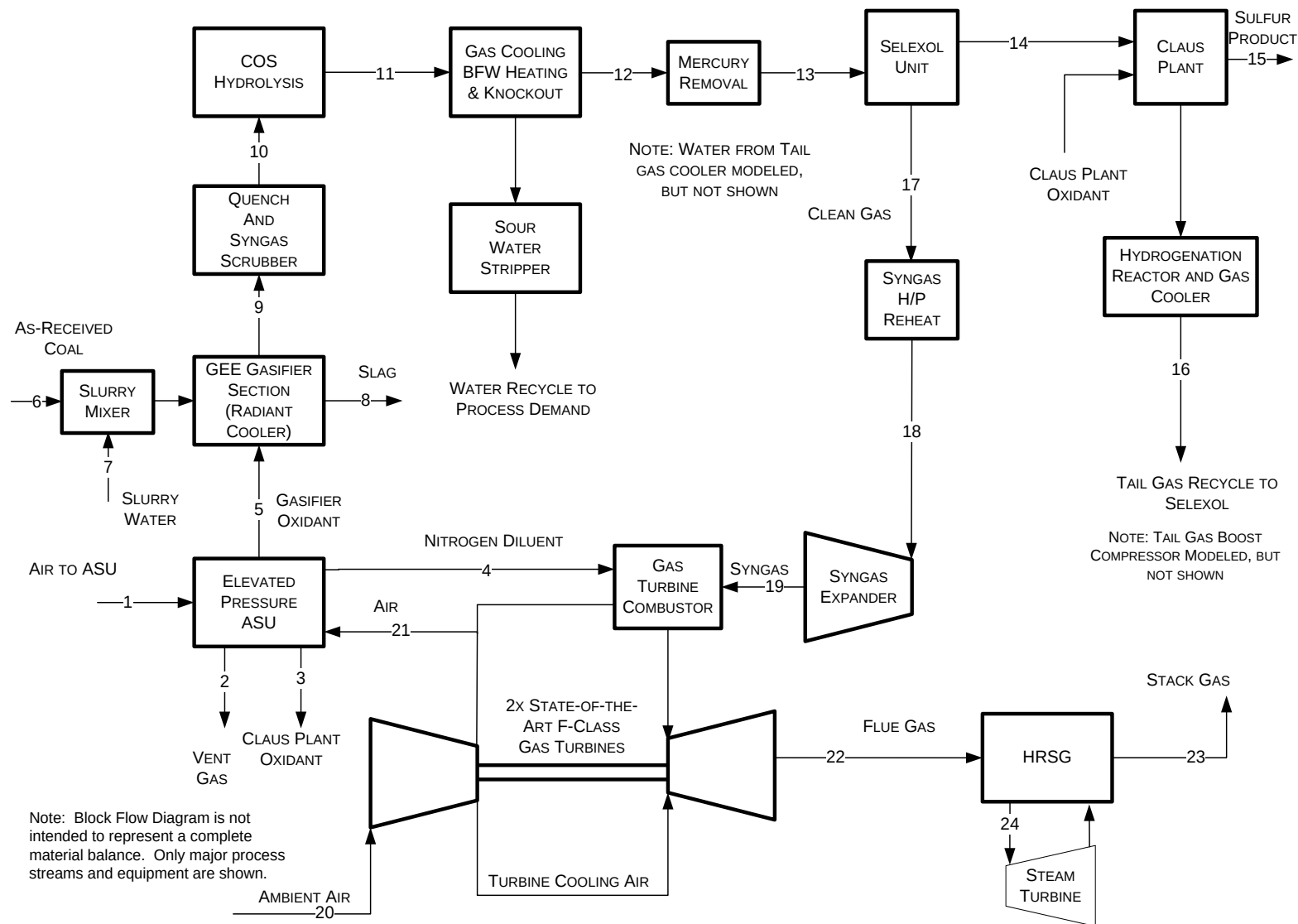
order to limit the amount of solids recycled to the quench chamber, a continuous blowdown stream is removed from the bottom of the syngas quench. The blowdown is sent to the vacuum flash drum in the black water flash section. The circulating quench water is pumped by circulating pumps to the quench gasifier.

3.4.2.4 Syngas Scrubber/Sour Water Stripper

Syngas exiting the water quench passes to a syngas scrubber where a water wash is used to remove remaining chlorides, NH_3 , SO_2 , and PM. The syngas exits the scrubber still saturated at 206°C (403°F) before it is preheated to 223°C (433°F) (stream 10) prior to entering the COS hydrolysis reactor.

The sour water stripper removes NH_3 , SO_2 , and other impurities from the scrubber and other waste streams. The stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from SGCs. Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the SRU. Remaining water is sent to wastewater treatment.

Exhibit 3-75 Case B5A block flow diagram, GEE IGCC without CO₂ capture



Source: NETL

Exhibit 3-76 Case B5A stream table, GEE IGCC without capture

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0234	0.0318	0.0023	0.0318	0.0000	0.0000	0.0000	0.0086	0.0068	0.0068	0.0099	0.0097	0.0000
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0012	0.0009	0.0009	0.0013	0.0013	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3579	0.2825	0.2825	0.4151	0.3977	0.0002
CO ₂	0.0003	0.0081	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1366	0.1078	0.1079	0.1586	0.1811	0.6124
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0001	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	0.3416	0.2696	0.2696	0.3961	0.3812	0.0000
H ₂ O	0.0099	0.2084	0.0000	0.0003	0.0000	0.0000	0.9994	0.0000	0.1358	0.3181	0.3180	0.0012	0.0012	0.0128
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0002	0.0002	0.0000	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0057	0.0059	0.0085	0.0086	0.1817
N ₂	0.7732	0.5614	0.0178	0.9919	0.0178	0.0000	0.0000	0.0000	0.0080	0.0063	0.0063	0.0092	0.0191	0.1905
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0000	0.0021	0.0020	0.0020	0.0000	0.0000	0.0000
O ₂	0.2074	0.1987	0.9504	0.0054	0.9504	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0023
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	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	21,872	1,065	102	19,382	5,298	0	4,829	0	22,212	28,142	28,142	19,153	19,992	951
V-L Flowrate (kg/hr)	631,159	28,895	3,290	543,851	170,484	0	87,000	0	446,029	552,590	552,590	390,592	422,589	36,883
Solids Flowrate (kg/hr)	0	0	0	0	0	211,781	0	23,236	0	0	0	0	0	0
Temperature (°C)	15	20	32	93	32	15	146	1,316	677	223	223	35	35	45
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	0.10	5.79	5.62	5.55	5.48	5.41	5.24	5.21	5.17
Steam Table Enthalpy (kJ/kg) ^A	30.23	36.81	26.67	92.52	26.67	---	558.58	---	1,427.78	1,067.54	1,067.44	40.35	36.92	-1.24
AspenPlus Enthalpy (kJ/kg) ^B	-97.58	-2,311.18	4.13	63.39	4.13	-2,116.95	-15,404.71	-727.24	-5,219.35	-7,386.27	-7,386.27	-5,336.38	-5,478.96	-6,427.70
Density (kg/m ³)	1.2	1.6	11.0	24.4	11.0	---	866.9	---	13.9	26.7	26.3	41.9	43.4	95.0
V-L Molecular Weight	28.857	27.130	32.181	28.060	32.181	---	18.015	---	20.081	19.636	19.636	20.393	21.138	38.795
V-L Flowrate (lb _{mole} /hr)	48,220	2,348	225	42,729	11,679	0	10,647	0	48,969	62,042	62,042	42,225	44,075	2,096
V-L Flowrate (lb/hr)	1,391,468	63,703	7,253	1,198,986	375,853	0	191,802	0	983,326	1,218,253	1,218,253	861,109	931,649	81,312
Solids Flowrate (lb/hr)	0	0	0	0	0	466,898	0	51,227	0	0	0	0	0	0
Temperature (°F)	59	68	90	199	90	59	295	2,400	1,250	433	433	95	94	112
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	14.7	840.0	815.0	805.0	795.0	785.0	760.0	755.0	750.0
Steam Table Enthalpy (Btu/lb) ^A	13.0	15.8	11.5	39.8	11.5	---	240.1	---	613.8	459.0	458.9	17.3	15.9	-0.5
AspenPlus Enthalpy (Btu/lb) ^B	-42.0	-993.6	1.8	27.3	1.8	-910.1	-6,622.8	-312.7	-2,243.9	-3,175.5	-3,175.5	-2,294.2	-2,355.5	-2,763.4
Density (lb/ft ³)	0.076	0.098	0.687	1.521	0.687	---	54.120	---	0.870	1.665	1.643	2.617	2.712	5.932

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 3-76 Case B5A stream table, GEE IGCC without capture (continued)

	15	16	17	18	19	20	21	22	23	24
V-L Mole Fraction										
Ar	0.0000	0.0040	0.0100	0.0100	0.0100	0.0092	0.0092	0.0089	0.0089	0.0000
CH ₄	0.0000	0.0000	0.0013	0.0013	0.0013	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0021	0.4089	0.4089	0.4089	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0000	0.6947	0.1562	0.1562	0.1562	0.0003	0.0003	0.0807	0.0807	0.0000
COS	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.0000	0.0415	0.3920	0.3920	0.3920	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0000	0.0018	0.0006	0.0006	0.0006	0.0099	0.0099	0.0638	0.0638	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ S	0.0000	0.0111	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N ₂	0.0000	0.2448	0.0309	0.0309	0.0309	0.7732	0.7732	0.7427	0.7427	0.0000
NH ₃	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	0.2074	0.2074	0.1039	0.1039	0.0000
SO ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	0	839	19,445	19,445	19,445	110,253	4,410	136,883	136,883	35,596
V-L Flowrate (kg/hr)	0	31,996	397,044	397,044	397,044	3,181,557	127,262	3,995,191	3,995,191	641,268
Solids Flowrate (kg/hr)	5,307	0	0	0	0	0	0	0	0	0
Temperature (°C)	174	38	45	241	196	15	432	589	132	561
Pressure (MPa, abs)	0.41	5.51	5.17	5.14	3.17	0.10	1.62	0.10	0.10	12.51
Steam Table Enthalpy (kJ/kg) ^A	---	-5.04	54.55	365.19	296.11	30.23	463.79	741.45	235.13	3,502.79
AspenPlus Enthalpy (kJ/kg) ^B	138.98	-7,219.57	-5,220.07	-4,909.43	-4,978.51	-97.58	335.98	-999.91	-1,506.23	-12,467.52
Density (kg/m ³)	5,288.2	97.7	40.0	24.1	16.4	1.2	7.9	0.4	0.9	35.1
V-L Molecular Weight	---	38.145	20.419	20.419	20.419	28.857	28.857	29.187	29.187	18.015
V-L Flowrate (lb _{mole} /hr)	0	1,849	42,869	42,869	42,869	243,066	9,723	301,776	301,776	78,475
V-L Flowrate (lb/hr)	0	70,540	875,333	875,333	875,333	7,014,133	280,565	8,807,888	8,807,888	1,413,754
Solids Flowrate (lb/hr)	11,699	0	0	0	0	0	0	0	0	0
Temperature (°F)	345	100	112	465	386	59	810	1,093	270	1,043
Pressure (psia)	59.3	799.5	750.0	745.0	460.0	14.7	234.9	15.2	15.2	1,814.7
Steam Table Enthalpy (Btu/lb) ^A	---	-2.2	23.5	157.0	127.3	13.0	199.4	318.8	101.1	1,505.9
AspenPlus Enthalpy (Btu/lb) ^B	59.8	-3,103.9	-2,244.2	-2,110.7	-2,140.4	-42.0	144.4	-429.9	-647.6	-5,360.1
Density (lb/ft ³)	330.129	6.098	2.499	1.508	1.026	0.076	0.495	0.027	0.057	2.191

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

3.4.2.5 COS Hydrolysis, Mercury Removal and Acid Gas Removal

Syngas exiting the scrubber (stream 10) passes through a COS hydrolysis reactor where about 99.5 percent of the COS is converted to CO₂ and H₂S (Section 3.1.5). The gas exiting the COS reactor (stream 11) passes through a series of heat exchangers and knockout (KO) drums to lower the syngas temperature to 35°C (95°F) and to separate entrained water. The cooled syngas (stream 12) then passes through dual carbon beds to remove approximately 97 percent of the Hg (Section 3.1.4).

Cool, particulate-free syngas (stream 13) enters the Selexol absorber unit at approximately 5.2 MPa (755 psia) and 34°C (94°F). In this absorber, H₂S is preferentially removed from the fuel gas stream along with smaller amounts of CO₂, COS, and other gases, such as hydrogen. The rich solution leaving the bottom of the absorber is heated against the lean solvent returning from the regenerator before entering the H₂S concentrator. A portion of the non-sulfur bearing absorbed gases is driven from the solvent in the H₂S concentrator using N₂ from the ASU as the stripping medium. The temperature of the H₂S concentrator overhead stream is reduced prior to entering the reabsorber where a second stage of H₂S absorption occurs. The rich solvent from the reabsorber is combined with the rich solvent from the absorber and sent to the stripper where it is regenerated through the indirect application of thermal energy via condensation of LP steam in a reboiler. The stripper acid gas stream (stream 14), consisting of 18 percent H₂S and 61 percent CO₂ (with the balance mostly N₂), is then sent to the Claus unit.

3.4.2.6 Claus Unit

Acid gas from the first-stage stripper of the Selexol unit is routed to the Claus plant. The Claus plant partially oxidizes the H₂S in the acid gas to elemental sulfur. About 5,307 kg/hr (11,699 lb/hr) of elemental sulfur (stream 15) are recovered from the fuel gas stream. This value represents an overall sulfur recovery efficiency of 99.6 percent.

Acid gas from the Selexol unit is preheated to 232°C (450°F). A portion of the acid gas along with all of the sour gas from the stripper and oxygen from the ASU are fed to the Claus furnace. In the furnace, H₂S is catalytically oxidized to SO₂ at a furnace temperature greater than 1,343°C (2,450°F), which must be maintained in order to thermally decompose all of the NH₃ present in the sour gas stream.

Following the thermal stage and condensation of sulfur, two reheater and two sulfur converters are used to obtain a per-pass H₂S conversion of approximately 99.9 percent. The Claus Plant tail gas is hydrogenated and recycled back to the Selexol process (stream 16). In the furnace waste heat boiler, 12,432 kg/hr (27,408 lb/hr) of 4.2 MPa (605 psia) steam are generated. This steam is used to satisfy all Claus process preheating and reheating requirements as well as to produce some steam for the medium-pressure steam header. The sulfur condensers produce 0.34 MPa (50 psig) of steam for the LP steam header and 2.9 MPa (415 psig) for IP steam.

3.4.2.7 Power Block

Clean syngas exiting the Selexol absorber is re-heated (stream 18) using HP BFW and then expanded to 3.2 MPa (460 psia) using an expansion turbine (stream 19). The syngas stream is diluted with nitrogen from the ASU (stream 4) and enters the state-of-the-art 2008 F-class CT burner. The CT compressor provides combustion air to the burner and also 17 percent of the air requirements in the ASU (stream 21). The exhaust gas exits the CT at 589°C (1,093°F) (stream 22) and enters the HRSG where additional heat is recovered until the flue gas exits the HRSG at

132°C (270°F) (stream 23) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced, commercially available steam turbine using a 12.4 MPa/562°C/562°C (1,800 psig/1,043°F/1,043°F) steam cycle 132°C (270°F) (stream 23) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced, commercially available steam turbine using a 12.4 MPa/562°C/562°C (1,800 psig/1,043°F/1,043°F) steam cycle.

3.4.2.8 Air Separation Unit

The elevated pressure ASU was described in Section 3.1.2. In Case B5A the ASU is designed to produce a nominal output of 4,171 tonnes/day (4,597 tpd) of 95 mol% O₂ for use in the gasifier (stream 5) and Claus plant (stream 3). The plant is designed with two production trains. The air compressor is powered by an electric motor. Approximately 13,051 tonnes/day (14,387 tpd) of nitrogen is also recovered, compressed, and used for dilution in the CT combustor (stream 4). About 4 percent of the CT air is used to supply approximately 17 percent of the ASU air requirements (stream 21) in the non-capture case.

3.4.2.9 Balance of Plant

Balance of plant items were covered in Sections 3.1.9, 3.1.10, and 3.1.12.

3.4.3 Key System Assumptions

System assumptions for cases B5A, B5B, and B5B-Q, GEE IGCC with and without CO₂ capture, are presented in Exhibit 3-77.

Exhibit 3-77 GEE IGCC plant study configuration matrix

Case	B5A	B5B	B5B-Q
Gasifier Pressure, MPa (psia)	5.6 (815)	5.6 (815)	5.6 (815)
O ₂ :Coal Ratio, kg O ₂ /kg dry coal	0.91	0.91	0.91
Carbon Conversion, %	98	98	98
Syngas HHV at Gasifier Outlet, kJ/Nm ³ (Btu/scf)	8,663 (233)	8,644 (232)	8,655 (232)
Steam Cycle, MPa/°C/°C (psig/°F/°F)	12.4/562/562 (1,800/1,043/1,043)	12.4/534/534 (1,800/994/994)	12.4/534/534 (1,800/994/994)
Condenser Pressure, mm Hg (in Hg)	51 (2.0)	51 (2.0)	51 (2.0)
Combustion Turbine	2x State-of-the-Art 2008 F-Class (232 MW output each)	2x State-of-the-Art 2008 F-Class (232 MW output each)	2x State-of-the-Art 2008 F-Class (232 MW output each)
Gasifier Technology	GEE Radiant	GEE Radiant	GEE Quench
Oxidant	95 vol% Oxygen	95 vol% Oxygen	95 vol% Oxygen
Coal	Illinois No. 6	Illinois No. 6	Illinois No. 6
Coal Slurry Solids Content, %	63	63	63
COS Hydrolysis	Yes	Occurs in SGS	Occurs in SGS
Sour Gas Shift	No	Yes	Yes
H ₂ S Separation	Selexol	Selexol 1 st Stage	Selexol 1 st Stage
Sulfur Removal, %	99.7	99.9	99.8
Sulfur Recovery	Claus Plant with Tail Gas Recycle to Selexol/ Elemental Sulfur	Claus Plant with Tail Gas Recycle to Selexol/ Elemental Sulfur	Claus Plant with Tail Gas Recycle to Selexol/ Elemental Sulfur
Particulate Control	Water Quench, Scrubber, and AGR Absorber	Water Quench, Scrubber, and AGR Absorber	Water Quench, Scrubber, and AGR Absorber
Mercury Control	Carbon Bed	Carbon Bed	Carbon Bed
NO _x Control	Multi Nozzle Quiet Combustor (MNQC) (LNB) and N ₂ Dilution	MNQC (LNB) and N ₂ Dilution	MNQC (LNB) and N ₂ Dilution
CO ₂ Separation	N/A	Selexol 2 nd Stage	Selexol 2 nd Stage
Overall Carbon Capture	N/A	90.0%	90.0%
CO ₂ Sequestration	N/A	Off-site Saline Formation	Off-site Saline Formation

3.4.3.1 Balance of Plant – Cases B5A, B5B, and B5B-Q

The balance of plant assumptions are common to all cases and are presented in Exhibit 3-11.

3.4.4 Sparing Philosophy

The sparing philosophy for cases B5A, B5B, and B5B-Q is provided below. Dual trains are used to accommodate the size of commercial CTs. There is no redundancy other than normal sparing of rotating equipment. The plant design consists of the following major subsystems:

- Two ASUs (2 x 50%)
- Two trains of slurry preparation and slurry pumps (2 x 50%)
- Two trains of gasification, including gasifier, SGC, quench and scrubber (2 x 50%)
- Two trains of syngas clean-up process (2 x 50%)
- Two trains of Selexol AGR, single-stage in Case B5A and two-stage in Case B5B, (2 x 50%) and one Claus-based SRU (1 x 100%)
- Two CO₂ compression systems (2 x 50%) in cases B5B and B5B-Q
- Two CT/HRSG tandems (2 x 50%)
- One steam turbine (1 x 100%)

3.4.5 Case B5A - Performance Results

The plant produces a net output of 622 MWe at a net plant efficiency of 39.0 percent (HHV basis). GEE has reported a net plant efficiency of 38.5 percent for their reference plant, and they also presented a range of efficiencies of 38.5-40 percent depending on fuel type. (16), (51) Typically the higher efficiencies result from fuel blends that include petroleum coke.

Overall performance for the plant is summarized in Exhibit 3-78; Exhibit 3-79 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for approximately 78 percent of the auxiliary load between the main air compressor, nitrogen compressor, oxygen compressor, and ASU auxiliaries. The cooling water system, including the CWP's and the cooling tower fan, accounts for approximately 5 percent of the auxiliary load, and the BFW pumps account for an additional 3 percent. All other individual auxiliary loads are less than 3 percent of the total.

Exhibit 3-78 Case B5A plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	8
Steam Turbine Power, MWe	276
Total Gross Power, MWe	748
Air Separation Unit Main Air Compressor, kWe	53,820
Oxygen Compressor, kWe	10,260
Nitrogen Compressors, kWe	33,350
CO ₂ Compression, kWe	0
Acid Gas Removal, kWe	2,590
Balance of Plant, kWe	25,840
Total Auxiliaries, MWe	126
Net Power, MWe	622
HHV Net Plant Efficiency (%)	39.0%
HHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	9,240 (8,758)
HHV Cold Gas Efficiency, %	79.6%
HHV Combustion Turbine Efficiency, %	37.5%
LHV Net Plant Efficiency (%)	40.4%
LHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	8,912 (8,447)
LHV Cold Gas Efficiency, %	76.2%
LHV Combustion Turbine Efficiency, %	40.6%
Steam Turbine Cycle Efficiency, %	43.3%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	8,323 (7,889)
Condenser Duty, GJ/hr (MMBtu/hr)	1,536 (1,456)
As-Received Coal Feed, kg/hr (lb/hr)	211,781 (466,898)
HHV Thermal Input, kWt	1,596,309
LHV Thermal Input, kWt	1,539,660
Raw Water Withdrawal, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.029 (7.6)
Raw Water Consumption, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.023 (6.1)
O ₂ :Coal	0.761

Exhibit 3-79 Case B5A plant power summary

Power Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	8
Steam Turbine Power, MWe	276
Total Gross Power, MWe	748
Auxiliary Load Summary	
Coal Handling, kWe	460
Coal Milling, kWe	2,180
Sour Water Recycle Slurry Pump, kWe	180
Slag Handling, kWe	1,120
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	53,820
Oxygen Compressor, kWe	10,260
Nitrogen Compressors, kWe	33,350
Feedwater Pumps, kWe	3,980
Quench Water Pump, kWe	520
Syngas Recycle Compressor, kWe	0
CO ₂ Compression, kWe	0
Scrubber Pumps, kWe	220
Acid Gas Removal, kWe	2,590
Combustion Turbine Auxiliaries, kWe	1,000
Steam Turbine Auxiliaries, kWe	100
Condensate Pumps, kWe	230
Circulating Water Pumps, kWe	4,220
Ground Water Pumps, kWe	430
Cooling Tower Fans, kWe	2,180
Claus Plant/TGTU Auxiliaries, kWe	250
Claus Plant TG Recycle Compressor, kWe	2,090
Miscellaneous Balance of Plant ^{A,B} , kWe	3,000
Transformer Losses, kWe	2,680
Total Auxiliaries, MWe	126
Net Power, MWe	622

^AIncludes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.4.5.1 Environmental Performance

The environmental targets for emissions of Hg, NO_x, SO₂, and PM were presented in Section 2.4. A summary of the plant air emissions for Case B5A is presented in Exhibit 3-80. All HCl is assumed to be removed and is, therefore, not reported.

Exhibit 3-80 Case B5A air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year) ^A	kg/MWh (lb/MWh) ^B
SO ₂	0.001 (0.001)	21 (24)	0.004 (0.009)
NO _x	0.025 (0.059)	1,023 (1,128)	0.195 (0.430)
Particulates	0.003 (0.0071)	123 (136)	0.023 (0.052)
Hg	1.77E-7 (4.12E-7)	0.007 (0.008)	1.36E-6 (3.00E-6)
CO ₂	85 (197)	3,407,878 (3,756,542)	650 (1,434)
CO ₂ ^C	-	-	782 (1,724)

^ACalculations based on an 80 percent capacity factor

^BEmissions are based on gross power except where otherwise noted

^CCO₂ emissions based on net power instead of gross power

The low level of SO₂ emissions is achieved by capturing the sulfur in the gas by the Selexol AGR process. The AGR process removes over 99 percent of the sulfur compounds in the fuel gas down to a level of less than 30 ppmv. This results in a concentration in the flue gas of less than 4 ppmv. The H₂S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is hydrogenated to convert all sulfur species to H₂S and then recycled back to the Selexol process, thereby eliminating the need for a tail gas treatment unit.

NO_x emissions are limited by nitrogen dilution of the syngas to 15 ppmvd (as NO₂ @15 percent O₂). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process and ultimately destroyed in the Claus plant burner. This helps lower NO_x levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas quench in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of the mercury is captured from the syngas by dual activated carbon beds.

CO₂ emissions represent the uncontrolled discharge from the process.

The carbon balance for the plant is shown in Exhibit 3-81. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not neglected here since the Aspen model accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag and as CO₂ in the stack gas and ASU vent gas.

Exhibit 3-81 Case B5A carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	134,999 (297,623)	Stack Gas	132,715 (292,586)
Air (CO ₂)	519 (1,143)	CO ₂ Product	0 (0)
		ASU Vent	103 (227)
		Slag	2,700 (5,952)
Total	135,518 (298,766)	Total	135,518 (298,766)

Exhibit 3-82 shows the sulfur balances for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant and sulfur emitted in the stack gas. Sulfur in the slag is considered to be negligible.

Exhibit 3-82 Case B5A sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,308 (11,702)	Stack Gas	2 (3)
		CO ₂ Product	0 (0)
		Elemental Sulfur	(11,699)
Total	5,308 (11,702)	Total	5,308 (11,702)

Exhibit 3-83 shows the overall water balance for the plant. The water balance was explained in cases B1A (Shell) and B4A (E-GasTM), but is also presented here for completeness

Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is defined as the water removed from the ground or diverted from a surface-water source for use in the plant. For this report, it was assumed to be provided 50 percent by a POTW and 50 percent from groundwater. Raw water withdrawal can be represented by the water metered from a raw water source and used in the plant processes for any and all purposes, such as cooling tower makeup, BFW makeup, quench system makeup, and slag handling makeup. 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 3-83 Case B5A water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)
Slag Handling	0.50 (133)	0.50 (133)	–	–	–
Slurry Water	1.45 (384)	1.45 (384)	–	–	–
Quench/Wash	2.75 (726)	0.90 (237)	1.85 (489)	–	1.85 (489)
Humidifier	–	–	–	–	–
SWS Blowdown	–	–	–	0.03 (7.54)	-0.03 (-7.54)
Condenser Makeup	0.20 (54)	–	0.20 (54)	–	0.20 (54)
BFW Makeup	0.20 (54)	–	0.20 (54)	–	0.20 (54)
Gasifier Steam	–	–	–	–	–
Shift Steam	–	–	–	–	–
CT Steam Dilution	–	–	–	–	–
Cooling Tower	16.43 (4,341)	0.49 (129)	15.94 (4,211)	3.70 (976)	12.25 (3,235)
BFW Blowdown	–	0.20 (54)	-0.20 (-54)	–	-0.20 (-54)
SWS Blowdown	–	0.29 (75.38)	-0.29 (-75.38)	–	-0.29 (-75.38)
Total	21.34 (5,638)	3.34 (883)	18.00 (4,755)	3.72 (984)	14.27 (3,771)

3.4.5.2 Heat and Mass Balance Diagrams

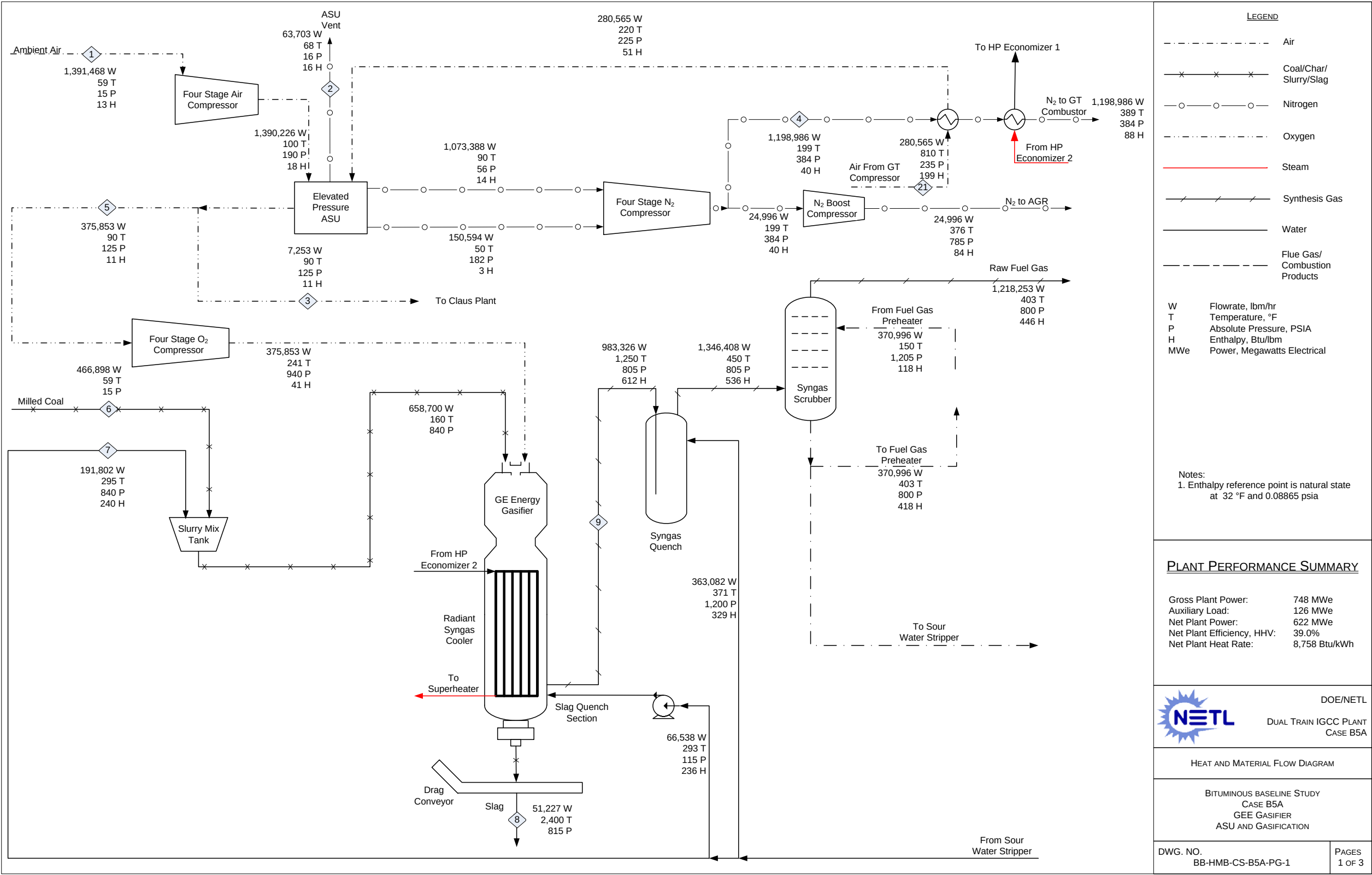
Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-84 through Exhibit 3-86:

- Coal gasification and ASU
- Syngas cleanup, sulfur recovery, and tail gas recycle
- Combined cycle power generation, steam, and FW

An overall plant energy balance is provided in tabular form in Exhibit 3-87. The power out is the combined CT, steam turbine, and expander power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-78) is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

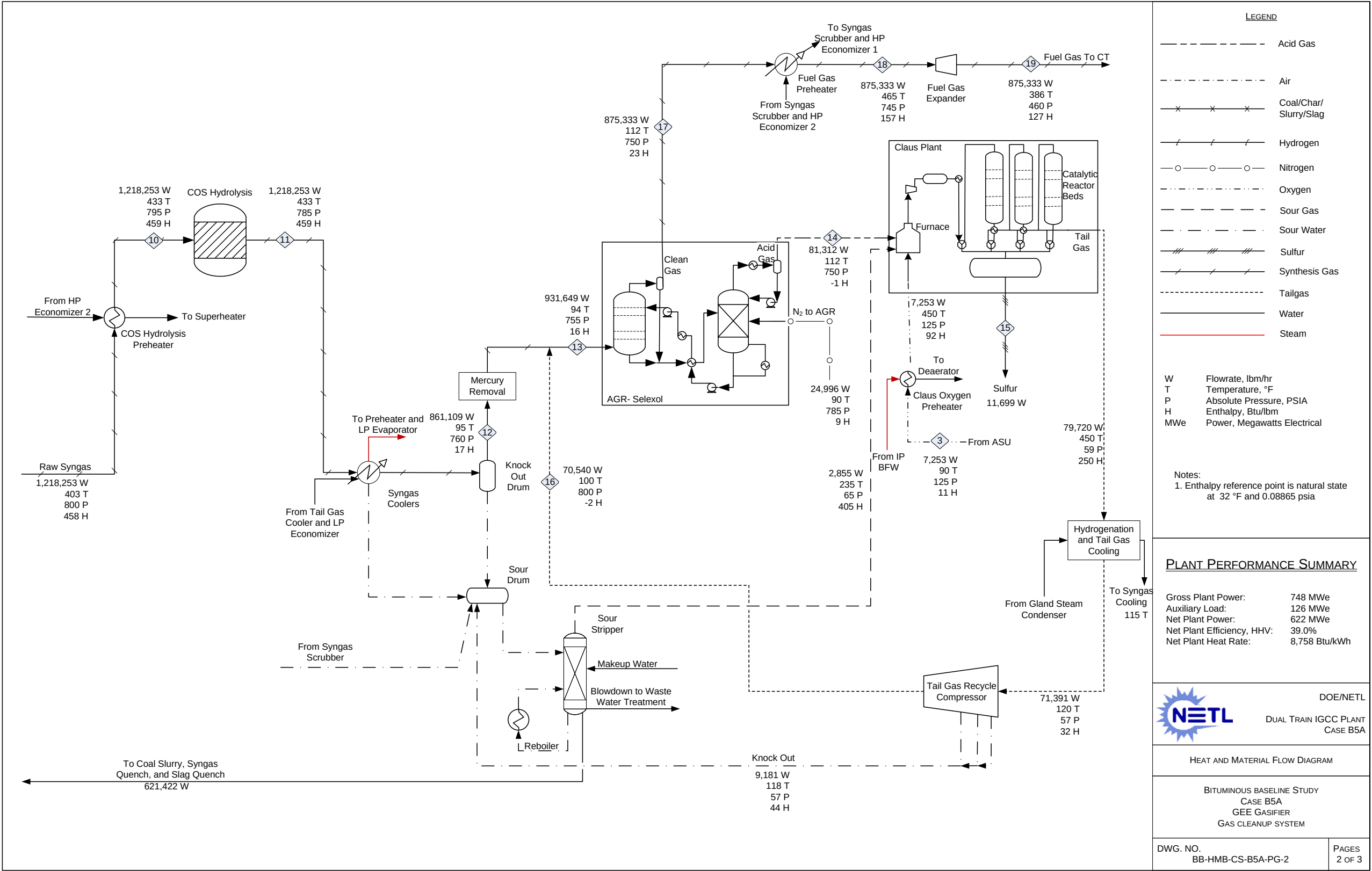
This page intentionally left blank

Exhibit 3-84 Case B5A coal gasification and ASU heat and mass balance schematic



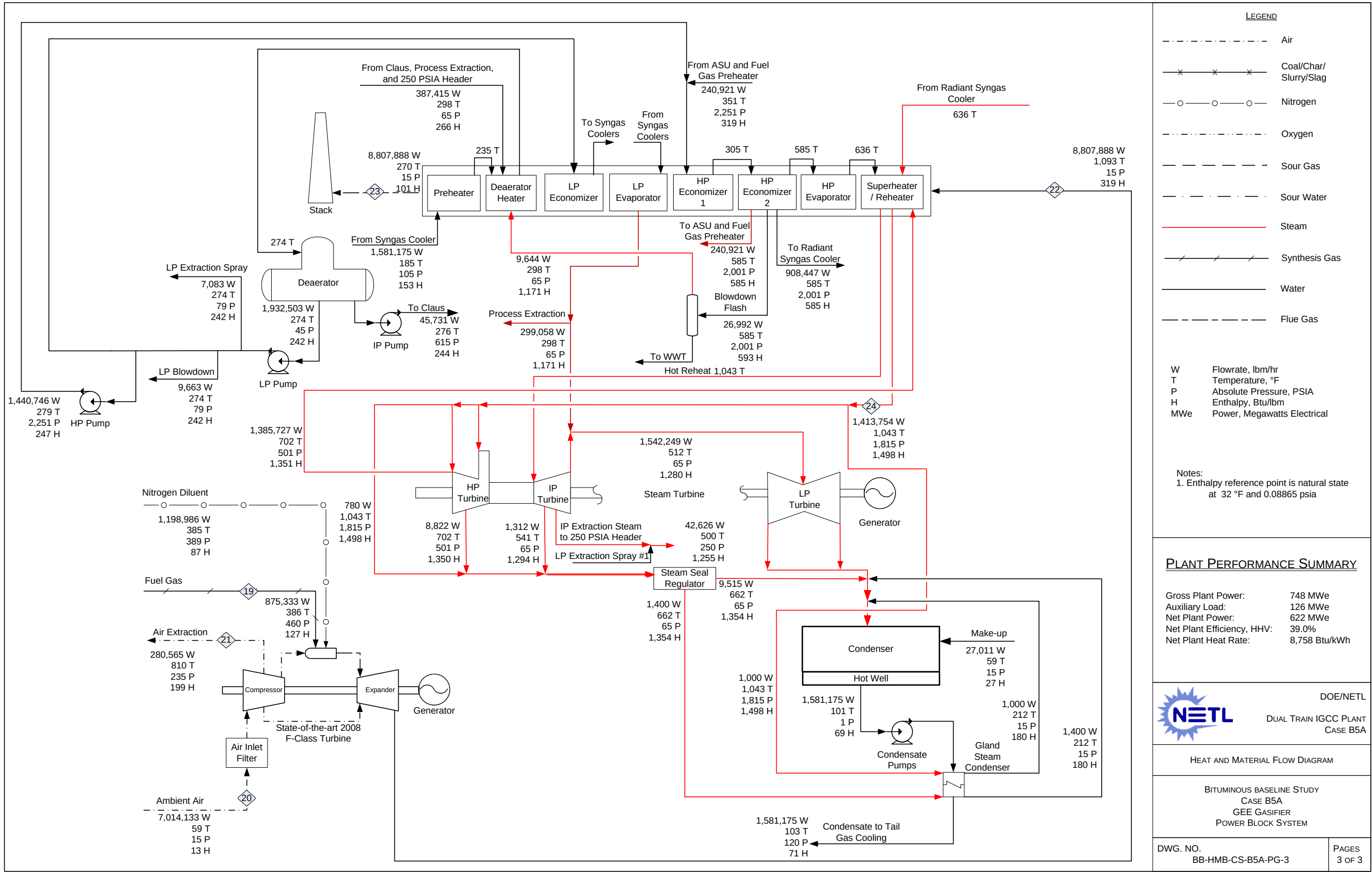
Source: NETL

Exhibit 3-85 Case B5A syngas cleanup heat and mass balance schematic



Source: NETL

Exhibit 3-86 Case B5A combined-cycle power generation heat and mass balance schematic



Source: NETL

This page intentionally left blank

Exhibit 3-87 Case B5A overall energy balance (0°C [32°F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In (GJ/hr)				
Coal	5,747 (5,447)	4.8 (4.6)	–	5,752 (5,451)
Air	–	115.2 (109.2)	–	115.2 (109.2)
Raw Water Makeup	–	67.7 (64.1)	–	67.7 (64.1)
Auxiliary Power	–	–	453 (429)	453 (429)
TOTAL	5,747 (5,447)	187.7 (177.9)	453 (429)	6,388 (6,054)
Heat Out (GJ/hr)				
ASU Vent	–	1.1 (1.0)	–	1.1 (1.0)
Slag	88.5 (83.9)	36.3 (34.4)	–	124.8 (118.3)
Stack Gas	–	939 (890)	–	939 (890)
Sulfur	49.2 (46.6)	0.6 (0.6)	–	49.8 (47.2)
Motor Losses and Design Allowances	–	–	53.9 (51.1)	53.9 (51.1)
Condenser	–	1,536 (1,456)	–	1,536 (1,456)
Non-Condenser Cooling Tower Loads	–	458 (434)	–	458 (434)
CO ₂	–	–	–	–
Blowdown	–	41.0 (38.9)	–	41.0 (38.9)
CO ₂ Capture Losses	–	–	–	–
<i>Ambient Losses^A</i>	–	157.9 (149.6)	–	157.9 (149.6)
Power	–	–	2,692 (2,552)	2,692 (2,552)
TOTAL	137.7 (130.5)	3,171 (3,005)	2,746 (2,603)	6,055 (5,739)
Unaccounted Energy ^B	–	333 (316)	–	333 (316)

^AAmbient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers.

^BBy difference.

3.4.6 Case B5A - Major Equipment List

Major equipment items for the GEE gasifier with no CO₂ capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the

cost estimates in Section 3.4.7. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

Case B5A – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	40 tonne (50 ton)	2	1
7	Feeder	Vibratory	170 tonne/hr (190 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	350 tonne/hr (390 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	170 tonne (190 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/ tripper	350 tonne/hr (390 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	350 tonne/hr (390 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	780 tonne (860 ton)	3	0

Case B5A – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/h (90 tph)	3	0
2	Conveyor No. 6	Belt w/ tripper	230 tonne/h (260 tph)	1	0
3	Rod Mill Feed Hopper	Dual Outlet	470 tonne (510 ton)	1	0
4	Weigh Feeder	Belt	120 tonne/h (130 tph)	2	0
5	Rod Mill	Rotary	120 tonne/h (130 tph)	2	0
6	Slurry Water Storage Tank with Agitator	Field erected	287,520 liters (75,950 gal)	2	0
7	Slurry Water Pumps	Centrifugal	800 lpm (210 gpm)	2	1
8	Trommel Screen	Coarse	160 tonne/h (180 tph)	2	0
9	Rod Mill Discharge Tank with Agitator	Field erected	376,120 liters (99,360 gal)	2	0
10	Rod Mill Product Pumps	Centrifugal	3,100 lpm (800 gpm)	2	2
11	Slurry Storage Tank with Agitator	Field erected	1,128,400 liters (298,100 gal)	2	0
12	Slurry Recycle Pumps	Centrifugal	6,300 lpm (1,700 gpm)	2	2
13	Slurry Product Pumps	Positive displacement	3,100 lpm (800 gpm)	2	2

Case B5A – Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1,083,000 liters (286,000 gal)	2	0
2	Condensate Pumps	Vertical canned	6,620 lpm @ 90 m H ₂ O (1,750 gpm @ 300 ft H ₂ O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	494,000 kg/hr (1,088,000 lb/hr)	2	0
4	Intermediate-Pressure Feedwater Pump	Horizontal centrifugal, single stage	8,360 lpm @ 30 m H ₂ O (2,210 gpm @ 90 ft H ₂ O)	2	1
5	High-Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 6,230 lpm @ 1,800 m H ₂ O (1,650 gpm @ 6,100 ft H ₂ O)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
6	High-Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,040 lpm @ 220 m H ₂ O (280 gpm @ 730 ft H ₂ O)	2	1
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded Screw	28 m ³ /min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m ³ /min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	252 GJ/hr (239 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	90,400 lpm @ 20 m H ₂ O (23,900 gpm @ 70 ft H ₂ O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H ₂ O (1,000 gpm @ 350 ft H ₂ O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H ₂ O (700 gpm @ 250 ft H ₂ O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	4,630 lpm @ 20 m H ₂ O (1,220 gpm @ 60 ft H ₂ O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	3,080 lpm @ 270 m H ₂ O (810 gpm @ 880 ft H ₂ O)	3	1
16	Filtered Water Pumps	Stainless steel, single suction	2,090 lpm @ 50 m H ₂ O (550 gpm @ 160 ft H ₂ O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	1,002,000 liter (265,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	300 lpm (80 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0

Case B5A – Account 4: Gasifier, ASU, and Accessories Including Low Temperature Heat Recovery and Fuel Gas Saturation

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized slurry-feed, entrained bed	2,800 tonne/day, 5.6 MPa (3,100 tpd, 815 psia)	2	0
2	Synthesis Gas Cooler	Vertical downflow radiant heat exchanger with outlet quench chamber	245,000 kg/hr (541,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	336,000 kg/hr (741,000 lb/hr) Design efficiency 90%	2	0
4	Raw Gas Coolers	Shell and tube with condensate drain	216,000 kg/hr (476,000 lb/hr)	8	0
5	Raw Gas Knockout Drum	Vertical with mist eliminator	215,000 kg/hr, 35°C, 5.3 MPa (475,000 lb/hr, 95°F, 765 psia)	2	0
6	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	336,000 kg/hr (741,000 lb/hr) syngas	2	0
7	ASU Main Air Compressor	Centrifugal, multi-stage	5,000 m ³ /min @ 1.3 MPa (168,000 scfm @ 190 psia)	2	0
8	Cold Box	Vendor design	2,300 tonne/day (2,500 tpd) of 95% purity oxygen	2	0
9	Oxygen Compressor	Centrifugal, multi-stage	1,000 m ³ /min (41,000 scfm) Suction - 1.0 MPa (130 psia) Discharge - 6.5 MPa (940 psia)	2	0
10	Primary Nitrogen Compressor	Centrifugal, multi-stage	4,000 m ³ /min (133,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
11	Secondary Nitrogen Compressor	Centrifugal, single-stage	530 m ³ /min (19,000 scfm) Suction - 1.3 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0
12	AGR Nitrogen Boost Compressor	Centrifugal, single-stage	90 m ³ /min (3,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 5.4 MPa (790 psia)	2	0
13	Extraction Air Heat Exchanger	Gas-to-gas, vendor design	70,000 kg/hr, 432°C, 1.6 MPa (154,000 lb/hr, 810°F, 235 psia)	2	0

Case B5A – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	215,000 kg/hr (474,000 lb/hr), 35°C (95°F), 5.2 MPa (760 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	215,000 kg/hr (474,000 lb/hr), 35°C (95°F), 5.2 MPa (760 psia)	2	0
3	Sulfur Plant	Claus type	140 tonne/day (154 tpd)	1	0
4	COS Hydrolysis Reactor	Fixed bed, catalytic	304,000 kg/hr (670,000 lb/hr), 221°C (430°F), 5.4 MPa (790 psia)	2	0
5	Acid Gas Removal Plant	Selexol	232,000 kg/hr (512,000 lb/hr), 35°C (94°F), 5.2 MPa (755 psia)	2	0
6	Hydrogenation Reactor	Fixed bed, catalytic	40,000 kg/hr (88,000 lb/hr) 232°C (450°F) 0.4 MPa (58.9 psia)	1	0
7	Tail Gas Recycle Compressor	Centrifugal	35,000 kg/hr (78,000 lb/hr) each	1	0

Case B5A – Account 6: Combustion Turbine and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	State-of-the-art 2008 F-class	230 MW	2	0
2	Combustion Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0
3	Syngas Expansion Turbine/Generator	Turbo Expander	218,400 kg/h (481,400 lb/h) 5.1 MPa (745 psia) Inlet 3.2 MPa (460 psia) Outlet	2	0

Case B5A – Account 7: HRSG, Ducting, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.4 m (27 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 352,698 kg/hr, 12.4 MPa/561°C (777,565 lb/hr, 1,800 psig/1,043°F)	2	0

Case B5A – Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	291 MW 12.4 MPa/561°C/561°C (1,800 psig/ 1,043°F/1,043°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	320 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,690GJ/hr (1,600 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0

Case B5A – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	423,000 lpm @ 30 m (112,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 2,360 GJ/hr (2,240 MMBtu/hr) heat duty	1	0

Case B5A – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	244,000 liters (64,000 gal)	2	0
2	Slag Crusher	Roll	13 tonne/hr (14 tph)	2	0
3	Slag Depressurizer	Lock Hopper	13 tonne/hr (14 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	147,000 liters (39,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	66,000 liters (17,000 gal)	2	0
6	Slag Conveyor	Drag chain	13 tonne/hr (14 tph)	2	0
7	Slag Separation Screen	Vibrating	13 tonne/hr (14 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	13 tonne/hr (14 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	208,000 liters (55,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	50 lpm @ 14 m H ₂ O (10 gpm @ 46 ft H ₂ O)	2	2
11	Grey Water Storage Tank	Field erected	66,000 liters (18,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	240 lpm @ 560 m H ₂ O (60 gpm @ 1,850 ft H ₂ O)	2	2
13	Slag Storage Bin	Vertical, field erected	900 tonne (1,000 tons)	2	0
14	Unloading Equipment	Telescoping chute	110 tonne/hr (120 tph)	1	0

Case B5A – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	24 kV/345 kV, 290 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 54 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 29 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 4 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case B5A – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.4.7 Case B5A - Cost Estimating

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-88 shows a detailed breakdown of the capital costs; Exhibit 3-89 shows the owner's costs, TOC, and TASC; Exhibit 3-90 shows the initial and annual O&M costs; and Exhibit 3-91 shows the COE breakdown.

The estimated TPC of the GEE gasifier with no CO₂ capture is \$2,449/kW. Process contingency represents 2.5 percent of the TPC, and project contingency represents 13.5 percent. The COE is \$102.6/MWh.

Exhibit 3-88 Case B5A total plant cost details

Case:		B5A – GEE Radiant IGCC w/o CO ₂					Estimate Type:			Conceptual	
Plant Size (MW,net):		622					Cost Base:			Jun 2011	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
1		Coal & Sorbent Handling									
1.1	Coal Receive & Unload	\$4,435	\$0	\$2,137	\$0	\$6,572	\$657	\$0	\$1,446	\$8,675	\$14
1.2	Coal Stackout & Reclaim	\$5,731	\$0	\$1,370	\$0	\$7,101	\$710	\$0	\$1,562	\$9,373	\$15
1.3	Coal Conveyors & Yd Crush	\$5,328	\$0	\$1,356	\$0	\$6,684	\$668	\$0	\$1,470	\$8,823	\$14
1.4	Other Coal Handling	\$1,394	\$0	\$314	\$0	\$1,708	\$171	\$0	\$376	\$2,254	\$4
1.9	Coal & Sorbent Hnd. Foundations	\$0	\$2,961	\$7,739	\$0	\$10,701	\$1,070	\$0	\$2,354	\$14,125	\$23
Subtotal		\$16,887	\$2,961	\$12,917	\$0	\$32,765	\$3,277	\$0	\$7,208	\$43,250	\$70
2		Coal & Sorbent Prep & Feed									
2.1	Coal Crushing & Drying	w/ 2.3	\$0	w/ 2.3	\$0	\$0	\$0	\$0	\$0	\$0	\$0
2.2	Prepared Coal Storage & Feed	\$1,891	\$454	\$292	\$0	\$2,638	\$264	\$0	\$580	\$3,482	\$6
2.3	Slurry Prep & Feed	\$25,812	\$0	\$11,252	\$0	\$37,064	\$3,706	\$1,853	\$8,525	\$51,149	\$82
2.4	Misc. Coal Prep & Feed	\$1,040	\$760	\$2,238	\$0	\$4,037	\$404	\$0	\$888	\$5,329	\$9
2.9	Coal & Sorbent Feed Foundation	\$0	\$3,814	\$3,273	\$0	\$7,087	\$709	\$0	\$1,559	\$9,355	\$15
Subtotal		\$28,743	\$5,028	\$17,055	\$0	\$50,826	\$5,083	\$1,853	\$11,552	\$69,314	\$111
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	\$3,461	\$5,974	\$3,132	\$0	\$12,567	\$1,257	\$0	\$2,765	\$16,589	\$27
3.2	Water Makeup & Pretreating	\$758	\$78	\$416	\$0	\$1,252	\$125	\$0	\$413	\$1,791	\$3
3.3	Other Feedwater Subsystems	\$1,946	\$643	\$575	\$0	\$3,164	\$316	\$0	\$696	\$4,176	\$7
3.4	Service Water Systems	\$443	\$883	\$3,045	\$0	\$4,372	\$437	\$0	\$1,443	\$6,251	\$10
3.5	Other Boiler Plant Systems	\$2,388	\$892	\$2,196	\$0	\$5,476	\$548	\$0	\$1,205	\$7,228	\$12
3.6	FO Supply Sys & Nat Gas	\$17,740	\$718	\$665	\$0	\$19,123	\$1,912	\$0	\$4,207	\$25,242	\$41
3.7	Waste Treatment Equipment	\$1,025	\$0	\$635	\$0	\$1,660	\$166	\$0	\$548	\$2,374	\$4
3.8	Misc. Power Plant Equipment	\$1,277	\$170	\$664	\$0	\$2,111	\$211	\$0	\$697	\$3,019	\$5
Subtotal		\$29,037	\$9,360	\$11,328	\$0	\$49,725	\$4,972	\$0	\$11,973	\$66,670	\$107
4		Gasifier & Accessories									
4.1	Syngas Cooler Gasifier System	\$136,778	\$0	\$75,333	\$0	\$212,111	\$21,211	\$29,509	\$40,271	\$303,102	\$487
4.2	Syngas Cooler	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.3	ASU & Oxidant Compression	\$195,452	\$0	w/equip.	\$0	\$195,452	\$19,545	\$0	\$21,500	\$236,496	\$380
4.4	Scrubber & LT Heat Recovery	\$7,080	\$5,763	\$5,957	\$0	\$18,800	\$1,880	\$0	\$4,136	\$24,816	\$40
4.5	Black Water & Sour Gas Section	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.6	Soot Recovery & SARU	\$2,159	\$1,025	\$2,013	\$0	\$5,197	\$520	\$0	\$1,143	\$6,860	\$11
4.8	Major Component Rigging	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.9	Gasification Foundations	\$0	\$7,429	\$6,482	\$0	\$13,911	\$1,391	\$0	\$3,826	\$19,128	\$31
Subtotal		\$341,469	\$14,218	\$89,784	\$0	\$445,471	\$44,547	\$29,509	\$70,875	\$590,403	\$949
5A		Gas Cleanup & Piping									
5A.1	Single Stage Selexol	\$50,584	\$0	\$42,628	\$0	\$93,212	\$9,321	\$0	\$20,507	\$123,040	\$198
5A.2	Elemental Sulfur Plant	\$12,121	\$2,363	\$15,532	\$0	\$30,015	\$3,002	\$0	\$6,603	\$39,620	\$64
5A.3	Mercury Removal	\$1,515	\$0	\$1,145	\$0	\$2,661	\$266	\$133	\$612	\$3,672	\$6
5A.4	COS Hydrolysis	\$4,309	\$0	\$5,589	\$0	\$9,898	\$990	\$0	\$2,178	\$13,066	\$21
5A.7	Fuel Gas Piping	\$0	\$882	\$577	\$0	\$1,459	\$146	\$0	\$321	\$1,926	\$3

Case:		B5A – GEE Radiant IGCC w/o CO ₂				Estimate Type:		Conceptual			
Plant Size (MW,net):		622				Cost Base:		Jun 2011			
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
5A.9	HGCU Foundations	\$0	\$798	\$538	\$0	\$1,336	\$134	\$0	\$441	\$1,911	\$3
	Subtotal	\$68,530	\$4,043	\$66,009	\$0	\$138,582	\$13,858	\$133	\$30,662	\$183,234	\$295
6	Combustion Turbine & Accessories										
6.1	Combustion Turbine Generator	\$104,200	\$0	\$7,506	\$0	\$111,706	\$11,171	\$5,585	\$12,846	\$141,308	\$227
6.2	Syngas Expander	\$7,146	\$0	\$981	\$0	\$8,126	\$813	\$0	\$1,341	\$10,280	\$17
6.9	Combustion Turbine Foundations	\$0	\$1,016	\$1,175	\$0	\$2,191	\$219	\$0	\$723	\$3,133	\$5
	Subtotal	\$111,346	\$1,016	\$9,662	\$0	\$122,024	\$12,202	\$5,585	\$14,910	\$154,721	\$249
7	HRSG, Ducting, & Stack										
7.1	Heat Recovery Steam Generator	\$31,080	\$0	\$6,019	\$0	\$37,099	\$3,710	\$0	\$4,081	\$44,890	\$72
7.3	Ductwork	\$0	\$2,080	\$1,458	\$0	\$3,538	\$354	\$0	\$778	\$4,670	\$8
7.4	Stack	\$4,014	\$0	\$1,498	\$0	\$5,511	\$551	\$0	\$606	\$6,669	\$11
7.9	HRSG, Duct & Stack Foundations	\$0	\$764	\$767	\$0	\$1,531	\$153	\$0	\$505	\$2,190	\$4
	Subtotal	\$35,094	\$2,844	\$9,741	\$0	\$47,679	\$4,768	\$0	\$5,971	\$58,418	\$94
8	Steam Turbine Generator										
8.1	Steam TG & Accessories	\$38,955	\$0	\$5,979	\$0	\$44,934	\$4,493	\$0	\$4,943	\$54,370	\$87
8.2	Turbine Plant Auxiliaries	\$243	\$0	\$554	\$0	\$798	\$80	\$0	\$88	\$965	\$2
8.3	Condenser & Auxiliaries	\$3,300	\$0	\$1,878	\$0	\$5,178	\$518	\$0	\$570	\$6,265	\$10
8.4	Steam Piping	\$15,350	\$0	\$6,657	\$0	\$22,006	\$2,201	\$0	\$6,052	\$30,258	\$49
8.9	TG Foundations	\$0	\$1,147	\$2,027	\$0	\$3,174	\$317	\$0	\$1,047	\$4,539	\$7
	Subtotal	\$57,848	\$1,147	\$17,094	\$0	\$76,089	\$7,609	\$0	\$12,699	\$96,397	\$155
9	Cooling Water System										
9.1	Cooling Towers	\$4,180	\$0	\$1,270	\$0	\$5,450	\$545	\$0	\$899	\$6,894	\$11
9.2	Circulating Water Pumps	\$2,012	\$0	\$146	\$0	\$2,158	\$216	\$0	\$356	\$2,730	\$4
9.3	Circ. Water System Auxiliaries	\$177	\$0	\$25	\$0	\$202	\$20	\$0	\$33	\$256	\$0
9.4	Circ. Water Piping	\$0	\$7,847	\$1,901	\$0	\$9,747	\$975	\$0	\$2,144	\$12,867	\$21
9.5	Make-up Water System	\$428	\$0	\$589	\$0	\$1,017	\$102	\$0	\$224	\$1,342	\$2
9.6	Component Cooling Water Sys	\$897	\$1,073	\$737	\$0	\$2,707	\$271	\$0	\$595	\$3,573	\$6
9.9	Circ. Water System Foundations	\$0	\$2,576	\$4,576	\$0	\$7,152	\$715	\$0	\$2,360	\$10,227	\$16
	Subtotal	\$7,695	\$11,495	\$9,243	\$0	\$28,433	\$2,843	\$0	\$6,612	\$37,889	\$61
10	Ash & Spent Sorbent Handling Systems										
10.1	Slag Dewatering & Cooling	\$14,251	\$7,858	\$15,855	\$0	\$37,965	\$3,797	\$0	\$4,176	\$45,938	\$74
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.6	Ash Storage Silos	\$693	\$0	\$749	\$0	\$1,442	\$144	\$0	\$238	\$1,825	\$3
10.7	Ash Transport & Feed Equipment	\$954	\$0	\$223	\$0	\$1,177	\$118	\$0	\$194	\$1,489	\$2
10.8	Misc. Ash Handling Equipment	\$1,436	\$1,760	\$522	\$0	\$3,718	\$372	\$0	\$613	\$4,703	\$8
10.9	Ash/Spent Sorbent Foundation	\$0	\$58	\$76	\$0	\$135	\$13	\$0	\$44	\$193	\$0
	Subtotal	\$17,335	\$9,676	\$17,426	\$0	\$44,437	\$4,444	\$0	\$5,266	\$54,147	\$87
11	Accessory Electric Plant										
11.1	Generator Equipment	\$1,165	\$0	\$1,134	\$0	\$2,299	\$230	\$0	\$253	\$2,782	\$4
11.2	Station Service Equipment	\$4,601	\$0	\$423	\$0	\$5,024	\$502	\$0	\$553	\$6,080	\$10
11.3	Switchgear & Motor Control	\$8,493	\$0	\$1,579	\$0	\$10,071	\$1,007	\$0	\$1,662	\$12,740	\$20

Case:		B5A – GEE Radiant IGCC w/o CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		622					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
11.4	Conduit & Cable Tray	\$0	\$4,315	\$13,302	\$0	\$17,617	\$1,762	\$0	\$4,845	\$24,223	\$39
11.5	Wire & Cable	\$0	\$8,320	\$5,062	\$0	\$13,382	\$1,338	\$0	\$3,680	\$18,400	\$30
11.6	Protective Equipment	\$0	\$805	\$2,988	\$0	\$3,793	\$379	\$0	\$626	\$4,798	\$8
11.7	Standby Equipment	\$276	\$0	\$275	\$0	\$552	\$55	\$0	\$91	\$698	\$1
11.8	Main Power Transformers	\$18,037	\$0	\$175	\$0	\$18,211	\$1,821	\$0	\$3,005	\$23,037	\$37
11.9	Electrical Foundations	\$0	\$183	\$498	\$0	\$680	\$68	\$0	\$225	\$973	\$2
Subtotal		\$32,572	\$13,623	\$25,436	\$0	\$71,630	\$7,163	\$0	\$14,938	\$93,732	\$151
12		Instrumentation & Control									
12.1	IGCC Control Equipment	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.2	Combustion Turbine Control	w/6.1	\$0	w/6.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.3	Steam Turbine Control	w/8.1	\$0	w/8.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.4	Other Major Component Control	\$1,187	\$0	\$809	\$0	\$1,996	\$200	\$100	\$344	\$2,640	\$4
12.5	Signal Processing Equipment	w/12.7	\$0	w/12.7	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.6	Control Boards, Panels & Racks	\$273	\$0	\$178	\$0	\$451	\$45	\$23	\$104	\$623	\$1
12.7	Computer & Accessories	\$6,332	\$0	\$207	\$0	\$6,539	\$654	\$327	\$752	\$8,272	\$13
12.8	Instrument Wiring & Tubing	\$0	\$2,438	\$4,614	\$0	\$7,052	\$705	\$353	\$2,027	\$10,137	\$16
12.9	Other I & C Equipment	\$4,233	\$0	\$2,097	\$0	\$6,330	\$633	\$317	\$1,092	\$8,372	\$13
Subtotal		\$12,025	\$2,438	\$7,906	\$0	\$22,368	\$2,237	\$1,118	\$4,319	\$30,043	\$48
13		Improvements to Site									
13.1	Site Preparation	\$0	\$118	\$2,689	\$0	\$2,807	\$281	\$0	\$926	\$4,014	\$6
13.2	Site Improvements	\$0	\$2,103	\$2,973	\$0	\$5,077	\$508	\$0	\$1,675	\$7,260	\$12
13.3	Site Facilities	\$3,769	\$0	\$4,231	\$0	\$8,000	\$800	\$0	\$2,640	\$11,440	\$18
Subtotal		\$3,769	\$2,222	\$9,893	\$0	\$15,884	\$1,588	\$0	\$5,242	\$22,714	\$37
14		Buildings & Structures									
14.1	Combustion Turbine Area	\$0	\$317	\$179	\$0	\$496	\$50	\$0	\$109	\$655	\$1
14.2	Steam Turbine Building	\$0	\$2,930	\$4,172	\$0	\$7,102	\$710	\$0	\$1,172	\$8,984	\$14
14.3	Administration Building	\$0	\$1,013	\$734	\$0	\$1,747	\$175	\$0	\$288	\$2,210	\$4
14.4	Circulation Water Pumphouse	\$0	\$200	\$106	\$0	\$305	\$31	\$0	\$50	\$386	\$1
14.5	Water Treatment Buildings	\$0	\$623	\$607	\$0	\$1,230	\$123	\$0	\$203	\$1,556	\$3
14.6	Machine Shop	\$0	\$518	\$354	\$0	\$873	\$87	\$0	\$144	\$1,104	\$2
14.7	Warehouse	\$0	\$837	\$540	\$0	\$1,377	\$138	\$0	\$227	\$1,742	\$3
14.8	Other Buildings & Structures	\$0	\$501	\$390	\$0	\$891	\$89	\$0	\$196	\$1,177	\$2
14.9	Waste Treating Building & Str.	\$0	\$1,121	\$2,140	\$0	\$3,261	\$326	\$0	\$717	\$4,304	\$7
Subtotal		\$0	\$8,059	\$9,222	\$0	\$17,282	\$1,728	\$0	\$3,107	\$22,117	\$36
Total		\$762,350	\$88,130	\$312,716	\$0	\$1,163,195	\$116,320	\$38,199	\$205,336	\$1,523,050	\$2,449

Exhibit 3-89 Case B5A owner's costs

Description	\$/1,000	\$/kW
Pre-Production Costs		
6 Months All Labor	\$14,667	\$24
1 Month Maintenance Materials	\$3,346	\$5
1 Month Non-fuel Consumables	\$452	\$1
1 Month Waste Disposal	\$473	\$1
25% of 1 Months Fuel Cost at 100% CF	\$2,920	\$5
2% of TPC	\$30,461	\$49
Total	\$52,318	\$84
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	\$23,916	\$38
0.5% of TPC (spare parts)	\$7,615	\$12
Total	\$31,532	\$51
Other Costs		
Initial Cost for Catalyst and Chemicals	\$11,013	\$18
Land	\$900	\$1
Other Owner's Costs	\$228,457	\$367
Financing Costs	\$41,122	\$66
Total Overnight Costs (TOC)	\$1,888,393	\$3,036
TASC Multiplier (IOU, high-risk, 35 year)	1.140	
Total As-Spent Cost (TASC)	\$2,152,768	\$3,461

Exhibit 3-90 Case B5A initial and annual operating and maintenance costs

Case:	B5A – GEE Radiant IGCC w/o CO ₂			Cost Base:	Jun 2011
Plant Size (MW _{net}):	622	Heat Rate-net (Btu/kWh):	8,758	Capacity Factor (%):	80
Operating & Maintenance Labor					
Operating Labor			Operating Labor Requirements per Shift		
Operating Labor Rate (base):	39.70	\$/hour	Skilled Operator:	2.0	
Operating Labor Burden:	30.00	% of base	Operator:	9.0	
Labor O-H Charge Rate:	25.00	% of labor	Foreman:	1.0	
			Lab Tech's, etc.:	3.0	
			Total:	15.0	
Fixed Operating Costs					
			Annual Cost		
			(\$)	(\$/kW-net)	
Annual Operating Labor:				\$6,781,554	\$10.904
Maintenance Labor:				\$16,685,386	\$26.828
Administrative & Support Labor:				\$5,866,735	\$9.433
Property Taxes and Insurance:				\$30,460,991	\$48.977
Total:				\$59,794,666	\$96.142
Variable Operating Costs					
			(\$)	(\$/MWh-net)	
Maintenance Material:				\$32,119,526	\$7.36931
Consumables					
	Consumption			Cost (\$)	
	Initial Fill	Per Day	Per Unit	Initial Fill	
Water (/1000 gallons):	0	3,423	\$1.67	\$0	\$1,673,317
Makeup and Waste Water Treatment Chemicals (lbs):	0	20,395	\$0.27	\$0	\$1,595,062
Carbon (Mercury Removal) (lb):	92,086	158	\$5.50	\$506,474	\$253,237
COS Catalyst (m³):	422	0.29	\$3,751.70	\$1,582,971	\$316,594
Selexol Solution (gal):	242,554	36	\$36.79	\$8,923,873	\$386,544
Claus Catalyst (ft³):	w/equip.	1.94	\$203.15	\$0	\$115,175
Subtotal:				\$11,013,317	\$4,339,929
Waste Disposal					
Spent Mercury Catalyst (lb.):	0	158	\$0.65	\$0	\$29,932
Flyash (ton):	0	0	\$0.00	\$0	\$0
Slag (ton):	0	615	\$25.11	\$0	\$4,507,185
Subtotal:				\$0	\$4,537,118
By-Products					
Sulfur (tons):	0	140	\$0.00	\$0	\$0
Subtotal:				\$0	\$0
Variable Operating Costs Total:				\$11,013,317	\$40,996,573
Fuel Cost					
Illinois Number 6 (ton):	0	5,603	\$68.54	\$0	\$112,132,176
Total:				\$0	\$112,132,176

Exhibit 3-91 Case B5A COE breakdown

Component	Value, \$/MWh	Percentage
Capital	53.7	52%
Fixed	13.7	13%
Variable	9.4	9%
Fuel	25.7	25%
Total (Excluding T&S)	102.6	N/A
CO ₂ T&S	0.0	0%
Total (Including T&S)	102.6	N/A

3.4.8 Case B5B - GEE Radiant IGCC with CO₂ Capture

Case B5B is configured to produce electric power with CO₂ capture. The plant configuration is the same as Case B5A, namely two gasifier trains, two state-of-the-art 2008 F-class turbines, two HRSGs, and one steam turbine. The gross power output from the plant is constrained by the capacity of the two CTs, and since the CO₂ capture process increases the auxiliary load on the plant, the net output is significantly reduced relative to Case B5A.

The process description for Case B5B is similar to Case B5A with several notable exceptions to accommodate CO₂ capture. A BFD and stream tables for Case B5B are shown in Exhibit 3-92 and Exhibit 3-93, respectively. Instead of repeating the entire process description, only differences from Case B5A are reported here.

3.4.8.1 Gasification

The gasification process is the same as Case B5A with the exception that total coal feed to the two gasifiers is 5,302 tonnes/day (5,844 tpd) (stream 6) and the ASU provides 4,342 tonnes/day (4,786 tpd) of 95 percent oxygen to the gasifier and Claus plant (streams 3 and 5).

3.4.8.2 Raw Gas Cooling/Particulate Removal

Raw gas cooling and particulate removal are the same as Case B5A with the exception that approximately 443,118 kg/hr (976,891 lb/hr) of saturated steam at 13.8 MPa (2,000 psia) is generated in the radiant SGCs.

3.4.8.3 Syngas Scrubber/Sour Water Stripper

No differences from Case B5A.

3.4.8.4 Sour Gas Shift

The SGS process was described in Section 3.1.3. In Case B5B steam (stream 11) is added to the syngas exiting the scrubber to adjust the H₂O:CO molar ratio to 2:1 prior to the first SGS reactor. The hot syngas exiting the first stage of SGS is used to generate the steam that is added in stream 11. A second stage of SGS results in 97 percent overall conversion of the CO to CO₂. The warm syngas from the second stage of SGS (stream 12) is cooled to 236°C (456°F) by preheating the unshifted syngas prior to the SGS. The SGS catalyst also serves to hydrolyze COS, thus eliminating the need for a separate COS hydrolysis reactor. Following the second SGS cooler, the syngas is further cooled to 35°C (95°F) prior to the mercury removal beds.

3.4.8.5 Mercury Removal and AGR

Mercury removal is the same as in Case B5A.

The AGR process in Case B5B is a two-stage Selexol process where H₂S is removed in the first stage and CO₂ in the second stage of absorption, as previously described in Section 3.1.5. The process results in three product streams: the clean syngas, a CO₂-rich stream and an acid gas feed to the Claus plant. The acid gas (stream 18) contains 35 percent H₂S and 52 percent CO₂ with the balance primarily N₂. The CO₂-rich stream is discussed further in the CO₂ compression section.

Exhibit 3-93 Case B5B stream table, GEE IGCC with capture

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0166	0.0318	0.0023	0.0318	0.0000	0.0000	0.0000	0.0086	0.0068	0.0000	0.0054	0.0071	0.0071
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0011	0.0009	0.0000	0.0007	0.0009	0.0009
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3576	0.2823	0.0000	0.0060	0.0078	0.0077
CO ₂	0.0003	0.0054	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1380	0.1089	0.0000	0.3082	0.4019	0.4055
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0001	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	0.3406	0.2689	0.0000	0.4366	0.5692	0.5649
H ₂ O	0.0099	0.1363	0.0000	0.0003	0.0000	0.0000	0.9995	0.0000	0.1369	0.3190	1.0000	0.2325	0.0012	0.0013
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0002	0.0000	0.0001	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0057	0.0000	0.0047	0.0061	0.0061
N ₂	0.7732	0.7061	0.0178	0.9920	0.0178	0.0000	0.0000	0.0000	0.0070	0.0055	0.0000	0.0044	0.0058	0.0064
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0000	0.0019	0.0016	0.0000	0.0013	0.0000	0.0000
O ₂	0.2074	0.1356	0.9504	0.0054	0.9504	0.0000	0.0000	0.0000	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	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	27,361	1,650	96	20,050	5,526	0	5,037	0	23,122	29,284	7,193	36,477	27,978	28,367
V-L Flowrate (kg/hr)	789,549	45,326	3,080	562,613	177,826	0	90,747	0	465,238	575,977	129,586	705,563	552,385	564,913
Solids Flowrate (kg/hr)	0	0	0	0	0	220,902	0	24,237	0	0	0	0	0	0
Temperature (°C)	15	18	32	93	32	15	142	1,316	677	206	288	240	35	35
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	0.10	5.79	5.62	5.55	5.52	5.52	5.41	5.14	5.10
Steam Table Enthalpy (kJ/kg) ^A	30.23	35.64	26.67	92.50	26.67	---	537.77	---	1,428.30	1,066.52	2,917.15	942.88	37.11	36.40
AspenPlus Enthalpy (kJ/kg) ^B	-97.58	-1,492.10	4.13	63.47	4.13	-2,116.95	-15,427.03	-727.24	-5,246.11	-7,406.90	-13,053.16	-8,863.69	-8,084.69	-8,087.64
Density (kg/m ³)	1.2	1.5	11.0	24.4	11.0	---	872.0	---	14.0	27.2	25.6	24.8	40.7	40.9
V-L Molecular Weight	28.857	27.476	32.181	28.060	32.181	---	18.015	---	20.121	19.669	18.015	19.343	19.744	19.914
V-L Flowrate (lb _{mole} /hr)	60,320	3,637	211	44,204	12,182	0	11,105	0	50,975	64,560	15,858	80,418	61,680	62,539
V-L Flowrate (lb/hr)	1,740,658	99,926	6,790	1,240,349	392,039	0	200,062	0	1,025,674	1,269,812	285,687	1,555,499	1,217,800	1,245,421
Solids Flowrate (lb/hr)	0	0	0	0	0	487,005	0	53,433	0	0	0	0	0	0
Temperature (°F)	59	65	90	199	90	59	287	2,400	1,250	403	550	463	95	95
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	14.7	840.0	815.0	805.0	800.0	800.0	785.0	745.0	740.0
Steam Table Enthalpy (Btu/lb) ^A	13.0	15.3	11.5	39.8	11.5	---	231.2	---	614.1	458.5	1,254.1	405.4	16.0	15.7
AspenPlus Enthalpy (Btu/lb) ^B	-42.0	-641.5	1.8	27.3	1.8	-910.1	-6,632.4	-312.7	-2,255.4	-3,184.4	-5,611.9	-3,810.7	-3,475.8	-3,477.1
Density (lb/ft ³)	0.076	0.091	0.687	1.521	0.687	---	54.440	---	0.871	1.699	1.597	1.550	2.544	2.552

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 3-93 Case B5B stream table, GEE IGCC with capture (continued)

	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction											
Ar	0.0115	0.0115	0.0115	0.0002	0.0018	0.0000	0.0103	0.0092	0.0091	0.0091	0.0000
CH ₄	0.0015	0.0015	0.0015	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0124	0.0124	0.0124	0.0002	0.0022	0.0000	0.0064	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0502	0.0502	0.0502	0.9948	0.5214	0.0000	0.6664	0.0003	0.0083	0.0083	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.9139	0.9139	0.9139	0.0048	0.1028	0.0000	0.2562	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0001	0.0001	0.0001	0.0000	0.0226	0.0000	0.0017	0.0099	0.1222	0.1222	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.3477	0.0000	0.0050	0.0000	0.0000	0.0000	0.0000
N ₂	0.0105	0.0105	0.0105	0.0000	0.0008	0.0000	0.0541	0.7732	0.7541	0.7541	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
O ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.1064	0.1064	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	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	17,423	17,423	17,423	10,425	497	0	390	110,253	139,657	139,657	34,514
V-L Flowrate (kg/hr)	90,178	90,178	90,178	456,650	17,683	0	12,529	3,181,557	3,834,349	3,834,349	621,772
Solids Flowrate (kg/hr)	0	0	0	0	0	5,524	0	0	0	0	0
Temperature (°C)	35	241	196	51	48	178	38	15	562	132	534
Pressure (MPa, abs)	5.10	5.07	3.17	15.27	0.16	0.12	5.51	0.10	0.10	0.10	12.51
Steam Table Enthalpy (kJ/kg) ^A	195.53	1,389.05	1,124.24	-162.31	74.87	---	5.30	30.23	834.76	343.83	3,432.88
AspenPlus Enthalpy (kJ/kg) ^B	-4,051.34	-2,857.82	-3,122.63	-9,120.19	-6,107.59	141.99	-8,217.57	-97.58	-587.80	-1,078.73	-12,537.44
Density (kg/m ³)	10.1	6.0	4.2	641.8	2.2	5,280.5	77.9	1.2	0.4	0.9	36.7
V-L Molecular Weight	5.176	5.176	5.176	43.805	35.588	---	32.150	28.857	27.456	27.456	18.015
V-L Flowrate (lb _{mole} /hr)	38,412	38,412	38,412	22,983	1,095	0	859	243,066	307,891	307,891	76,089
V-L Flowrate (lb/hr)	198,808	198,808	198,808	1,006,740	38,985	0	27,621	7,014,133	8,453,292	8,453,292	1,370,772
Solids Flowrate (lb/hr)	0	0	0	0	0	12,178	0	0	0	0	0
Temperature (°F)	95	465	384	124	119	352	100	59	1,044	270	994
Pressure (psia)	740.0	735.0	460.0	2,214.7	23.7	17.3	799.5	14.7	15.2	15.2	1,814.7
Steam Table Enthalpy (Btu/lb) ^A	84.1	597.2	483.3	-69.8	32.2	---	2.3	13.0	358.9	147.8	1,475.9
AspenPlus Enthalpy (Btu/lb) ^B	-1,741.8	-1,228.6	-1,342.5	-3,921.0	-2,625.8	61.0	-3,532.9	-42.0	-252.7	-463.8	-5,390.1
Density (lb/ft ³)	0.630	0.377	0.260	40.068	0.137	329.649	4.864	0.076	0.026	0.053	2.293

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

3.4.8.6 CO₂ Compression and Dehydration

CO₂ from the AGR process is flashed at three pressure levels to separate CO₂ and decrease H₂ losses to the CO₂ product pipeline. The HP CO₂ stream is flashed at 2.0 MPa (289.7 psia), compressed, and recycled back to the CO₂ absorber. The MP CO₂ stream is flashed at 1.0 MPa (149.7 psia). The LP CO₂ stream is flashed at 0.1 MPa (16.7 psia), compressed to 1.0 MPa (149.5 psia), and combined with the MP CO₂ stream. The combined stream is compressed from 2.1 MPa (149.5 psia) to an SC condition at 15.3 MPa (2,215 psia) using a multiple-stage, intercooled compressor. During compression, the CO₂ stream is dehydrated to a dew point of -40°C (-40°F) with triethylene glycol. The raw CO₂ stream from the Selexol process contains over 99 percent CO₂. The CO₂ (stream 17) is transported to the plant fence line and is sequestration ready. CO₂ T&S costs were estimated using the methodology described in Section 2.6.

3.4.8.7 Claus Unit

The Claus plant is the same as Case B5A with the following exceptions:

- 5,528 kg/hr (12,178 lb/hr) of sulfur (stream 19) are produced
- The waste heat boiler generates 13,555 kg/hr (29,884 lb/hr) of 4.0 MPa (575 psia) steam of which 12,679 kg/hr (27,953 lb/hr) is available to the medium pressure steam header

3.4.8.8 Power Block

Clean syngas from the AGR plant is heated to 241°C (465°F) using HP BFW before passing through an expansion turbine. The clean syngas (stream 16) is diluted with nitrogen (stream 4) and then enters the CT burner. There is no integration between the CT and the ASU in this case. The exhaust gas (stream 22) exits the CT at 562°C (1,044°F) and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 132°C (270°F) (stream 23) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available steam turbine using a 12.4 MPa/534°C/534°C (1,800 psig/994°F/994°F) steam cycle.

3.4.8.9 Air Separation Unit (ASU)

The same elevated pressure ASU is used in Case B5B and produces 4,342 tonnes/day (4,786 tpd) of 95 mol% oxygen and 14,591 tonnes/day (16,084 tpd) of nitrogen. There is no integration between the ASU and the CT in the CO₂ capture case.

3.4.9 Case B5B - Performance Results

The Case B5B modeling assumptions were presented previously in Section 3.4.3.

The plant produces a net output of 543 MW at a net plant efficiency of 32.6 percent (HHV basis). Overall performance for the entire plant is summarized in Exhibit 3-94; Exhibit 3-95 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for nearly 60 percent of the auxiliary load between the main air compressor, nitrogen compressor, oxygen compressor, and ASU auxiliaries. The two-stage Selexol process and CO₂ compression account for an additional 26 percent of the auxiliary power load. The BFW pumps and cooling water system (CWPs and cooling tower fan) compose over 6 percent of the load, leaving 8 percent of the auxiliary load for all other systems.

Exhibit 3-94 Case B5B plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	7
Steam Turbine Power, MWe	264
Total Gross Power, MWe	734
Air Separation Unit Main Air Compressor, kWe	67,330
Oxygen Compressor, kWe	10,640
Nitrogen Compressors, kWe	35,640
CO ₂ Compression, kWe	31,160
Acid Gas Removal, kWe	19,230
Balance of Plant, kWe	26,870
Total Auxiliaries, MWe	191
Net Power, MWe	543
HHV Net Plant Efficiency (%)	32.6%
HHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	11,034 (10,459)
HHV Cold Gas Efficiency, %	79.2%
HHV Combustion Turbine Efficiency, %	36.0%
LHV Net Plant Efficiency (%)	33.8%
LHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	10,643 (10,087)
LHV Cold Gas Efficiency, %	75.9%
LHV Combustion Turbine Efficiency, %	42.4%
Steam Turbine Cycle Efficiency, %	42.4%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	8,496 (8,053)
Condenser Duty, GJ/hr (MMBtu/hr)	1,513 (1,434)
As-Received Coal Feed, kg/hr (lb/hr)	220,902 (487,005)
HHV Thermal Input, kWt	1,665,056
LHV Thermal Input, kWt	1,605,967
Raw Water Withdrawal, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.041 (10.7)
Raw Water Consumption, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.033 (8.8)
O ₂ :Coal	0.761

Exhibit 3-95 Case B5B plant power summary

Power Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	7
Steam Turbine Power, MWe	264
Total Gross Power, MWe	734
Auxiliary Load Summary	
Coal Handling, kWe	470
Coal Milling, kWe	2,270
Sour Water Recycle Slurry Pump, kWe	190
Slag Handling, kWe	1,160
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	67,330
Oxygen Compressor, kWe	10,640
Nitrogen Compressors, kWe	35,640
Feedwater Pumps, kWe	4,180
Quench Water Pump, kWe	540
Syngas Recycle Compressor, kWe	0
CO ₂ Compression, kWe	31,160
Scrubber Pumps, kWe	230
Acid Gas Removal, kWe	19,230
Combustion Turbine Auxiliaries, kWe	1,000
Steam Turbine Auxiliaries, kWe	100
Condensate Pumps, kWe	280
Circulating Water Pumps, kWe	4,630
Ground Water Pumps, kWe	530
Cooling Tower Fans, kWe	2,400
Claus Plant/TGTU Auxiliaries, kWe	250
Claus Plant TG Recycle Compressor, kWe	1,780
Miscellaneous Balance of Plant ^{A,B} , kWe	3,000
Transformer Losses, kWe	2,860
Total Auxiliaries, MWe	191
Net Power, MWe	543

^AIncludes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.4.9.1 Environmental Performance

The environmental targets for emissions of Hg, NO_x, SO₂, and PM were presented in Section 2.4. A summary of the plant air emissions for Case B5B is presented in Exhibit 3-96. All HCl is assumed to be removed and is, therefore, not reported.

Exhibit 3-96 Case B5B air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year) ^A	kg/MWh (lb/MWh) ^B
SO ₂	0.001 (0.002)	39 (43)	0.008 (0.017)
NO _x	0.021 (0.049)	878 (967)	0.171 (0.376)
Particulates	0.003 (0.0071)	128 (141)	0.025 (0.055)
Hg	1.67E-7 (3.88E-7)	0.007 (0.008)	1.36E-6 (3.00E-6)
CO ₂	8 (20)	355,435 (391,800)	69 (152)
CO ₂ ^C	-	-	93 (206)

^ACalculations based on an 80 percent capacity factor

^BEmissions are based on gross power except where otherwise noted

^CCO₂ emissions based on net power instead of gross power

The low level of SO₂ emissions is achieved by capturing the sulfur in the gas by the two-stage Selexol AGR process. As a result of achieving the 90 percent CO₂ removal target, the sulfur compounds are removed to an extent that exceeds the environmental target in Section 2.3. The clean syngas exiting the AGR process has a sulfur concentration of approximately 5 ppmv. This results in a concentration in the flue gas of less than 1 ppmv. The H₂S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is hydrogenated to convert all sulfur species to H₂S and then recycled back to the Selexol process, thereby eliminating the need for a tail gas treatment unit.

NO_x emissions are limited by nitrogen dilution to 15 ppmvd (as NO₂ @15 percent O₂). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process. This helps lower NO_x levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas quench in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of the mercury is captured from the syngas by dual activated carbon beds.

Ninety two percent of the CO₂ from the syngas is captured in the AGR system and compressed for sequestration. Because not all CO is converted to CO₂ in the shift reactors, the overall carbon removal is 90.0 percent.

The carbon balance for the plant is shown in Exhibit 3-97. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon leaves the plant as unburned carbon in the slag and as CO₂ in the stack gas, ASU vent gas, and the captured CO₂ product. The carbon capture efficiency is defined as one minus the amount of carbon in the stack gas relative to the total carbon in less carbon contained in the slag, represented by the following fraction:

$$\frac{\text{Carbon in Stack}}{(\text{Total Carbon In}) - (\text{Carbon in Slag})} = \left(1 - \left(\frac{30,516}{311,631 - 6,209}\right) * \right) 100 = 90.0\%$$

Exhibit 3-97 Case B5B carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	140,813 (310,440)	Stack Gas	13,842 (30,516)
Air (CO ₂)	540 (1,191)	CO ₂ Product	124,589 (274,672)
		ASU Vent	107 (237)
		Slag	2,816 (6,209)
Total	141,353 (311,631)	Total	141,353 (311,631)

Exhibit 3-98 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant and sulfur emitted in the stack gas. Sulfur in the slag is considered to be negligible.

Exhibit 3-98 Case B5B sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,537 (12,206)	Stack Gas	3 (6)
		CO ₂ Product	10 (23)
		Elemental Sulfur	5,524 (12,178)
Total	5,537 (12,206)	Total	5,537 (12,206)

Exhibit 3-99 shows the overall water balance for the plant. The exhibit is presented in an identical manner as for cases B1A-B5A.

Exhibit 3-99 Case B5B water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)
Slag Handling	0.53 (139)	0.53 (139)	–	–	–
Slurry Water	1.51 (400)	1.51 (400)	–	–	–
Quench/Wash	2.86 (757)	0.72 (191)	2.14 (566)	–	2.14 (566)
Humidifier	–	–	–	–	–
SWS Blowdown	–	–	–	0.03 (7.30)	-0.03 (-7.30)
Condenser Makeup	2.37 (627)	–	2.37 (627)	–	2.37 (627)
BFW Makeup	0.21 (56)	–	0.21 (56)	–	0.21 (56)
Gasifier Steam	–	–	–	–	–
Shift Steam	2.16 (571)	–	2.16 (571)	–	2.16 (571)
CT Steam Dilution	–	–	–	–	–
Cooling Tower	18.06 (4,770)	0.49 (129)	17.57 (4,641)	4.06 (1,073)	13.51 (3,568)
BFW Blowdown	–	0.21 (56)	-0.21 (-56)	–	-0.21 (-56)
SWS Blowdown	–	0.28 (72.96)	-0.28 (-72.96)	–	-0.28 (-72.96)
Total	25.33 (6,692)	3.25 (858)	22.08 (5,834)	4.09 (1,080)	18.00 (4,754)

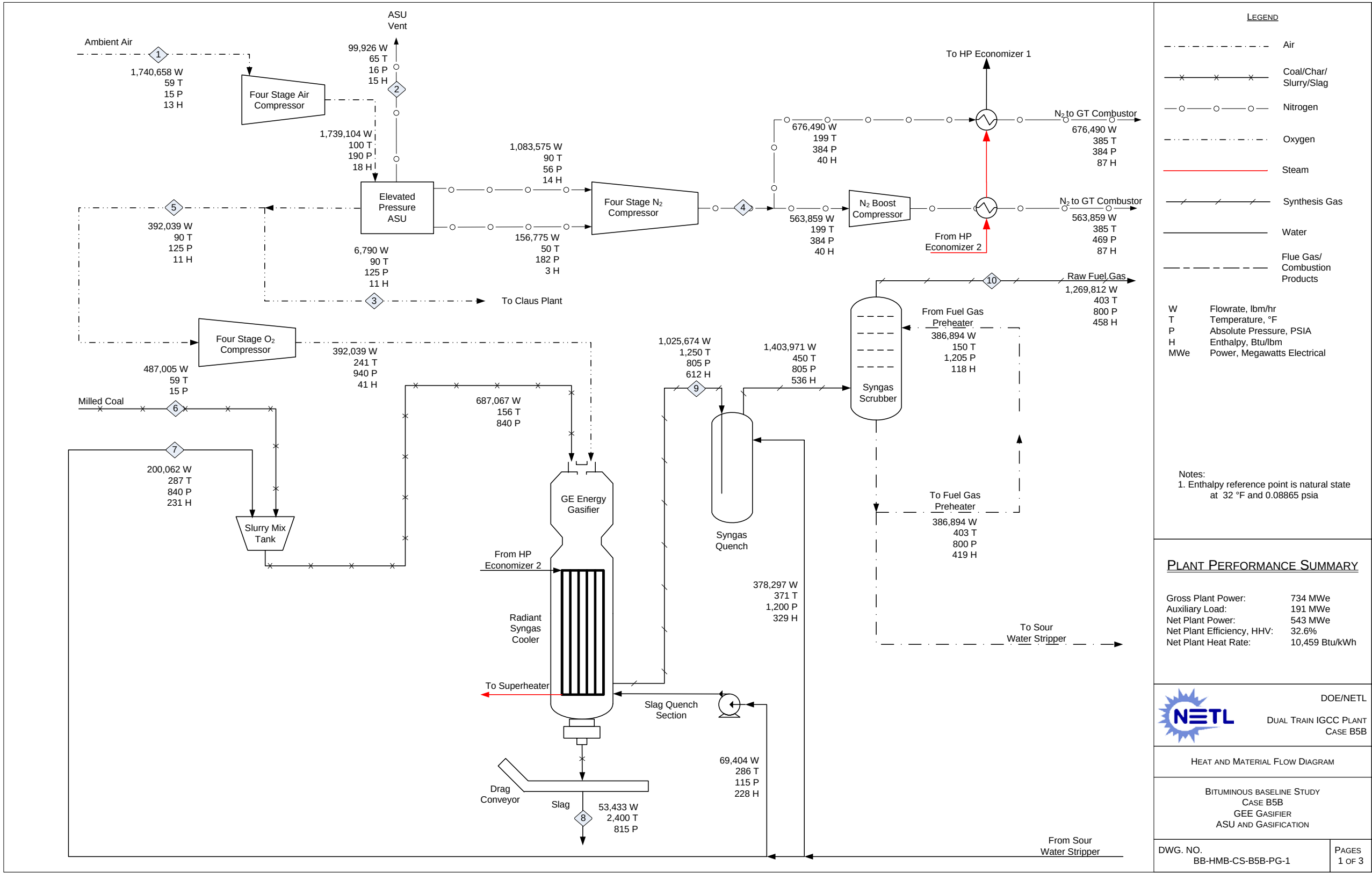
3.4.9.2 Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-100 through Exhibit 3-102:

- Coal gasification and ASU
- Syngas cleanup, sulfur recovery, and tail gas recycle
- Combined cycle power generation, steam, and FW

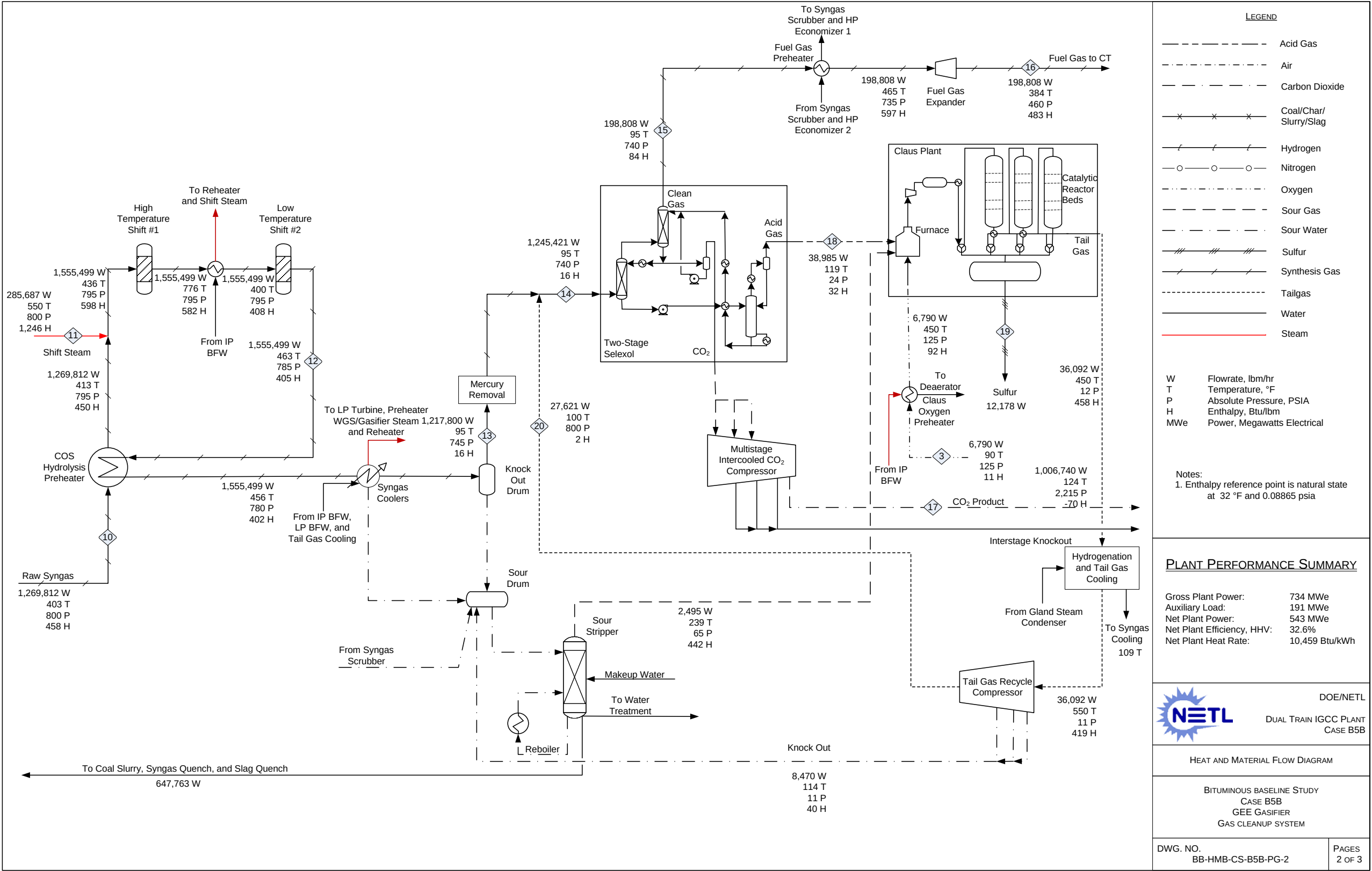
An overall plant energy balance is presented in tabular form in Exhibit 3-103. The power out is the combined CT, steam turbine, and expander power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-94) is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

Exhibit 3-100 Case B5B coal gasification and ASU heat and mass balance schematic



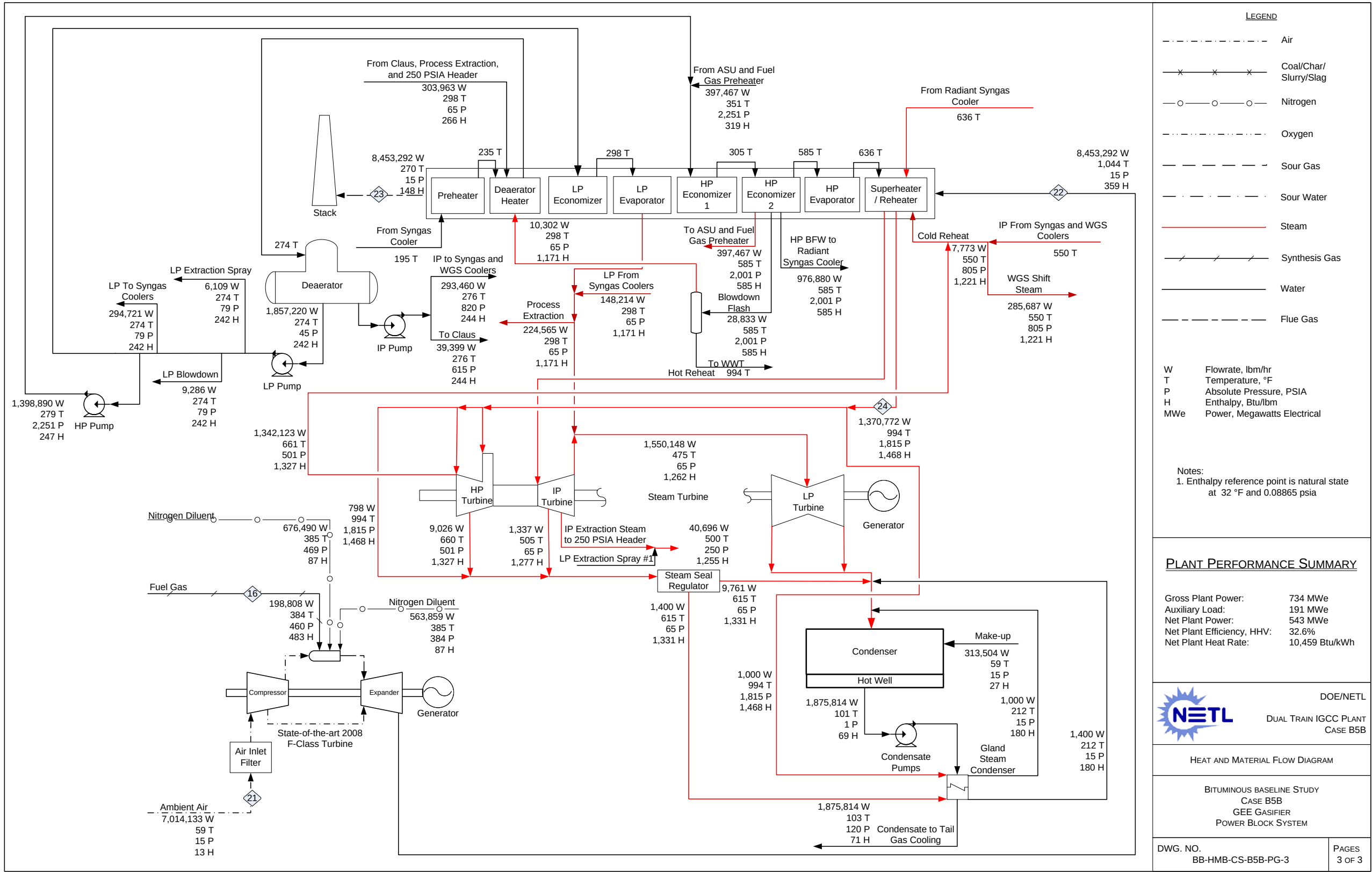
Source: NETL

Exhibit 3-101 Case B5B syngas cleanup heat and mass balance schematic



Source: NETL

Exhibit 3-102 Case B5B combined-cycle power generation heat and mass balance schematic



Source: NETL

This page intentionally left blank

Exhibit 3-103 Case B5B overall energy balance (0°C [32°F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In (GJ/hr)				
Coal	5,994 (5,681)	5.0 (4.7)	–	5,999 (5,686)
Air	–	120.0 (113.8)	–	120.0 (113.8)
Raw Water Makeup	–	83.0 (78.7)	–	83.0 (78.7)
Auxiliary Power	–	–	687 (651)	687 (651)
TOTAL	5,994 (5,681)	208.1 (197.2)	687 (651)	6,889 (6,530)
Heat Out (GJ/hr)				
ASU Vent	–	1.6 (1.5)	–	1.6 (1.5)
Slag	92.3 (87.5)	37.8 (35.9)	–	130.2 (123.4)
Stack Gas	–	1,318 (1,250)	–	1,318 (1,250)
Sulfur	51.2 (48.5)	0.6 (0.6)	–	51.8 (49.1)
Motor Losses and Design Allowances	–	–	60.1 (57.0)	60.1 (57.0)
Condenser	–	1,513 (1,434)	–	1,513 (1,434)
Non-Condenser Cooling Tower Loads	–	668 (633)	–	668 (633)
CO ₂	–	-74.1 (-70.2)	–	-74.1 (-70.2)
Blowdown	–	43.3 (41.0)	–	43.3 (41.0)
CO ₂ Capture Losses	–	176.4 (167.2)	–	176.4 (167.2)
<i>Ambient Losses^A</i>	–	<i>162.4 (154.0)</i>	–	<i>162.4 (154.0)</i>
Power	–	–	2,643 (2,505)	2,643 (2,505)
TOTAL	143.5 (136.0)	3,848 (3,647)	2,703 (2,562)	6,694 (6,345)
Unaccounted Energy ^B	–	195 (185)	–	195 (185)

^AAmbient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers.

3.4.10 Case B5B - Major Equipment List

Major equipment items for the GEE gasifier with CO₂ capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 3.4.11. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

Case B5B – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	50 tonne (50 ton)	2	1
7	Feeder	Vibratory	180 tonne/hr (200 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	180 tonne (200 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	800 tonne (900 ton)	3	0

Case B5B – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/h (90 tph)	3	0
2	Conveyor No. 6	Belt w/ tripper	240 tonne/h (270 tph)	1	0
3	Rod Mill Feed Hopper	Dual Outlet	490 tonne (540 ton)	1	0
4	Weigh Feeder	Belt	120 tonne/h (130 tph)	2	0
5	Rod Mill	Rotary	120 tonne/h (130 tph)	2	0
6	Slurry Water Storage Tank with Agitator	Field erected	299,900 liters (79,220 gal)	2	0
7	Slurry Water Pumps	Centrifugal	830 lpm (220 gpm)	2	1
8	Trommel Screen	Coarse	170 tonne/h (190 tph)	2	0
9	Rod Mill Discharge Tank with Agitator	Field erected	392,320 liters (103,640 gal)	2	0
10	Rod Mill Product Pumps	Centrifugal	3,300 lpm (900 gpm)	2	2
11	Slurry Storage Tank with Agitator	Field erected	1,176,900 liters (310,900 gal)	2	0
12	Slurry Recycle Pumps	Centrifugal	6,500 lpm (1,700 gpm)	2	2
13	Slurry Product Pumps	Positive displacement	3,300 lpm (900 gpm)	2	2

Case B5B – Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1,094,000 liters (289,000 gal)	2	0
2	Condensate Pumps	Vertical canned	7,850 lpm @ 90 m H ₂ O (2,070 gpm @ 300 ft H ₂ O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	546,000 kg/hr (1,205,000 lb/hr)	2	0
4	Intermediate-Pressure Feedwater Pump	Horizontal centrifugal, single stage	8,030 lpm @ 30 m H ₂ O (2,120 gpm @ 90 ft H ₂ O)	2	1
5	High-Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 6,050 lpm @ 1,800 m H ₂ O (1,600 gpm @ 6,100 ft H ₂ O)	2	1
6	High-Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,720 lpm @ 220 m H ₂ O (450 gpm @ 730 ft H ₂ O)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded Screw	28 m ³ /min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m ³ /min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	368 GJ/hr (348 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	131,900 lpm @ 20 m H ₂ O (34,800 gpm @ 70 ft H ₂ O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H ₂ O (1,000 gpm @ 350 ft H ₂ O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H ₂ O (700 gpm @ 250 ft H ₂ O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	5,730 lpm @ 20 m H ₂ O (1,510 gpm @ 60 ft H ₂ O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	2,870 lpm @ 270 m H ₂ O (760 gpm @ 880 ft H ₂ O)	4	1
16	Filtered Water Pumps	Stainless steel, single suction	3,350 lpm @ 50 m H ₂ O (890 gpm @ 160 ft H ₂ O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	1,610,000 liter (425,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	1,490 lpm (390 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0

Case B5B – Account 4: Gasifier, ASU, and Accessories Including Low Temperature Heat Recovery and Fuel Gas Saturation

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized slurry-feed, entrained bed	2,900 tonne/day, 5.6 MPa (3,200 tpd, 815 psia)	2	0
2	Synthesis Gas Cooler	Vertical downflow radiant heat exchanger with outlet quench chamber	256,000 kg/hr (564,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	350,000 kg/hr (772,000 lb/hr) Design efficiency 90%	2	0
4	Raw Gas Coolers	Shell and tube with condensate drain	388,000 kg/hr (856,000 lb/hr)	8	0
5	Raw Gas Knockout Drum	Vertical with mist eliminator	304,000 kg/hr, 35°C, 5.2 MPa (671,000 lb/hr, 95°F, 750 psia)	2	0
6	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	350,000 kg/hr (772,000 lb/hr) syngas	2	0
7	ASU Main Air Compressor	Centrifugal, multi-stage	6,000 m ³ /min @ 1.3 MPa (210,000 scfm @ 190 psia)	2	0
8	Cold Box	Vendor design	2,400 tonne/day (2,600 tpd) of 95% purity oxygen	2	0
9	Oxygen Compressor	Centrifugal, multi-stage	1,000 m ³ /min (42,000 scfm) Suction - 1.0 MPa (130 psia) Discharge - 6.5 MPa (940 psia)	2	0
10	Primary Nitrogen Compressor	Centrifugal, multi-stage	4,000 m ³ /min (134,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0
11	Secondary Nitrogen Compressor	Centrifugal, single-stage	550 m ³ /min (19,000 scfm) Suction - 1.3 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
12	Syngas Dilution Nitrogen Boost Compressor	Centrifugal, single-stage	2,000 m ³ /min (70,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 3.2 MPa (470 psia)	2	0

Case B5B – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	304,000 kg/hr (670,000 lb/hr), 35°C (95°F), 5.1 MPa (745 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	304,000 kg/hr (670,000 lb/hr), 35°C (95°F), 5.1 MPa (745 psia)	2	0
3	Sulfur Plant	Claus type	146 tonne/day (161 tpd)	1	0
4	Water Gas Shift Reactors	Fixed bed, catalytic	194,000 kg/hr (428,000 lb/hr), 227°C (440°F), 5.4 MPa (790 psia)	4	0
5	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 157 GJ/hr (149 MMBtu/hr) Exchanger 2: 3 GJ/hr (3 MMBtu/hr)	4	0
6	Acid Gas Removal Plant	Two-stage Selexol	311,000 kg/hr (685,000 lb/hr), 35°C (95°F), 5.1 MPa (740 psia)	2	0
7	Hydrogenation Reactor	Fixed bed, catalytic	18,000 kg/hr (40,000 lb/hr), 232°C (450°F), 0.1 MPa (12.3 psia)	1	0
8	Tail Gas Recycle Compressor	Centrifugal	14,000 kg/hr (30,000 lb/hr) each	1	0
9	CO ₂ Compressor	Integrally geared, multi-stage centrifugal	1,100 m ³ /min @ 15.3 MPa (40,000 scfm @ 2,215 psia)	4	0

Case B5B – Account 6: Combustion Turbine and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	State-of-the-art 2008 F-class	230 MW	2	0
2	Combustion Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0
3	Syngas Expansion Turbine/Generator	Turbo Expander	49,600 kg/h (109,300 lb/h) 5.1 MPa (735 psia) Inlet 3.2 MPa (460 psia) Outlet	2	0

Case B5B – Account 7: HRSG, Ducting, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.5 m (28 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 341,975 kg/hr, 12.4 MPa/534°C (753,925 lb/hr, 1,800 psig/994°F)	2	0

Case B5B – Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	277 MW 12.4 MPa/534°C/534°C (1,800 psig/994°F/994°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	310 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,660GJ/hr (1,580 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0

Case B5B – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	465,000 lpm @ 30 m (123,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 2,590 GJ/hr (2,460 MMBtu/hr) heat duty	1	0

Case B5B – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	254,000 liters (67,000 gal)	2	0
2	Slag Crusher	Roll	13 tonne/hr (15 tph)	2	0
3	Slag Depressurizer	Lock Hopper	13 tonne/hr (15 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	153,000 liters (40,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	69,000 liters (18,000 gal)	2	0
6	Slag Conveyor	Drag chain	13 tonne/hr (15 tph)	2	0
7	Slag Separation Screen	Vibrating	13 tonne/hr (15 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	13 tonne/hr (15 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	217,000 liters (57,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	60 lpm @ 14 m H ₂ O (10 gpm @ 46 ft H ₂ O)	2	2
11	Grey Water Storage Tank	Field erected	69,000 liters (18,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	250 lpm @ 560 m H ₂ O (60 gpm @ 1,850 ft H ₂ O)	2	2
13	Slag Storage Bin	Vertical, field erected	1,000 tonne (1,100 tons)	2	0
14	Unloading Equipment	Telescoping chute	110 tonne/hr (120 tph)	1	0

Case B5B – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 80 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 48 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 7 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case B5B – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.4.11 Case B5B - Cost Estimating

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-104 shows a detailed breakdown of the capital costs; Exhibit 3-105 shows the owner's costs, TOC, and TASC; Exhibit 3-106 shows the initial and annual O&M costs; and Exhibit 3-107 shows the COE breakdown.

The estimated TPC of the GEE gasifier with CO₂ capture is \$3,387/kW. Process contingency represents 4.2 percent of the TPC, and project contingency represents 13.8 percent. The COE, including CO₂ T&S costs of \$9.2/MWh, is \$144.7/MWh.

Exhibit 3-104 Case B5B total plant cost details

Case:		B5B – GEE Radiant IGCC w/ CO ₂					Estimate Type:			Conceptual	
Plant Size (MW,net):		543					Cost Base:			Jun 2011	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
1		Coal & Sorbent Handling									
1.1	Coal Receive & Unload	\$4,552	\$0	\$2,194	\$0	\$6,746	\$675	\$0	\$1,484	\$8,905	\$16
1.2	Coal Stackout & Reclaim	\$5,882	\$0	\$1,407	\$0	\$7,289	\$729	\$0	\$1,604	\$9,621	\$18
1.3	Coal Conveyors & Yd Crush	\$5,469	\$0	\$1,392	\$0	\$6,861	\$686	\$0	\$1,509	\$9,056	\$17
1.4	Other Coal Handling	\$1,431	\$0	\$322	\$0	\$1,753	\$175	\$0	\$386	\$2,314	\$4
1.9	Coal & Sorbent Hnd. Foundations	\$0	\$3,040	\$7,944	\$0	\$10,984	\$1,098	\$0	\$2,417	\$14,499	\$27
Subtotal		\$17,335	\$3,040	\$13,259	\$0	\$33,633	\$3,363	\$0	\$7,399	\$44,396	\$82
2		Coal & Sorbent Prep & Feed									
2.1	Coal Crushing & Drying	w/2.3	\$0	w/2.3	\$0	\$0	\$0	\$0	\$0	\$0	\$0
2.2	Prepared Coal Storage & Feed	\$1,944	\$467	\$301	\$0	\$2,712	\$271	\$0	\$597	\$3,580	\$7
2.3	Slurry Prep & Feed	\$26,551	\$0	\$11,574	\$0	\$38,125	\$3,812	\$1,906	\$8,769	\$52,612	\$97
2.4	Misc. Coal Prep & Feed	\$1,069	\$781	\$2,301	\$0	\$4,151	\$415	\$0	\$913	\$5,479	\$10
2.9	Coal & Sorbent Feed Foundation	\$0	\$3,922	\$3,365	\$0	\$7,287	\$729	\$0	\$1,603	\$9,619	\$18
Subtotal		\$29,564	\$5,170	\$17,541	\$0	\$52,275	\$5,227	\$1,906	\$11,882	\$71,290	\$131
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	\$3,381	\$5,835	\$3,059	\$0	\$12,275	\$1,227	\$0	\$2,700	\$16,203	\$30
3.2	Water Makeup & Pretreating	\$876	\$91	\$481	\$0	\$1,448	\$145	\$0	\$478	\$2,071	\$4
3.3	Other Feedwater Subsystems	\$1,901	\$628	\$561	\$0	\$3,090	\$309	\$0	\$680	\$4,079	\$8
3.4	Service Water Systems	\$512	\$1,022	\$3,521	\$0	\$5,055	\$505	\$0	\$1,668	\$7,228	\$13
3.5	Other Boiler Plant Systems	\$2,761	\$1,032	\$2,539	\$0	\$6,331	\$633	\$0	\$1,393	\$8,358	\$15
3.6	FO Supply Sys & Nat Gas	\$17,735	\$710	\$658	\$0	\$19,104	\$1,910	\$0	\$4,203	\$25,217	\$46
3.7	Waste Treatment Equipment	\$1,185	\$0	\$734	\$0	\$1,919	\$192	\$0	\$633	\$2,745	\$5
3.8	Misc. Power Plant Equipment	\$1,326	\$177	\$689	\$0	\$2,193	\$219	\$0	\$724	\$3,135	\$6
Subtotal		\$29,677	\$9,494	\$12,243	\$0	\$51,415	\$5,142	\$0	\$12,479	\$69,035	\$127
4		Gasifier & Accessories									
4.1	Syngas Cooler Gasifier System	\$140,246	\$0	\$77,258	\$0	\$217,504	\$21,750	\$30,249	\$41,297	\$310,800	\$572
4.2	Syngas Cooler	w/4.1	\$0	w/ 4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.3	ASU & Oxidant Compression	\$234,789	\$0	w/equip.	\$0	\$234,789	\$23,479	\$0	\$25,827	\$284,094	\$523
4.4	Scrubber & LT Heat Recovery	\$7,292	\$5,935	\$6,135	\$0	\$19,363	\$1,936	\$0	\$4,260	\$25,560	\$47
4.5	Black Water & Sour Gas Section	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.6	Soot Recovery & SARU	\$2,224	\$1,056	\$2,073	\$0	\$5,353	\$535	\$0	\$1,178	\$7,066	\$13
4.8	Major Component Rigging	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.9	Gasification Foundations	\$0	\$7,429	\$6,482	\$0	\$13,911	\$1,391	\$0	\$3,826	\$19,128	\$35
Subtotal		\$384,551	\$14,421	\$91,949	\$0	\$490,920	\$49,092	\$30,249	\$76,387	\$646,648	\$1,190
5A		Gas Cleanup & Piping									
5A.1	Double Stage Selexol	\$160,990	\$0	w/equip.	\$0	\$160,990	\$16,099	\$32,198	\$41,857	\$251,145	\$462
5A.2	Elemental Sulfur Plant	\$12,451	\$2,427	\$15,954	\$0	\$30,833	\$3,083	\$0	\$6,783	\$40,699	\$75
5A.3	Mercury Removal	\$1,973	\$0	\$1,491	\$0	\$3,464	\$346	\$173	\$797	\$4,780	\$9
5A.4	Shift Reactors	\$11,566	\$0	\$4,624	\$0	\$16,190	\$1,619	\$0	\$3,562	\$21,370	\$39
5A.7	Fuel Gas Piping	\$0	\$812	\$531	\$0	\$1,344	\$134	\$0	\$296	\$1,774	\$3

Case:		B5B – GEE Radiant IGCC w/ CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		543					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
5A.9	HGCU Foundations	\$0	\$735	\$495	\$0	\$1,231	\$123	\$0	\$406	\$1,760	\$3
	Subtotal	\$186,980	\$3,974	\$23,096	\$0	\$214,051	\$21,405	\$32,371	\$53,701	\$321,528	\$592
5B		CO₂ Compression									
5B.2	CO ₂ Compression & Drying	\$40,237	\$6,036	\$17,430	\$0	\$63,703	\$6,370	\$0	\$14,015	\$84,088	\$155
	Subtotal	\$40,237	\$6,036	\$17,430	\$0	\$63,703	\$6,370	\$0	\$14,015	\$84,088	\$155
6		Combustion Turbine & Accessories									
6.1	Combustion Turbine Generator	\$111,210	\$0	\$7,881	\$0	\$119,091	\$11,909	\$11,909	\$14,291	\$157,201	\$289
6.2	Syngas Expander	\$6,691	\$0	\$918	\$0	\$7,609	\$761	\$0	\$1,256	\$9,626	\$18
6.9	Combustion Turbine Foundations	\$0	\$1,016	\$1,175	\$0	\$2,191	\$219	\$0	\$723	\$3,133	\$6
	Subtotal	\$117,901	\$1,016	\$9,975	\$0	\$128,892	\$12,889	\$11,909	\$16,270	\$169,960	\$313
7		HRSG, Ducting, & Stack									
7.1	Heat Recovery Steam Generator	\$29,560	\$0	\$5,723	\$0	\$35,283	\$3,528	\$0	\$3,881	\$42,693	\$79
7.3	Ductwork	\$0	\$2,109	\$1,478	\$0	\$3,587	\$359	\$0	\$789	\$4,735	\$9
7.4	Stack	\$4,070	\$0	\$1,519	\$0	\$5,589	\$559	\$0	\$615	\$6,762	\$12
7.9	HRSG, Duct & Stack Foundations	\$0	\$775	\$778	\$0	\$1,553	\$155	\$0	\$512	\$2,220	\$4
	Subtotal	\$33,630	\$2,884	\$9,498	\$0	\$46,012	\$4,601	\$0	\$5,797	\$56,411	\$104
8		Steam Turbine Generator									
8.1	Steam TG & Accessories	\$37,100	\$0	\$5,758	\$0	\$42,858	\$4,286	\$0	\$4,714	\$51,858	\$95
8.2	Turbine Plant Auxiliaries	\$235	\$0	\$535	\$0	\$771	\$77	\$0	\$85	\$932	\$2
8.3	Condenser & Auxiliaries	\$3,290	\$0	\$1,868	\$0	\$5,158	\$516	\$0	\$567	\$6,241	\$11
8.4	Steam Piping	\$15,068	\$0	\$6,534	\$0	\$21,602	\$2,160	\$0	\$5,941	\$29,703	\$55
8.9	TG Foundations	\$0	\$1,108	\$1,958	\$0	\$3,067	\$307	\$0	\$1,012	\$4,385	\$8
	Subtotal	\$55,693	\$1,108	\$16,654	\$0	\$73,456	\$7,346	\$0	\$12,319	\$93,121	\$171
9		Cooling Water System									
9.1	Cooling Towers	\$4,490	\$0	\$1,360	\$0	\$5,850	\$585	\$0	\$965	\$7,400	\$14
9.2	Circulating Water Pumps	\$2,175	\$0	\$163	\$0	\$2,338	\$234	\$0	\$386	\$2,957	\$5
9.3	Circ. Water System Auxiliaries	\$189	\$0	\$27	\$0	\$216	\$22	\$0	\$36	\$273	\$1
9.4	Circ. Water Piping	\$0	\$8,386	\$2,031	\$0	\$10,417	\$1,042	\$0	\$2,292	\$13,751	\$25
9.5	Make-up Water System	\$484	\$0	\$665	\$0	\$1,150	\$115	\$0	\$253	\$1,517	\$3
9.6	Component Cooling Water Sys	\$959	\$1,147	\$787	\$0	\$2,893	\$289	\$0	\$636	\$3,818	\$7
9.9	Circ. Water System Foundations	\$0	\$2,738	\$4,866	\$0	\$7,604	\$760	\$0	\$2,509	\$10,874	\$20
	Subtotal	\$8,296	\$12,271	\$9,900	\$0	\$30,467	\$3,047	\$0	\$7,077	\$40,591	\$75
10		Ash & Spent Sorbent Handling Systems									
10.1	Slag Dewatering & Cooling	\$14,630	\$8,067	\$16,277	\$0	\$38,975	\$3,897	\$0	\$4,287	\$47,159	\$87
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.6	Ash Storage Silos	\$710	\$0	\$767	\$0	\$1,476	\$148	\$0	\$244	\$1,868	\$3
10.7	Ash Transport & Feed Equipment	\$977	\$0	\$228	\$0	\$1,205	\$120	\$0	\$199	\$1,524	\$3
10.8	Misc. Ash Handling Equipment	\$1,470	\$1,801	\$534	\$0	\$3,806	\$381	\$0	\$628	\$4,814	\$9
10.9	Ash/Spent Sorbent Foundation	\$0	\$60	\$78	\$0	\$138	\$14	\$0	\$46	\$197	\$0
	Subtotal	\$17,787	\$9,928	\$17,885	\$0	\$45,600	\$4,560	\$0	\$5,403	\$55,563	\$102
11		Accessory Electric Plant									

Case:		B5B – GEE Radiant IGCC w/ CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		543					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
11.1	Generator Equipment	\$1,152	\$0	\$1,122	\$0	\$2,274	\$227	\$0	\$250	\$2,752	\$5
11.2	Station Service Equipment	\$5,513	\$0	\$507	\$0	\$6,020	\$602	\$0	\$662	\$7,285	\$13
11.3	Switchgear & Motor Control	\$10,176	\$0	\$1,891	\$0	\$12,068	\$1,207	\$0	\$1,991	\$15,266	\$28
11.4	Conduit & Cable Tray	\$0	\$5,170	\$15,938	\$0	\$21,109	\$2,111	\$0	\$5,805	\$29,024	\$53
11.5	Wire & Cable	\$0	\$9,969	\$6,065	\$0	\$16,034	\$1,603	\$0	\$4,409	\$22,047	\$41
11.6	Protective Equipment	\$0	\$806	\$2,995	\$0	\$3,801	\$380	\$0	\$627	\$4,808	\$9
11.7	Standby Equipment	\$274	\$0	\$273	\$0	\$547	\$55	\$0	\$90	\$692	\$1
11.8	Main Power Transformers	\$19,599	\$0	\$172	\$0	\$19,771	\$1,977	\$0	\$3,262	\$25,011	\$46
11.9	Electrical Foundations	\$0	\$180	\$491	\$0	\$672	\$67	\$0	\$222	\$961	\$2
Subtotal		\$36,715	\$16,126	\$29,455	\$0	\$82,297	\$8,230	\$0	\$17,319	\$107,845	\$199
12		Instrumentation & Control									
12.1	IGCC Control Equipment	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.2	Combustion Turbine Control	w/6.1	\$0	w/6.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.3	Steam Turbine Control	w/8.1	\$0	w/8.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.4	Other Major Component Control	\$1,321	\$0	\$900	\$0	\$2,221	\$222	\$111	\$383	\$2,937	\$5
12.5	Signal Processing Equipment	w/12.7	\$0	w/12.7	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.6	Control Boards, Panels & Racks	\$304	\$0	\$199	\$0	\$502	\$50	\$25	\$116	\$693	\$1
12.7	Computer & Accessories	\$7,046	\$0	\$230	\$0	\$7,277	\$728	\$364	\$837	\$9,205	\$17
12.8	Instrument Wiring & Tubing	\$0	\$2,713	\$5,135	\$0	\$7,847	\$785	\$392	\$2,256	\$11,280	\$21
12.9	Other I & C Equipment	\$4,710	\$0	\$2,334	\$0	\$7,044	\$704	\$352	\$1,215	\$9,316	\$17
Subtotal		\$13,381	\$2,713	\$8,797	\$0	\$24,891	\$2,489	\$1,245	\$4,807	\$33,431	\$62
13		Improvements to Site									
13.1	Site Preparation	\$0	\$123	\$2,799	\$0	\$2,922	\$292	\$0	\$964	\$4,178	\$8
13.2	Site Improvements	\$0	\$2,189	\$3,095	\$0	\$5,284	\$528	\$0	\$1,744	\$7,556	\$14
13.3	Site Facilities	\$3,923	\$0	\$4,404	\$0	\$8,327	\$833	\$0	\$2,748	\$11,907	\$22
Subtotal		\$3,923	\$2,312	\$10,297	\$0	\$16,532	\$1,653	\$0	\$5,456	\$23,641	\$44
14		Buildings & Structures									
14.1	Combustion Turbine Area	\$0	\$317	\$179	\$0	\$496	\$50	\$0	\$109	\$655	\$1
14.2	Steam Turbine Building	\$0	\$2,849	\$4,056	\$0	\$6,904	\$690	\$0	\$1,139	\$8,734	\$16
14.3	Administration Building	\$0	\$1,057	\$766	\$0	\$1,823	\$182	\$0	\$301	\$2,306	\$4
14.4	Circulation Water Pumphouse	\$0	\$198	\$105	\$0	\$303	\$30	\$0	\$50	\$384	\$1
14.5	Water Treatment Buildings	\$0	\$720	\$702	\$0	\$1,422	\$142	\$0	\$235	\$1,799	\$3
14.6	Machine Shop	\$0	\$541	\$370	\$0	\$911	\$91	\$0	\$150	\$1,152	\$2
14.7	Warehouse	\$0	\$873	\$563	\$0	\$1,437	\$144	\$0	\$237	\$1,818	\$3
14.8	Other Buildings & Structures	\$0	\$523	\$407	\$0	\$930	\$93	\$0	\$205	\$1,228	\$2
14.9	Waste Treating Building & Str.	\$0	\$1,169	\$2,233	\$0	\$3,403	\$340	\$0	\$749	\$4,492	\$8
Subtotal		\$0	\$8,248	\$9,382	\$0	\$17,630	\$1,763	\$0	\$3,175	\$22,568	\$42
Total		\$975,671	\$98,741	\$297,361	\$0	\$1,371,773	\$137,177	\$77,680	\$253,485	\$1,840,115	\$3,387

Exhibit 3-105 Case B5B owner's costs

Description	\$/1,000	\$/kW
Pre-Production Costs		
6 Months All Labor	\$16,298	\$30
1 Month Maintenance Materials	\$3,682	\$7
1 Month Non-fuel Consumables	\$676	\$1
1 Month Waste Disposal	\$494	\$1
25% of 1 Months Fuel Cost at 100% CF	\$3,046	\$6
2% of TPC	\$36,802	\$68
Total	\$60,998	\$112
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	\$25,290	\$47
0.5% of TPC (spare parts)	\$9,201	\$17
Total	\$34,491	\$63
Other Costs		
Initial Cost for Catalyst and Chemicals	\$16,548	\$30
Land	\$900	\$2
Other Owner's Costs	\$276,017	\$508
Financing Costs	\$49,683	\$91
Total Overnight Costs (TOC)	\$2,278,752	\$4,195
TASC Multiplier (IOU, high-risk, 35 year)	1.140	
Total As-Spent Cost (TASC)	\$2,597,777	\$4,782

Exhibit 3-106 Case B5B initial and annual operating and maintenance costs

Case:	B5B – GEE Radiant IGCC w/ CO ₂			Cost Base:	Jun 2011
Plant Size (MW _{net}):	543	Heat Rate-net (Btu/kWh):	10,459	Capacity Factor (%):	80
Operating & Maintenance Labor					
Operating Labor			Operating Labor Requirements per Shift		
Operating Labor Rate (base):	39.70	\$/hour	Skilled Operator:	2.0	
Operating Labor Burden:	30.00	% of base	Operator:	10.0	
Labor O-H Charge Rate:	25.00	% of labor	Foreman:	1.0	
			Lab Tech's, etc.:	3.0	
			Total:	16.0	
Fixed Operating Costs					
			Annual Cost		
			(\$)	(\$/kW-net)	
Annual Operating Labor:				\$7,233,658	\$13.316
Maintenance Labor:				\$18,843,231	\$34.687
Administrative & Support Labor:				\$6,519,222	\$12.001
Property Taxes and Insurance:				\$36,802,296	\$67.747
Total:				\$69,398,406	\$127.751
Variable Operating Costs					
			(\$)	(\$/MWh-net)	
Maintenance Material:				\$35,348,580	\$9.28526
Consumables					
	Consumption			Cost (\$)	
	Initial Fill	Per Day	Per Unit	Initial Fill	
Water (/1000 gallons):	0	4,201	\$1.67	\$0	\$2,053,253
Makeup and Waste Water Treatment Chemicals (lbs):	0	25,026	\$0.27	\$0	\$1,957,230
Carbon (Mercury Removal) (lb):	135,182	231	\$5.50	\$743,501	\$371,751
Shift Catalyst (ft³):	6,246	4.28	\$771.99	\$4,822,025	\$964,405
Selexol Solution (gal):	298,498	95	\$36.79	\$10,982,126	\$1,020,094
Claus Catalyst (ft³):	w/equip	2.01	\$203.15	\$0	\$119,487
Subtotal:				\$16,547,652	\$6,486,219
Waste Disposal					
Spent Mercury Catalyst (lb.):	0	231	\$0.65	\$0	\$43,941
Flyash (ton):	0	0	\$0.00	\$0	\$0
Slag (ton):	0	641	\$25.11	\$0	\$4,701,292
Subtotal:				\$0	\$4,745,232
By-Products					
Sulfur (tons):	0	146	\$0.00	\$0	\$0
Subtotal:				\$0	\$0
Variable Operating Costs Total:				\$16,547,652	\$46,580,031
Fuel Cost					
Illinois Number 6 (ton):	0	5,844	\$68.54	\$0	\$116,961,258
Total:				\$0	\$116,961,258

Exhibit 3-107 Case B5B COE breakdown

Component	Value, \$/MWh	Percentage
Capital	74.2	55%
Fixed	18.2	13%
Variable	12.2	9%
Fuel	30.7	23%
Total (Excluding T&S)	135.4	N/A
CO ₂ T&S	9.2	7%
Total (Including T&S)	144.7	N/A

3.4.12 Case B5B-Q - GEE Quench IGCC with CO₂ Capture

Case B5B-Q is configured to produce electric power with CO₂ capture. The plant configuration is the same as Case B5B, namely two gasifier trains, two state-of-the-art 2008 F-class turbines, two HRSGs, and one steam turbine.

The process description for Case B5B-Q is similar to Case B5B with the exception that the gasifier is operated in quench-only mode. A BFD and stream tables for Case B5B-Q are shown in Exhibit 3-108 and Exhibit 3-109, respectively. Rather than repeating the entire process description, only differences from Case B5B are reported here.

3.4.12.1 Gasification

The gasification process is the same as Case B5B with the exception that the syngas exiting the gasifier passes through a water quench where the temperature is reduced from 1,316°C (2,400°F) to 288°C (550°F). The total coal feed to the two gasifiers is 5,301 tonnes/day (5,844 tpd) (stream 6) and the ASU provides 4,343 tonnes/day (4,787 tpd) of 95 percent oxygen to the gasifier and Claus plant (streams 3 and 5).

3.4.12.2 Raw Gas Cooling/Particulate Removal

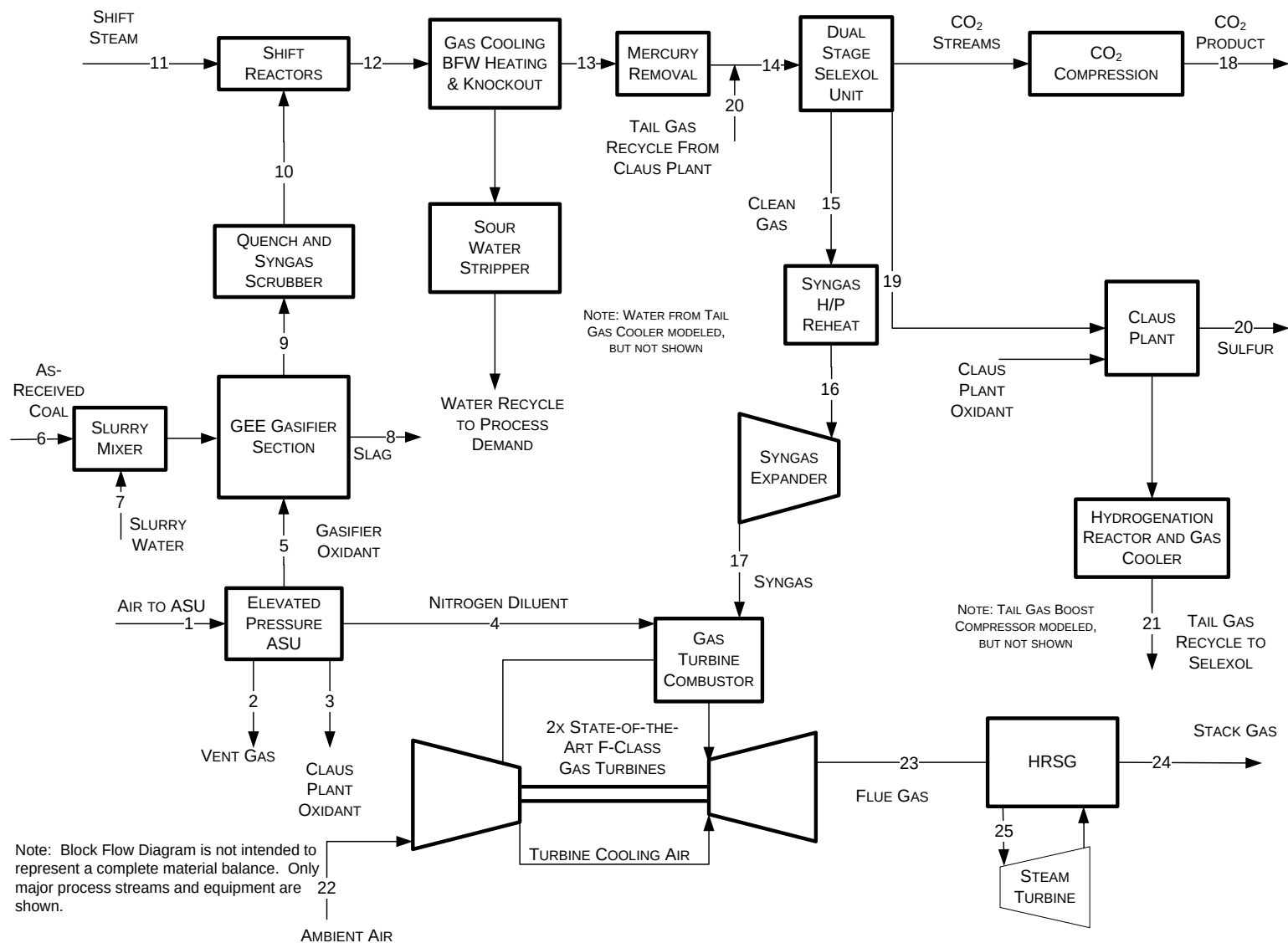
Particulate is largely removed in the quench tank and no additional heat recovery occurs prior to the syngas scrubber.

3.4.12.3 Syngas Scrubber/Sour Water Stripper

The outlet temperature from the syngas scrubber is 227°C (440°F) (stream 10).

3.4.12.4 Sour Gas Shift

The SGS process is the same as described for Case B5B with a 97 percent overall conversion of the CO to CO₂. The warm syngas from the second stage of SGS (stream 12) is cooled to 244°C (471°F) by heating the syngas entering the first shift reactor. The syngas is then cooled to 193°C (380°F) by producing IP steam that is sent to the RH in the HRSG. The SGS catalyst also serves to hydrolyze COS thus eliminating the need for a separate COS hydrolysis reactor. Following the second SGS cooler the syngas is further cooled to 35°C (95°F) before entering the mercury removal beds.

Exhibit 3-108 Case B5B-Q block flow diagram, GEE quench-only IGCC with CO₂ capture


Source: NETL

Exhibit 3-109 Case B5B-Q stream table, GEE quench-only IGCC with capture

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0092	0.0165	0.0318	0.0023	0.0318	0.0000	0.0000	0.0000	0.0086	0.0053	0.0000	0.0053	0.0071	0.0071
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0011	0.0007	0.0000	0.0007	0.0009	0.0009
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3576	0.2197	0.0000	0.0060	0.0081	0.0081
CO ₂	0.0003	0.0054	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1380	0.0848	0.0000	0.2985	0.4017	0.4053
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0001	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	0.3406	0.2093	0.0000	0.4229	0.5690	0.5647
H ₂ O	0.0099	0.1355	0.0000	0.0003	0.0000	0.0000	0.9997	0.0000	0.1369	0.4700	1.0000	0.2562	0.0012	0.0012
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0001	0.0000	0.0001	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0045	0.0000	0.0046	0.0061	0.0061
N ₂	0.7732	0.7077	0.0178	0.9920	0.0178	0.0000	0.0000	0.0000	0.0070	0.0043	0.0000	0.0043	0.0058	0.0065
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0000	0.0019	0.0013	0.0000	0.0013	0.0000	0.0000
O ₂	0.2074	0.1349	0.9504	0.0054	0.9504	0.0000	0.0000	0.0000	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	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	27,367	1,660	97	20,046	5,526	0	5,037	0	23,121	37,625	0	37,625	27,964	28,358
V-L Flowrate (kg/hr)	789,736	45,606	3,132	562,476	177,816	0	90,741	0	465,212	726,201	0	726,201	552,117	564,722
Solids Flowrate (kg/hr)	0	0	0	0	0	220,890	0	24,235	0	0	0	0	0	0
Temperature (°C)	15	18	32	93	32	15	146	1,316	1,316	227	288	249	35	35
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	0.10	5.79	5.62	5.62	5.58	5.52	5.45	5.17	5.14
Steam Table Enthalpy (kJ/kg) ^A	30.23	35.62	26.67	92.50	26.67	---	559.14	---	2,635.52	1,479.99	2,918.18	1,017.28	36.94	36.25
AspenPlus Enthalpy (kJ/kg) ^B	-97.58	-1,483.25	4.13	63.47	4.13	-2,116.95	-15,408.22	-727.24	-4,038.89	-8,545.15	-13,053.16	-8,965.22	-8,082.42	-8,084.87
Density (kg/m ³)	1.2	1.5	11.0	24.4	11.0	---	867.1	---	8.5	25.9	25.6	24.5	41.0	41.2
V-L Molecular Weight	28.857	27.479	32.181	28.060	32.181	---	18.015	---	20.121	19.301	18.015	19.301	19.744	19.914
V-L Flowrate (lb _{mole} /hr)	60,334	3,659	215	44,193	12,182	0	11,105	0	50,972	82,949	0	82,949	61,649	62,518
V-L Flowrate (lb/hr)	1,741,069	100,544	6,906	1,240,047	392,017	0	200,051	0	1,025,616	1,600,998	0	1,600,998	1,217,210	1,244,998
Solids Flowrate (lb/hr)	0	0	0	0	0	486,978	0	53,430	0	0	0	0	0	0
Temperature (°F)	59	65	90	199	90	59	295	2,400	2,400	440	550	481	95	95
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	14.7	840.0	815.0	815.0	810.0	800.0	790.0	750.0	745.0
Steam Table Enthalpy (Btu/lb) ^A	13.0	15.3	11.5	39.8	11.5	---	240.4	---	1,133.1	636.3	1,254.6	437.4	15.9	15.6
AspenPlus Enthalpy (Btu/lb) ^B	-42.0	-637.7	1.8	27.3	1.8	-910.1	-6,624.3	-312.7	-1,736.4	-3,673.8	-5,611.9	-3,854.3	-3,474.8	-3,475.9
Density (lb/ft ³)	0.076	0.091	0.687	1.521	0.687	---	54.131	---	0.530	1.619	1.597	1.529	2.562	2.570

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 3-109 Case B5B-Q stream table, GEE quench-only IGCC with capture (continued)

	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction											
Ar	0.0115	0.0115	0.0115	0.0002	0.0018	0.0000	0.0103	0.0092	0.0091	0.0091	0.0000
CH ₄	0.0015	0.0015	0.0015	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0130	0.0130	0.0130	0.0002	0.0023	0.0000	0.0061	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0501	0.0501	0.0501	0.9948	0.5218	0.0000	0.6606	0.0003	0.0083	0.0083	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂	0.9133	0.9133	0.9133	0.0048	0.1029	0.0000	0.2596	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0001	0.0001	0.0001	0.0000	0.0225	0.0000	0.0017	0.0099	0.1222	0.1222	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ S	0.0000	0.0000	0.0000	0.0000	0.3472	0.0000	0.0043	0.0000	0.0000	0.0000	0.0000
N ₂	0.0105	0.0105	0.0105	0.0000	0.0008	0.0000	0.0574	0.7732	0.7541	0.7541	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
O ₂	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.1064	0.1064	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	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mole} /hr)	17,424	17,424	17,424	10,415	496	0	394	110,253	139,653	139,653	20,013
V-L Flowrate (kg/hr)	90,438	90,438	90,438	456,223	17,655	0	12,604	3,181,557	3,834,472	3,834,472	360,534
Solids Flowrate (kg/hr)	0	0	0	0	0	5,523	0	0	0	0	0
Temperature (°C)	35	241	195	51	48	178	38	15	562	132	534
Pressure (MPa, abs)	5.14	5.10	3.17	15.27	0.16	0.12	5.51	0.10	0.10	0.10	12.51
Steam Table Enthalpy (kJ/kg) ^A	194.96	1,385.17	1,117.49	-162.29	74.70	---	5.84	30.23	834.65	343.46	3,432.96
AspenPlus Enthalpy (kJ/kg) ^B	-4,047.98	-2,857.76	-3,125.45	-9,120.15	-6,110.54	141.95	-8,192.39	-97.58	-588.07	-1,079.26	-12,537.35
Density (kg/m ³)	10.2	6.1	4.2	641.8	2.2	5,280.6	77.3	1.2	0.4	0.9	36.7
V-L Molecular Weight	5.190	5.190	5.190	43.804	35.590	---	31.966	28.857	27.457	27.457	18.015
V-L Flowrate (lb _{mole} /hr)	38,414	38,414	38,414	22,961	1,094	0	869	243,066	307,882	307,882	44,120
V-L Flowrate (lb/hr)	199,383	199,383	199,383	1,005,800	38,923	0	27,788	7,014,133	8,453,564	8,453,564	794,842
Solids Flowrate (lb/hr)	0	0	0	0	0	12,177	0	0	0	0	0
Temperature (°F)	95	465	383	124	119	352	100	59	1,044	270	994
Pressure (psia)	745.0	740.0	460.0	2,214.7	23.7	17.3	799.5	14.7	15.2	15.2	1,814.7
Steam Table Enthalpy (Btu/lb) ^A	83.8	595.5	480.4	-69.8	32.1	---	2.5	13.0	358.8	147.7	1,475.9
AspenPlus Enthalpy (Btu/lb) ^B	-1,740.3	-1,228.6	-1,343.7	-3,921.0	-2,627.1	61.0	-3,522.1	-42.0	-252.8	-464.0	-5,390.1
Density (lb/ft ³)	0.636	0.381	0.261	40.066	0.137	329.655	4.823	0.076	0.026	0.053	2.293

^ASteam table reference conditions are 32.02°F & 0.089 psia^BAspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

3.4.12.5 Mercury Removal and Acid Gas Removal

Mercury removal and the AGR process are the same as in Case B5B.

The AGR process in Case B5B-Q is a two-stage Selexol process where H₂S is removed in the first stage and CO₂ in the second stage of absorption as previously described in Section 3.1.5. The process results in three product streams, the clean syngas, a CO₂-rich stream and an acid gas feed to the Claus plant. The acid gas (stream 18) contains 35 percent H₂S and 52 percent CO₂ with the balance primarily H₂.

3.4.12.6 CO₂ Compression and Dehydration

The CO₂ compression and dehydration scheme is the same as for Case B5B.

3.4.12.7 Claus Unit

The Claus plant is the same as Case B5B with the following exceptions:

- 5,523 kg/hr (12,177 lb/hr) of sulfur (stream 19) are produced
- The waste heat boiler generates 12,957 kg/hr (28,564 lb/hr) of 3.0 MPa (430 psia) steam

3.4.12.8 Power Block

Clean syngas from the AGR plant is heated to 241°C (465°F) first using hot water from the syngas scrubber followed by HP BFW before passing through an expansion turbine. The clean syngas (stream 16) is diluted with nitrogen (stream 4) and then enters the CT burner. There is no integration between the CT and the ASU in this case. The exhaust gas (stream 22) exits the CT at 562°C (1,044°F) and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 132°C (270°F) (stream 23) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available steam turbine using a nominal 12.4 MPa/538°C/538°C (1,800 psig/1,000°F/1,000°F) steam cycle.

3.4.12.9 Air Separation Unit

The same elevated pressure ASU is used in Case B5B-Q and produces 4,343 tonnes/day (4,787 tpd) of 95 mol% oxygen and 14,595 tonnes/day (16,087 tpd) of nitrogen. There is no integration between the ASU and the CT in the CO₂ capture cases.

3.4.13 Case B5B-Q - Performance Results

The plant produces a net output of 494 MW at a net plant efficiency of 29.7 percent (HHV basis). Overall performance for the entire plant is summarized in Exhibit 3-110 and Exhibit 3-111, which includes auxiliary power requirements. The ASU accounts for 60 percent of the auxiliary load between the main air compressor, nitrogen compressor, oxygen compressor, and ASU auxiliaries. The two-stage Selexol process and CO₂ compression account for an additional 26 percent of the auxiliary power load. The BFW pumps and cooling water system (CWP and cooling tower fan) compose about 5 percent of the load, leaving 9 percent of the auxiliary load for all other systems.

Exhibit 3-110 Case B5B-Q plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	7
Steam Turbine Power, MWe	214
Total Gross Power, MWe	684
Air Separation Unit Main Air Compressor, kWe	67,350
Oxygen Compressor, kWe	10,640
Nitrogen Compressors, kWe	35,630
CO ₂ Compression, kWe	31,130
Acid Gas Removal, kWe	19,210
Balance of Plant, kWe	26,450
Total Auxiliaries, MWe	190
Net Power, MWe	494
HHV Net Plant Efficiency (%)	29.7%
HHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	12,138 (11,505)
HHV Cold Gas Efficiency, %	79.2%
HHV Combustion Turbine Efficiency, %	36.0%
LHV Net Plant Efficiency (%)	30.7%
LHV Net Plant Heat Rate, kJ/kWh (Btu/kWh)	11,708 (11,097)
LHV Cold Gas Efficiency, %	75.9%
LHV Combustion Turbine Efficiency, %	42.4%
Steam Turbine Cycle Efficiency, %	38.1%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	9,443 (8,950)
Condenser Duty, GJ/hr (MMBtu/hr)	1,452 (1,376)
As-Received Coal Feed, kg/hr (lb/hr)	220,890 (486,978)
HHV Thermal Input, kWt	1,664,962
LHV Thermal Input, kWt	1,605,876
Raw Water Withdrawal, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.047 (12.4)
Raw Water Consumption, (m ³ /min)/MW _{net} (gpm/MW _{net})	0.038 (10.1)
O ₂ :Coal	0.761

Exhibit 3-111 Case B5B-Q plant power summary

Power Summary	
Combustion Turbine Power, MWe	464
Sweet Gas Expander Power, MWe	7
Steam Turbine Power, MWe	214
Total Gross Power, MWe	684
Auxiliary Load Summary	
Coal Handling, kWe	470
Coal Milling, kWe	2,270
Sour Water Recycle Slurry Pump, kWe	200
Slag Handling, kWe	1,160
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	67,350
Oxygen Compressor, kWe	10,640
Nitrogen Compressors, kWe	35,630
Feedwater Pumps, kWe	2,620
Quench Water Pump, kWe	1,270
Syngas Recycle Compressor, kWe	0
CO ₂ Compression, kWe	31,130
Scrubber Pumps, kWe	480
Acid Gas Removal, kWe	19,210
Combustion Turbine Auxiliaries, kWe	1,000
Steam Turbine Auxiliaries, kWe	100
Condensate Pumps, kWe	230
Circulating Water Pumps, kWe	4,830
Ground Water Pumps, kWe	560
Cooling Tower Fans, kWe	2,500
Claus Plant/TGTU Auxiliaries, kWe	250
Claus Plant TG Recycle Compressor, kWe	1,800
Miscellaneous Balance of Plant ^{A,B} , kWe	3,000
Transformer Losses, kWe	2,710
Total Auxiliaries, MWe	190
Net Power, MWe	494

^AIncludes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.4.13.1 Environmental Performance

The environmental targets for emissions of Hg, NO_x, SO₂, and PM were presented in Section 2.3. A summary of the plant air emissions for Case B5B-Q is presented in Exhibit 3-112. All HCl is assumed to be removed and is, therefore, not reported.

Exhibit 3-112 Case B5B-Q air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year) ^A	kg/MWh (lb/MWh) ^B
SO ₂	0.001 (0.002)	39 (43)	0.008 (0.018)
NO _x	0.021 (0.049)	878 (968)	0.183 (0.404)
Particulates	0.003 (0.007)	128 (141)	0.027 (0.059)
Hg	1.55E-7 (3.61E-7)	0.007 (0.007)	1.36E-6 (3.00E-6)
CO ₂	8 (20)	358,178 (394,824)	75 (165)
CO ₂ ^C	-	-	104 (228)

^ACalculations based on an 80 percent capacity factor

^BEmissions are based on gross power except where otherwise noted

^CCO₂ emissions based on net power instead of gross power

The low level of SO₂ emissions is achieved by capturing the sulfur in the gas by the two-stage Selexol AGR process. As a result of achieving the 90 percent CO₂ removal target, the sulfur compounds are removed to an extent that exceeds the environmental target in Section 2.3. The clean syngas exiting the AGR process has a sulfur concentration of approximately 5 ppmv. This results in a concentration in the flue gas of less than less than 1 ppmv. The H₂S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is hydrogenated to convert all sulfur species to H₂S and then recycled back to the Selexol process, thereby eliminating the need for a tail gas treatment unit.

NO_x emissions are limited by nitrogen dilution to 15 ppmvd (as NO₂ @15 percent O₂). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process. This helps lower NO_x levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of the syngas quench in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of the mercury is captured from the syngas by dual activated carbon beds.

Ninety two percent of the CO₂ from the syngas is captured in the AGR system and compressed for sequestration. Because not all CO is converted to CO₂ in the shift reactors, the overall carbon removal is 90.0 percent.

The carbon balance for the plant is shown in Exhibit 3-113. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon leaves the plant as unburned carbon in the slag and as CO₂ in the stack gas, ASU vent gas, and the captured CO₂ product. The carbon capture efficiency is defined as one minus the amount of carbon in the stack gas relative to the total carbon in less carbon contained in the slag, represented by the following fraction:

$$\frac{\text{Carbon in Stack}}{(\text{Total Carbon In}) - (\text{Carbon in Slag})} = \left(1 - \left(\frac{30,752}{311,614 - 6,208}\right) * \right) 100 = 90.0\%$$

Exhibit 3-113 Case B5B-Q carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	140,805 (310,423)	Stack Gas	13,949 (30,752)
Air (CO ₂)	540 (1,191)	CO ₂ Product	124,473 (274,417)
		ASU Vent	107 (237)
		Slag	2,816 (6,208)
Total	141,346 (311,614)	Total	141,346 (311,614)

Exhibit 3-114 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur co-sequestered with the CO₂ product, and sulfur emitted in the stack gas. Sulfur in the slag is considered to be negligible.

Exhibit 3-114 Case B5B-Q sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,536 (12,206)	Stack Gas	3 (6)
		CO ₂ Product	10 (23)
		Elemental Sulfur	5,523 (12,177)
Total	5,536 (12,206)	Total	5,536 (12,206)

Exhibit 3-115 shows the overall water balance for the plant. The exhibit is presented in an identical manner as for cases B1A-B5B.

Exhibit 3-115 Case B5B-Q water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)	m ³ /min (gpm)
Slag Handling	0.53 (139)	0.53 (139)	–	–	–
Slurry Water	1.51 (400)	1.51 (400)	–	–	–
Quench/Wash	6.66 (1,759)	1.86 (493)	4.79 (1,267)	–	4.79 (1,267)
Humidifier	–	–	–	–	–
SWS Blowdown	–	–	–	0.04 (10.32)	-0.04 (-10.32)
Condenser Makeup	0.15 (40)	–	0.15 (40)	–	0.15 (40)
BFW Makeup	0.15 (40)	–	0.15 (40)	–	0.15 (40)
Gasifier Steam	–	–	–	–	–
Shift Steam	–	–	–	–	–
CT Steam Dilution	–	–	–	–	–
Cooling Tower	18.83 (4,974)	0.54 (143)	18.29 (4,831)	4.23 (1,119)	14.05 (3,712)
BFW Blowdown	–	0.15 (40)	-0.15 (-40)	–	-0.15 (-40)
SWS Blowdown	–	0.39 (103.15)	-0.39 (-103.15)	–	-0.39 (-103.15)
Total	27.68 (7,311)	4.45 (1,174)	23.23 (6,137)	4.27 (1,129)	18.96 (5,008)

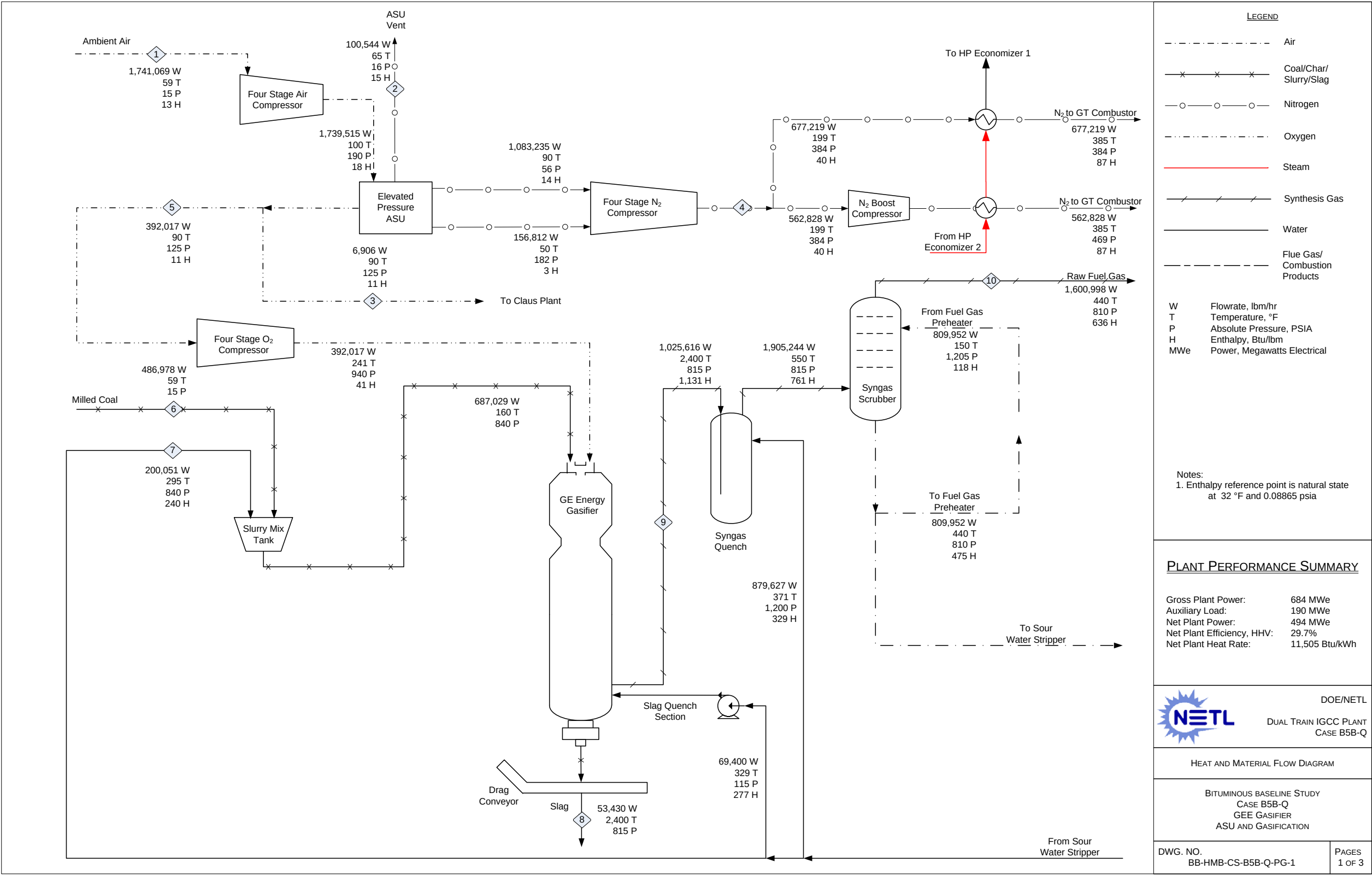
3.4.13.2 Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-116 through Exhibit 3-118:

- Coal gasification and ASU
- Syngas cleanup, sulfur recovery, and tail gas recycle
- Combined cycle power generation, steam, and FW

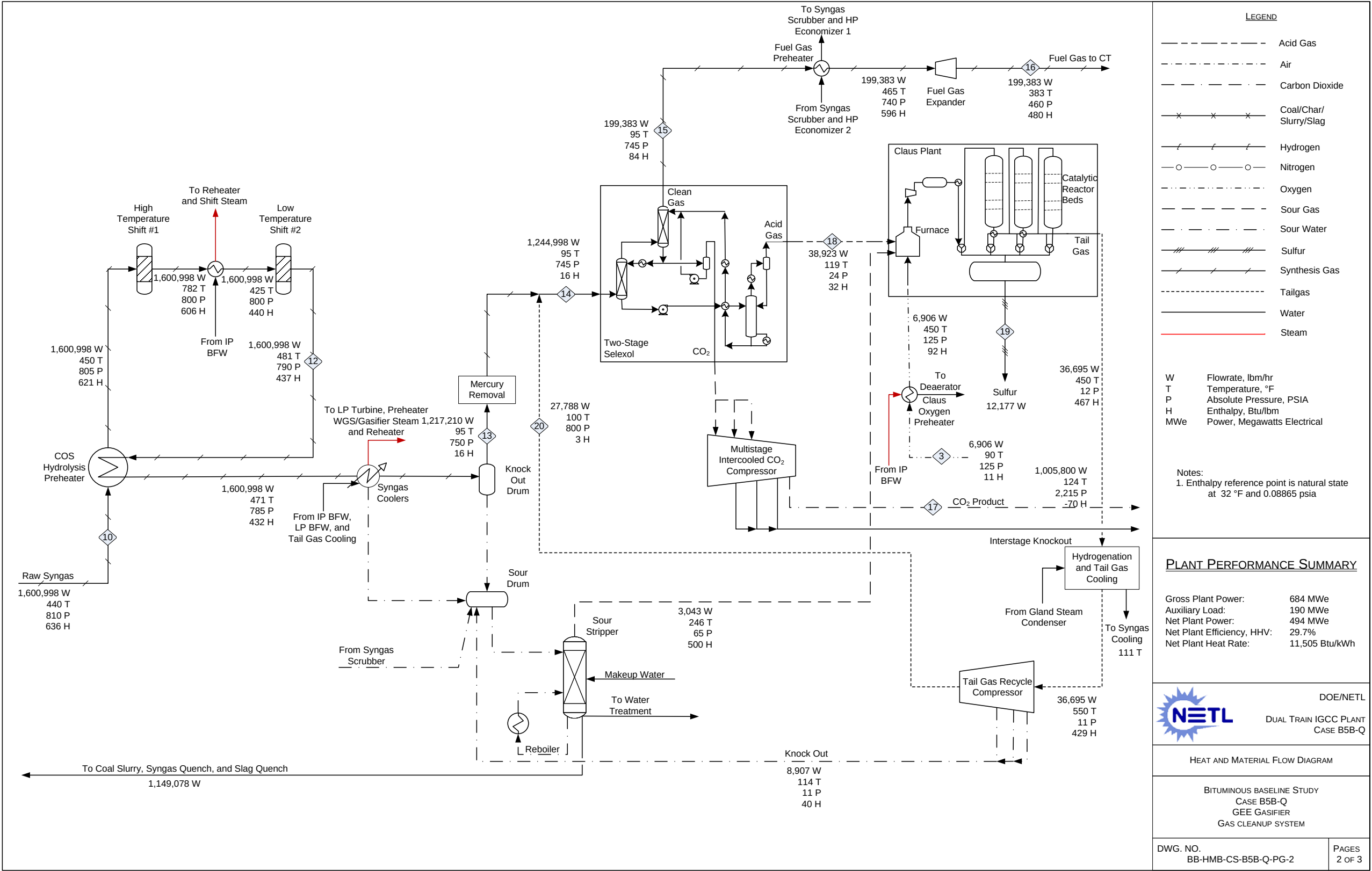
An overall plant energy balance is provided in tabular form in Exhibit 3-119. The power out is the combined CT, steam turbine, and expander power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-118) is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

Exhibit 3-116 Case B5B-Q coal gasification and ASU heat and mass balance schematic



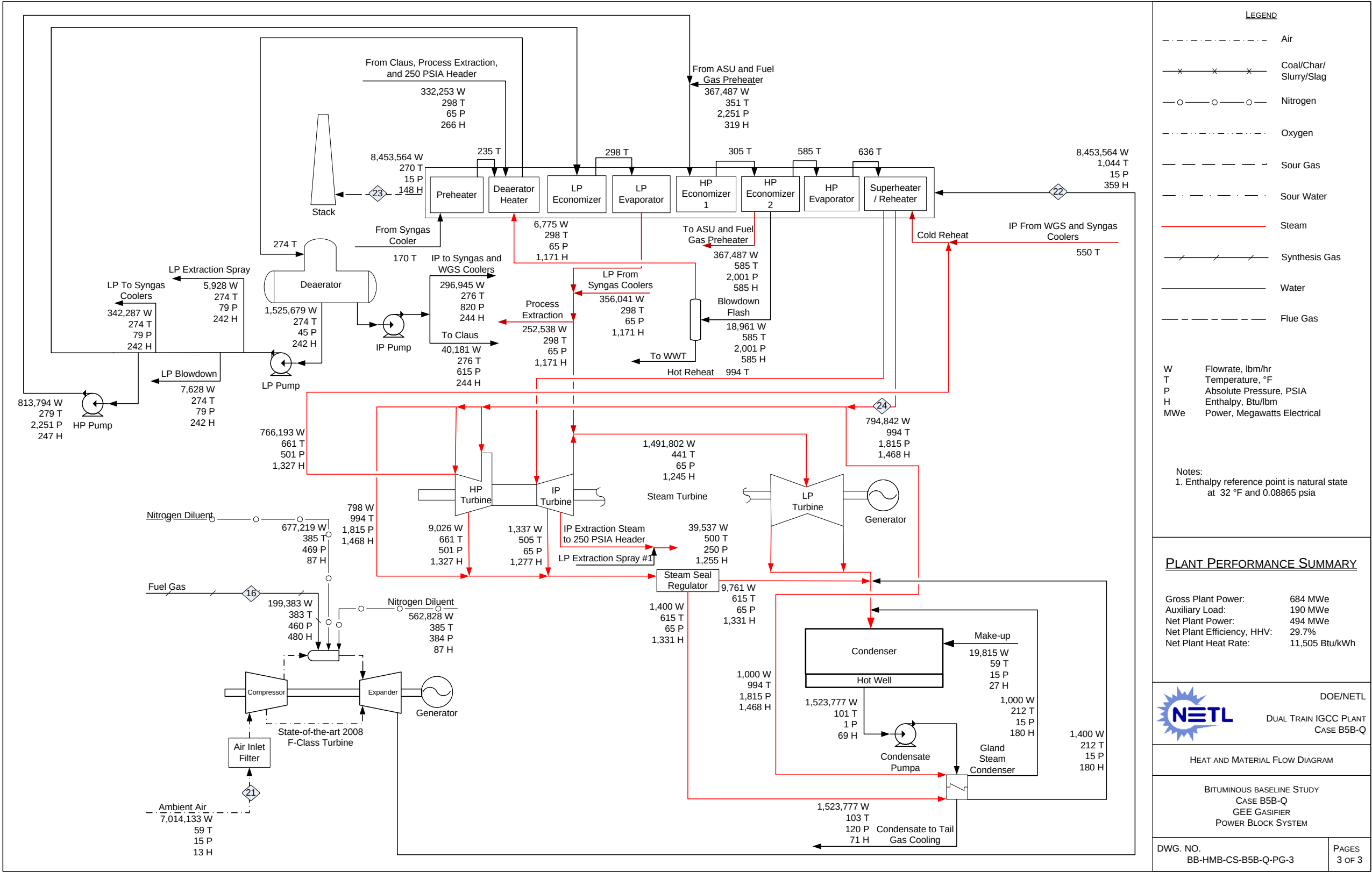
Source: NETL

Exhibit 3-117 Case B5B-Q syngas cleanup heat and mass balance schematic



Source: NETL

Exhibit 3-118 Case B5B-Q combined cycle power generation heat and mass balance schematic



Source: NETL

This page intentionally left blank

Exhibit 3-119 Case B5B-Q overall energy balance

	HHV	Sensible + Latent	Power	Total
Heat In (GJ/hr)				
Coal	5,994 (5,681)	5.0 (4.7)	–	5,999 (5,686)
Air	–	120.0 (113.8)	–	120.0 (113.8)
Raw Water Makeup	–	87.4 (82.8)	–	87.4 (82.8)
Auxiliary Power	–	–	685 (650)	685 (650)
TOTAL	5,994 (5,681)	212.4 (201.3)	685 (650)	6,892 (6,532)
Heat Out (GJ/hr)				
ASU Vent	–	1.6 (1.5)	–	1.6 (1.5)
Slag	92.3 (87.5)	37.8 (35.9)	–	130.2 (123.4)
Stack Gas	–	1,317 (1,248)	–	1,317 (1,248)
Sulfur	51.2 (48.5)	0.6 (0.6)	–	51.8 (49.1)
Motor Losses and Design Allowances	–	–	57.1 (54.2)	57.1 (54.2)
Condenser	–	1,452 (1,376)	–	1,452 (1,376)
Non-Condenser Cooling Tower Loads	–	830 (787)	–	830 (787)
CO ₂	–	-74.0 (-70.2)	–	-74.0 (-70.2)
Blowdown	–	50.0 (47.4)	–	50.0 (47.4)
CO ₂ Capture Losses	–	176.2 (167.0)	–	176.2 (167.0)
<i>Ambient Losses^A</i>	–	163.1 (154.6)	–	163.1 (154.6)
Power	–	–	2,463 (2,335)	2,463 (2,335)
TOTAL	143.5 (136.0)	3,955 (3,748)	2,520 (2,389)	6,618 (6,273)
Unaccounted Energy ^B	–	273 (259)	–	273 (259)

^AAmbient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers.

^BBy difference

3.4.14 Case B5B-Q - Major Equipment List

Major equipment items for the GEE quench-only gasifier with CO₂ capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 3.4.15. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

Case B5B-Q – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	50 tonne (50 ton)	2	1
7	Feeder	Vibratory	180 tonne/hr (200 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	180 tonne (200 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	360 tonne/hr (400 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	800 tonne (900 ton)	3	0

Case B5B-Q – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/h (90 tph)	3	0
2	Conveyor No. 6	Belt w/ tripper	240 tonne/h (270 tph)	1	0
3	Rod Mill Feed Hopper	Dual Outlet	490 tonne (540 ton)	1	0
4	Weigh Feeder	Belt	120 tonne/h (130 tph)	2	0
5	Rod Mill	Rotary	120 tonne/h (130 tph)	2	0
6	Slurry Water Storage Tank with Agitator	Field erected	299,880 liters (79,220 gal)	2	0
7	Slurry Water Pumps	Centrifugal	830 lpm (220 gpm)	2	1
8	Trommel Screen	Coarse	170 tonne/h (190 tph)	2	0
9	Rod Mill Discharge Tank with Agitator	Field erected	392,290 liters (103,630 gal)	2	0
10	Rod Mill Product Pumps	Centrifugal	3,300 lpm (900 gpm)	2	2
11	Slurry Storage Tank with Agitator	Field erected	1,176,900 liters (310,900 gal)	2	0
12	Slurry Recycle Pumps	Centrifugal	6,500 lpm (1,700 gpm)	2	2
13	Slurry Product Pumps	Positive displacement	3,300 lpm (900 gpm)	2	2

Case B5B-Q – Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	811,000 liters (214,000 gal)	2	0
2	Condensate Pumps	Vertical canned	6,380 lpm @ 90 m H ₂ O (1,690 gpm @ 300 ft H ₂ O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	465,000 kg/hr (1,025,000 lb/hr)	2	0
4	Intermediate-Pressure Feedwater Pump	Horizontal centrifugal, single stage	6,600 lpm @ 30 m H ₂ O (1,740 gpm @ 90 ft H ₂ O)	2	1
5	High-Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 3,520 lpm @ 1,800 m H ₂ O (930 gpm @ 6,100 ft H ₂ O)	2	1
6	High-Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,590 lpm @ 220 m H ₂ O (420 gpm @ 730 ft H ₂ O)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded Screw	28 m ³ /min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m ³ /min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	457 GJ/hr (433 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	163,800 lpm @ 20 m H ₂ O (43,300 gpm @ 70 ft H ₂ O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H ₂ O (1,000 gpm @ 350 ft H ₂ O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H ₂ O (700 gpm @ 250 ft H ₂ O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	6,030 lpm @ 20 m H ₂ O (1,590 gpm @ 60 ft H ₂ O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	3,010 lpm @ 270 m H ₂ O (800 gpm @ 880 ft H ₂ O)	4	1
16	Filtered Water Pumps	Stainless steel, single suction	4,170 lpm @ 50 m H ₂ O (1,100 gpm @ 160 ft H ₂ O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	2,003,000 liter (529,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	230 lpm (60 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0

Case B5B-Q – Account 4: Gasifier, ASU, and Accessories Including Low Temperature Heat Recovery and Fuel Gas Saturation

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized slurry-feed, entrained bed	2,900 tonne/day, 5.6 MPa (3,200 tpd, 815 psia)	2	0
2	Synthesis Gas Cooler	Vertical downflow radiant heat exchanger with outlet quench chamber	256,000 kg/hr (564,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	475,000 kg/hr (1,048,000 lb/hr) Design efficiency 90%	2	0
4	Raw Gas Coolers	Shell and tube with condensate drain	399,000 kg/hr (881,000 lb/hr)	8	0
5	Raw Gas Knockout Drum	Vertical with mist eliminator	304,000 kg/hr, 35°C, 5.2 MPa (671,000 lb/hr, 95°F, 755 psia)	2	0
6	Synthesis Gas Reheater	Shell and Tube	50,000 kg/hr (110,000 lb/hr)		
7	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	475,000 kg/hr (1,048,000 lb/hr) syngas	2	0
8	ASU Main Air Compressor	Centrifugal, multi-stage	6,000 m ³ /min @ 1.3 MPa (210,000 scfm @ 190 psia)	2	0
9	Cold Box	Vendor design	2,400 tonne/day (2,600 tpd) of 95% purity oxygen	2	0
10	Oxygen Compressor	Centrifugal, multi-stage	1,000 m ³ /min (42,000 scfm) Suction - 1.0 MPa (130 psia) Discharge - 6.5 MPa (940 psia)	2	0
11	Primary Nitrogen Compressor	Centrifugal, multi-stage	4,000 m ³ /min (134,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
12	Secondary Nitrogen Compressor	Centrifugal, single-stage	550 m ³ /min (19,000 scfm) Suction - 1.3 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0
13	Syngas Dilution Nitrogen Boost Compressor	Centrifugal, single-stage	2,000 m ³ /min (70,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 3.2 MPa (470 psia)	2	0

Case B5B-Q – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	304,000 kg/hr (669,000 lb/hr), 35°C (95°F), 5.2 MPa (750 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	304,000 kg/hr (669,000 lb/hr), 35°C (95°F), 5.2 MPa (750 psia)	2	0
3	Sulfur Plant	Claus type	146 tonne/day (161 tpd)	1	0
4	Water Gas Shift Reactors	Fixed bed, catalytic	200,000 kg/hr (440,000 lb/hr), 232°C (450°F), 5.5 MPa (800 psia)	4	0
5	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 154 GJ/hr (146 MMBtu/hr) Exchanger 2: 4 GJ/hr (4 MMBtu/hr)	4	0
6	Acid Gas Removal Plant	Two-stage Selexol	311,000 kg/hr (685,000 lb/hr), 35°C (95°F), 5.1 MPa (745 psia)	2	0
7	Hydrogenation Reactor	Fixed bed, catalytic	18,000 kg/hr (40,000 lb/hr), 232°C (450°F), 0.1 MPa (12.3 psia)	1	0
8	Tail Gas Recycle Compressor	Centrifugal	14,000 kg/hr (31,000 lb/hr) each	1	0
9	CO ₂ Compressor	Integrally geared, multi-stage centrifugal	1,100 m ³ /min @ 15.3 MPa (39,900 scfm @ 2,215 psia)	4	0

Case B5B-Q – Account 6: Combustion Turbine and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	State-of-the-art 2008 F-class	230 MW	2	0
2	Combustion Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0
3	Syngas Expansion Turbine/Generator	Turbo Expander	49,700 kg/h (109,700 lb/h) 5.1 MPa (740 psia) Inlet 3.2 MPa (460 psia) Outlet	2	0

Case B5B-Q – Account 7: HRSG, Ducting, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.5 m (28 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 198,294 kg/hr, 12.4 MPa/534°C (437,163 lb/hr, 1,800 psig/994°F)	2	0

Case B5B-Q – Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	225 MW 12.4 MPa/534°C/534°C (1,800 psig/994°F/994°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	250 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,600GJ/hr (1,510 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0

Case B5B-Q – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	485,000 lpm @ 30 m (128,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 2,700 GJ/hr (2,560 MMBtu/hr) heat duty	1	0

Case B5B-Q – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	254,000 liters (67,000 gal)	2	0
2	Slag Crusher	Roll	13 tonne/hr (15 tph)	2	0
3	Slag Depressurizer	Lock Hopper	13 tonne/hr (15 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	153,000 liters (40,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	69,000 liters (18,000 gal)	2	0
6	Slag Conveyor	Drag chain	13 tonne/hr (15 tph)	2	0
7	Slag Separation Screen	Vibrating	13 tonne/hr (15 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	13 tonne/hr (15 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	217,000 liters (57,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	60 lpm @ 14 m H ₂ O (10 gpm @ 46 ft H ₂ O)	2	2
11	Grey Water Storage Tank	Field erected	69,000 liters (18,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	250 lpm @ 560 m H ₂ O (60 gpm @ 1,850 ft H ₂ O)	2	2
13	Slag Storage Bin	Vertical, field erected	1,000 tonne (1,100 tons)	2	0
14	Unloading Equipment	Telescoping chute	110 tonne/hr (120 tph)	1	0

Case B5B-Q – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	24 kV/345 kV, 200 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 80 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 48 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 7 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case B5B-Q – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.4.15 Case B5B-Q - Cost Estimating

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-120 shows a detailed breakdown of the capital costs; Exhibit 3-121 shows the owner's costs, TOC, and TASC; Exhibit 3-122 shows the initial and annual O&M costs; and Exhibit 3-123 shows the COE breakdown.

The estimated TPC of the GEE gasifier with CO₂ capture in quench-only configuration is \$3,405/kW. Process contingency represents 4.0 percent of the TOC and project contingency represents 13.7 percent. The COE, including CO₂ T&S costs of \$10.2/MWh, is \$148.9/MWh. For comparison, the TPC and COE for Case B5B, GEE in radiant-only configuration with CO₂ capture, are \$3,387/kW and \$144.7/MWh.

Exhibit 3-120 Case B5B-Q total plant cost details

Case:		B5B-Q – GEE Quench IGCC w/ CO ₂					Estimate Type:			Conceptual	
Plant Size (MW,net):		494					Cost Base:			Jun 2011	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
1		Coal & Sorbent Handling									
1.1	Coal Receive & Unload	\$4,552	\$0	\$2,194	\$0	\$6,746	\$675	\$0	\$1,484	\$8,905	\$18
1.2	Coal Stackout & Reclaim	\$5,882	\$0	\$1,406	\$0	\$7,289	\$729	\$0	\$1,604	\$9,621	\$19
1.3	Coal Conveyors & Yd Crush	\$5,469	\$0	\$1,392	\$0	\$6,861	\$686	\$0	\$1,509	\$9,056	\$18
1.4	Other Coal Handling	\$1,431	\$0	\$322	\$0	\$1,753	\$175	\$0	\$386	\$2,314	\$5
1.9	Coal & Sorbent Hnd. Foundations	\$0	\$3,040	\$7,944	\$0	\$10,984	\$1,098	\$0	\$2,416	\$14,499	\$29
Subtotal		\$17,334	\$3,040	\$13,258	\$0	\$33,632	\$3,363	\$0	\$7,399	\$44,394	\$90
2		Coal & Sorbent Prep & Feed									
2.1	Coal Crushing & Drying	w/2.3	\$0	w/2.3	\$0	\$0	\$0	\$0	\$0	\$0	\$0
2.2	Prepared Coal Storage & Feed	\$1,944	\$467	\$301	\$0	\$2,712	\$271	\$0	\$597	\$3,580	\$7
2.3	Slurry Prep & Feed	\$23,378	\$0	\$14,694	\$0	\$38,072	\$3,807	\$1,904	\$8,757	\$52,539	\$106
2.4	Misc. Coal Prep & Feed	\$1,069	\$781	\$2,301	\$0	\$4,151	\$415	\$0	\$913	\$5,479	\$11
2.9	Coal & Sorbent Feed Foundation	\$0	\$3,922	\$3,365	\$0	\$7,287	\$729	\$0	\$1,603	\$9,618	\$19
Subtotal		\$26,391	\$5,170	\$20,660	\$0	\$52,221	\$5,222	\$1,904	\$11,869	\$71,217	\$144
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	\$2,281	\$3,936	\$2,064	\$0	\$8,280	\$828	\$0	\$1,822	\$10,930	\$22
3.2	Water Makeup & Pretreating	\$908	\$94	\$499	\$0	\$1,501	\$150	\$0	\$495	\$2,146	\$4
3.3	Other Feedwater Subsystems	\$1,282	\$424	\$379	\$0	\$2,085	\$208	\$0	\$459	\$2,752	\$6
3.4	Service Water Systems	\$531	\$1,059	\$3,650	\$0	\$5,240	\$524	\$0	\$1,729	\$7,493	\$15
3.5	Other Boiler Plant Systems	\$2,862	\$1,069	\$2,632	\$0	\$6,563	\$656	\$0	\$1,444	\$8,663	\$18
3.6	FO Supply Sys & Nat Gas	\$17,720	\$682	\$631	\$0	\$19,033	\$1,903	\$0	\$4,187	\$25,124	\$51
3.7	Waste Treatment Equipment	\$1,229	\$0	\$761	\$0	\$1,990	\$199	\$0	\$657	\$2,845	\$6
3.8	Misc. Power Plant Equipment	\$1,273	\$170	\$661	\$0	\$2,104	\$210	\$0	\$694	\$3,009	\$6
Subtotal		\$28,086	\$7,433	\$11,277	\$0	\$46,795	\$4,680	\$0	\$11,487	\$62,962	\$128
4		Gasifier & Accessories									
4.1	Quench Gasifier System	\$79,525	\$0	\$68,344	\$0	\$147,869	\$14,787	\$19,355	\$28,338	\$210,349	\$426
4.3	ASU & Oxidant Compression	\$231,253	\$0	w/equip.	\$0	\$231,253	\$23,125	\$0	\$25,438	\$279,816	\$567
4.4	Scrubber & LT Heat Recovery	\$8,555	\$0	\$0	\$0	\$8,555	\$855	\$0	\$1,882	\$11,292	\$23
4.5	Black Water & Sour Gas Section	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.8	Major Component Rigging	w/4.1&4.2	\$0	w/4.1&4.2	\$0	\$0	\$0	\$0	\$0	\$0	\$0
4.9	Gasification Foundations	\$0	\$3,060	\$2,670	\$0	\$5,730	\$573	\$0	\$1,576	\$7,879	\$16
Subtotal		\$319,333	\$3,060	\$71,014	\$0	\$393,407	\$39,341	\$19,355	\$57,233	\$509,336	\$1,031
5A		Gas Cleanup & Piping									
5A.1	Double Stage Selexol	\$160,990	\$0	w/equip.	\$0	\$160,990	\$16,099	\$32,198	\$41,857	\$251,145	\$509
5A.2	Elemental Sulfur Plant	\$12,451	\$2,427	\$15,954	\$0	\$30,831	\$3,083	\$0	\$6,783	\$40,697	\$82
5A.3	Mercury Removal	\$1,967	\$0	\$1,487	\$0	\$3,454	\$345	\$173	\$794	\$4,767	\$10
5A.4	Shift Reactors	\$12,108	\$0	\$4,841	\$0	\$16,949	\$1,695	\$0	\$3,729	\$22,373	\$45
5A.7	Fuel Gas Piping	\$0	\$810	\$530	\$0	\$1,340	\$134	\$0	\$295	\$1,769	\$4
5A.9	HGCU Foundations	\$0	\$733	\$494	\$0	\$1,227	\$123	\$0	\$405	\$1,755	\$4
Subtotal		\$187,516	\$3,970	\$23,306	\$0	\$214,792	\$21,479	\$32,371	\$53,863	\$322,505	\$653

Case:		B5B-Q – GEE Quench IGCC w/ CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		494					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
5B		CO ₂ Compression									
5B.2	CO ₂ Compression & Drying	\$40,213	\$6,032	\$17,419	\$0	\$63,663	\$6,366	\$0	\$14,006	\$84,035	\$170
Subtotal		\$40,213	\$6,032	\$17,419	\$0	\$63,663	\$6,366	\$0	\$14,006	\$84,035	\$170
6		Combustion Turbine & Accessories									
6.1	Combustion Turbine Generator	\$111,210	\$0	\$7,881	\$0	\$119,091	\$11,909	\$11,909	\$14,291	\$157,201	\$318
6.2	Syngas Expander	\$6,686	\$0	\$918	\$0	\$7,604	\$760	\$0	\$1,255	\$9,619	\$19
6.9	Combustion Turbine Foundations	\$0	\$1,016	\$1,175	\$0	\$2,191	\$219	\$0	\$723	\$3,133	\$6
Subtotal		\$117,896	\$1,016	\$9,974	\$0	\$128,886	\$12,889	\$11,909	\$16,269	\$169,953	\$344
7		HRSG, Ducting, & Stack									
7.1	Heat Recovery Steam Generator	\$29,570	\$0	\$5,726	\$0	\$35,296	\$3,530	\$0	\$3,883	\$42,708	\$86
7.3	Ductwork	\$0	\$2,109	\$1,478	\$0	\$3,587	\$359	\$0	\$789	\$4,734	\$10
7.4	Stack	\$4,069	\$0	\$1,518	\$0	\$5,587	\$559	\$0	\$615	\$6,761	\$14
7.9	HRSG, Duct & Stack Foundations	\$0	\$775	\$778	\$0	\$1,552	\$155	\$0	\$512	\$2,220	\$4
Subtotal		\$33,639	\$2,883	\$9,499	\$0	\$46,022	\$4,602	\$0	\$5,798	\$56,423	\$114
8		Steam Turbine Generator									
8.1	Steam TG & Accessories	\$35,245	\$0	\$4,866	\$0	\$40,111	\$4,011	\$0	\$4,412	\$48,535	\$98
8.2	Turbine Plant Auxiliaries	\$206	\$0	\$470	\$0	\$676	\$68	\$0	\$74	\$818	\$2
8.3	Condenser & Auxiliaries	\$3,200	\$0	\$1,812	\$0	\$5,012	\$501	\$0	\$551	\$6,064	\$12
8.4	Steam Piping	\$10,865	\$0	\$4,712	\$0	\$15,577	\$1,558	\$0	\$4,284	\$21,418	\$43
8.9	TG Foundations	\$0	\$951	\$1,680	\$0	\$2,630	\$263	\$0	\$868	\$3,761	\$8
Subtotal		\$49,516	\$951	\$13,540	\$0	\$64,006	\$6,401	\$0	\$10,190	\$80,597	\$163
9		Cooling Water System									
9.1	Cooling Towers	\$4,640	\$0	\$1,400	\$0	\$6,040	\$604	\$0	\$997	\$7,641	\$15
9.2	Circulating Water Pumps	\$2,236	\$0	\$170	\$0	\$2,405	\$241	\$0	\$397	\$3,043	\$6
9.3	Circ. Water System Auxiliaries	\$194	\$0	\$27	\$0	\$221	\$22	\$0	\$36	\$280	\$1
9.4	Circ. Water Piping	\$0	\$8,587	\$2,080	\$0	\$10,667	\$1,067	\$0	\$2,347	\$14,081	\$29
9.5	Make-up Water System	\$499	\$0	\$686	\$0	\$1,185	\$118	\$0	\$261	\$1,564	\$3
9.6	Component Cooling Water Sys	\$982	\$1,174	\$806	\$0	\$2,962	\$296	\$0	\$652	\$3,910	\$8
9.9	Circ. Water System Foundations	\$0	\$2,811	\$4,995	\$0	\$7,806	\$781	\$0	\$2,576	\$11,162	\$23
Subtotal		\$8,550	\$12,572	\$10,164	\$0	\$31,286	\$3,129	\$0	\$7,265	\$41,680	\$84
10		Ash & Spent Sorbent Handling Systems									
10.1	Slag Dewatering & Cooling	\$14,630	\$8,067	\$16,277	\$0	\$38,974	\$3,897	\$0	\$4,287	\$47,158	\$96
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
10.6	Ash Storage Silos	\$710	\$0	\$767	\$0	\$1,476	\$148	\$0	\$244	\$1,868	\$4
10.7	Ash Transport & Feed Equipment	\$977	\$0	\$228	\$0	\$1,205	\$120	\$0	\$199	\$1,524	\$3
10.8	Misc. Ash Handling Equipment	\$1,470	\$1,801	\$534	\$0	\$3,805	\$381	\$0	\$628	\$4,814	\$10
10.9	Ash/Spent Sorbent Foundation	\$0	\$60	\$78	\$0	\$138	\$14	\$0	\$46	\$197	\$0
Subtotal		\$17,786	\$9,928	\$17,884	\$0	\$45,598	\$4,560	\$0	\$5,403	\$55,561	\$113
11		Accessory Electric Plant									
11.1	Generator Equipment	\$1,106	\$0	\$1,077	\$0	\$2,183	\$218	\$0	\$240	\$2,641	\$5
11.2	Station Service Equipment	\$5,479	\$0	\$504	\$0	\$5,983	\$598	\$0	\$658	\$7,240	\$15

Case:		B5B-Q – GEE Quench IGCC w/ CO ₂					Estimate Type:		Conceptual		
Plant Size (MW,net):		494					Cost Base:		Jun 2011		
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/kW
11.3	Switchgear & Motor Control	\$10,113	\$0	\$1,880	\$0	\$11,993	\$1,199	\$0	\$1,979	\$15,171	\$31
11.4	Conduit & Cable Tray	\$0	\$5,139	\$15,840	\$0	\$20,978	\$2,098	\$0	\$5,769	\$28,845	\$58
11.5	Wire & Cable	\$0	\$9,907	\$6,028	\$0	\$15,935	\$1,594	\$0	\$4,382	\$21,911	\$44
11.6	Protective Equipment	\$0	\$806	\$2,995	\$0	\$3,801	\$380	\$0	\$627	\$4,808	\$10
11.7	Standby Equipment	\$265	\$0	\$264	\$0	\$529	\$53	\$0	\$87	\$669	\$1
11.8	Main Power Transformers	\$18,924	\$0	\$164	\$0	\$19,088	\$1,909	\$0	\$3,149	\$24,146	\$49
11.9	Electrical Foundations	\$0	\$172	\$468	\$0	\$639	\$64	\$0	\$211	\$914	\$2
Subtotal		\$35,887	\$16,024	\$29,218	\$0	\$81,130	\$8,113	\$0	\$17,103	\$106,346	\$215
12		Instrumentation & Control									
12.1	IGCC Control Equipment	w/4.1	\$0	w/4.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.2	Combustion Turbine Control	w/6.1	\$0	w/6.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.3	Steam Turbine Control	w/8.1	\$0	w/8.1	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.4	Other Major Component Control	\$1,271	\$0	\$866	\$0	\$2,138	\$214	\$107	\$369	\$2,827	\$6
12.5	Signal Processing Equipment	w/12.7	\$0	w/12.7	\$0	\$0	\$0	\$0	\$0	\$0	\$0
12.6	Control Boards, Panels & Racks	\$292	\$0	\$191	\$0	\$483	\$48	\$24	\$111	\$667	\$1
12.7	Computer & Accessories	\$6,782	\$0	\$222	\$0	\$7,004	\$700	\$350	\$805	\$8,860	\$18
12.8	Instrument Wiring & Tubing	\$0	\$2,611	\$4,942	\$0	\$7,553	\$755	\$378	\$2,171	\$10,857	\$22
12.9	Other I & C Equipment	\$4,534	\$0	\$2,246	\$0	\$6,780	\$678	\$339	\$1,170	\$8,966	\$18
Subtotal		\$12,879	\$2,611	\$8,467	\$0	\$23,957	\$2,396	\$1,198	\$4,626	\$32,177	\$65
13		Improvements to Site									
13.1	Site Preparation	\$0	\$119	\$2,706	\$0	\$2,826	\$283	\$0	\$932	\$4,041	\$8
13.2	Site Improvements	\$0	\$2,117	\$2,993	\$0	\$5,110	\$511	\$0	\$1,686	\$7,307	\$15
13.3	Site Facilities	\$3,794	\$0	\$4,259	\$0	\$8,053	\$805	\$0	\$2,657	\$11,516	\$23
Subtotal		\$3,794	\$2,236	\$9,958	\$0	\$15,988	\$1,599	\$0	\$5,276	\$22,864	\$46
14		Buildings & Structures									
14.1	Combustion Turbine Area	\$0	\$317	\$179	\$0	\$496	\$50	\$0	\$109	\$655	\$1
14.2	Steam Turbine Building	\$0	\$2,511	\$3,575	\$0	\$6,086	\$609	\$0	\$1,004	\$7,698	\$16
14.3	Administration Building	\$0	\$1,033	\$749	\$0	\$1,781	\$178	\$0	\$294	\$2,253	\$5
14.4	Circulation Water Pumphouse	\$0	\$194	\$103	\$0	\$296	\$30	\$0	\$49	\$375	\$1
14.5	Water Treatment Buildings	\$0	\$746	\$728	\$0	\$1,474	\$147	\$0	\$243	\$1,865	\$4
14.6	Machine Shop	\$0	\$529	\$361	\$0	\$890	\$89	\$0	\$147	\$1,126	\$2
14.7	Warehouse	\$0	\$853	\$551	\$0	\$1,404	\$140	\$0	\$232	\$1,776	\$4
14.8	Other Buildings & Structures	\$0	\$511	\$398	\$0	\$909	\$91	\$0	\$200	\$1,200	\$2
14.9	Waste Treating Building & Str.	\$0	\$1,143	\$2,182	\$0	\$3,325	\$332	\$0	\$731	\$4,389	\$9
Subtotal		\$0	\$7,836	\$8,825	\$0	\$16,661	\$1,666	\$0	\$3,009	\$21,337	\$43
Total		\$898,821	\$84,762	\$274,464	\$0	\$1,258,047	\$125,805	\$66,737	\$230,797	\$1,681,386	\$3,405

Exhibit 3-121 Case B5B-Q owner's costs

Description	\$/1,000	\$/kW
Pre-Production Costs		
6 Months All Labor	\$14,334	\$29
1 Month Maintenance Materials	\$3,191	\$6
1 Month Non-fuel Consumables	\$700	\$1
1 Month Waste Disposal	\$494	\$1
25% of 1 Months Fuel Cost at 100% CF	\$3,046	\$6
2% of TPC	\$33,628	\$68
Total	\$55,393	\$112
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	\$25,317	\$51
0.5% of TPC (spare parts)	\$8,407	\$17
Total	\$33,724	\$68
Other Costs		
Initial Cost for Catalyst and Chemicals	\$16,691	\$34
Land	\$900	\$2
Other Owner's Costs	\$252,208	\$511
Financing Costs	\$45,397	\$92
Total Overnight Costs (TOC)	\$2,085,698	\$4,224
TASC Multiplier (IOU, high-risk, 35 year)	1.140	
Total As-Spent Cost (TASC)	\$2,377,696	\$4,815

Exhibit 3-122 Case B5B-Q initial and annual operating and maintenance costs

Case:	B5B-Q – GEE Quench IGCC w/ CO ₂			Cost Base:	Jun 2011
Plant Size (MW _{net}):	494	Heat Rate-net (Btu/kWh):	11,505	Capacity Factor (%):	80
Operating & Maintenance Labor					
Operating Labor			Operating Labor Requirements per Shift		
Operating Labor Rate (base):	39.70	\$/hour	Skilled Operator:	2.0	
Operating Labor Burden:	30.00	% of base	Operator:	10.0	
Labor O-H Charge Rate:	25.00	% of labor	Foreman:	1.0	
			Lab Tech's, etc.:	3.0	
			Total:	16.0	
Fixed Operating Costs					
			Annual Cost		
			(\$)	(\$/kW-net)	
Annual Operating Labor:				\$7,233,658	\$14.649
Maintenance Labor:				\$15,699,956	\$31.795
Administrative & Support Labor:				\$5,733,403	\$11.611
Property Taxes and Insurance:				\$33,627,713	\$68.101
Total:				\$62,294,730	\$126.156
Variable Operating Costs					
			(\$)	(\$/MWh-net)	
Maintenance Material:				\$30,633,669	\$8.85243
Consumables					
	Consumption			Cost (\$)	
	Initial Fill	Per Day	Per Unit	Initial Fill	
Water (/1000 gallons):	0	4,419	\$1.67	\$0	\$2,159,895
Makeup and Waste Water Treatment Chemicals (lbs):	0	26,325	\$0.27	\$0	\$2,058,884
Carbon (Mercury Removal) (lb):	135,505	232	\$5.50	\$745,276	\$372,638
Shift Catalyst (ft³):	6,443	4.41	\$771.99	\$4,973,767	\$994,753
Selexol Solution (gal):	298,221	95	\$36.79	\$10,971,928	\$1,019,146
Claus Catalyst (ft³):	w/equip	2.01	\$203.15	\$0	\$119,480
Subtotal:				\$16,690,971	\$6,724,798
Waste Disposal					
Spent Mercury Catalyst (lb.):	0	232	\$0.65	\$0	\$44,045
Flyash (ton):	0	0	\$0.00	\$0	\$0
Slag (ton):	0	641	\$25.11	\$0	\$4,701,028
Subtotal:				\$0	\$4,745,073
By-Products					
Sulfur (tons):	0	146	\$0.00	\$0	\$0
Subtotal:				\$0	\$0
Variable Operating Costs Total:				\$16,690,971	\$42,103,540
Fuel Cost					
Illinois Number 6 (ton):	0	5,844	\$68.54	\$0	\$116,954,695
Total:				\$0	\$116,954,695

Exhibit 3-123 Case B5B-Q COE breakdown

Component	Value, \$/MWh	Percentage
Capital	74.7	54%
Fixed	18.0	13%
Variable	12.2	9%
Fuel	33.8	24%
Total (Excluding T&S)	138.7	N/A
CO ₂ T&S	10.2	7%
Total (Including T&S)	148.9	N/A

3.5 IGCC Case Summary

The performance results of the seven IGCC plant configurations modeled in this report are summarized in Exhibit 3-124.

Exhibit 3-124 Estimated performance and cost results for all cases

Case	Integrated Gasification Combined Cycle ^A						
	Shell		E-Gas™ FSQ		GEE R+Q		
	B1A	B1B	B4A	B4B	B5A	B5B	B5B-Q
PERFORMANCE							
Nominal CO ₂ Capture	0%	90%	0%	90%	0%	90%	90%
Capacity Factor	80%	80%	80%	80%	80%	80%	80%
Gross Power Output (MWe)	737	673	738	704	748	734	684
Auxiliary Power Requirement (MWe)	108	177	113	190	126	191	190
Net Power Output (MWe)	629	497	625	513	622	543	494
Coal Flow rate (lb/hr)	436,646	465,264	459,956	484,212	466,898	487,005	486,978
Natural Gas Flow rate (lb/hr)	N/A	N/A	N/A	N/A	N/A	N/A	N/A
HHV Thermal Input (kW _t)	1,492,878	1,590,722	1,572,575	1,655,503	1,596,309	1,665,056	1,664,962
Net Plant HHV Efficiency (%)	42.1%	31.2%	39.7%	31.0%	39.0%	32.6%	29.7%
Net Plant HHV Heat Rate (Btu/kWh)	8,100	10,927	8,585	11,002	8,758	10,459	11,505
Raw Water Withdrawal, gpm	4,150	5,652	4,382	5,757	4,755	5,834	6,137
Process Water Discharge, gpm	788	1,021	905	1,095	984	1,080	1,129
Raw Water Consumption, gpm	3,362	4,631	3,477	4,662	3,771	4,754	5,008
CO ₂ Capture Rate (%)	0	90	0	90	0	90	90
CO ₂ Emissions (lb/MMBtu)	197	20	199	20	197	20	20
CO ₂ Emissions (lb/MWh-gross)	1,361	161	1,448	158	1,434	152	165
CO ₂ Emissions (lb/MWh-net)	1,595	218	1,711	217	1,724	206	228
SO ₂ Emissions (lb/MMBtu)	0.004	0.002	0.012	0.002	0.001	0.002	0.002
SO ₂ Emissions (lb/MWh-gross)	0.029	0.017	0.085	0.017	0.009	0.017	0.018
NO _x Emissions (lb/MMBtu)	0.059	0.049	0.060	0.049	0.059	0.049	0.049
NO _x Emissions (lb/MWh-gross)	0.409	0.396	0.434	0.396	0.430	0.376	0.404
PM Emissions (lb/MMBtu)	0.007	0.007	0.007	0.007	0.007	0.007	0.007
PM Emissions (lb/MWh-gross)	0.049	0.057	0.052	0.057	0.052	0.055	0.059
Hg Emissions (lb/TBtu)	0.434	0.372	0.413	0.374	0.412	0.388	0.361
Hg Emissions (lb/MWh-gross)	3.00E-06	3.00E-06	3.00E-06	3.00E-06	3.00E-06	3.00E-06	3.00E-06
COST							
Total Plant Cost (2011\$/kW)	2,725	3,981	2,372	3,540	2,449	3,387	3,405
Bare Erected Cost	2,067	2,963	1,809	2,638	1,870	2,525	2,548
Home Office Expenses	207	296	181	264	187	253	255
Project Contingency	376	562	322	492	330	467	467
Process Contingency	75	160	61	147	61	143	135
Total Overnight Cost (2011\$/M)	2,107	2,442	1,830	2,252	1,888	2,279	2,086
Total Overnight Cost (2011\$/kW)	3,351	4,917	2,929	4,386	3,036	4,195	4,224
Owner's Costs	626	936	557	846	587	807	819
Total As-Spent Cost (2011\$/kW)	3,820	5,605	3,339	5,000	3,461	4,782	4,815
COE (\$/MWh) (excluding T&S)	107.0	152.6	99.8	141.9	102.6	135.4	138.7
Capital Costs	59.3	87.0	51.8	77.6	53.7	74.2	74.7
Fixed Costs	14.7	20.5	13.5	19.1	13.7	18.2	18.0
Variable Costs	9.3	13.0	9.2	12.8	9.4	12.2	12.2
Fuel Costs	23.8	32.1	25.2	32.3	25.7	30.7	33.8
COE (\$/MWh) (including T&S)	107.0	162.4	99.8	151.8	102.6	144.7	148.9
CO ₂ T&S Costs	0.0	9.8	0.0	9.9	0.0	9.2	10.2
CO ₂ Captured Cost (excluding T&S), \$/tonne ^B	N/A	78.9	N/A	66.5	N/A	63.2	61.0
CO ₂ Avoided Cost (including T&S), \$/tonne ^B	N/A	118.7	N/A	102.9	N/A	91.7	99.3

^AAll cases use high-risk financial assumptions consistent with NETL's "QGESS: Cost Estimation Methodology for NETL Assessments of Power Plant Performance." (2)

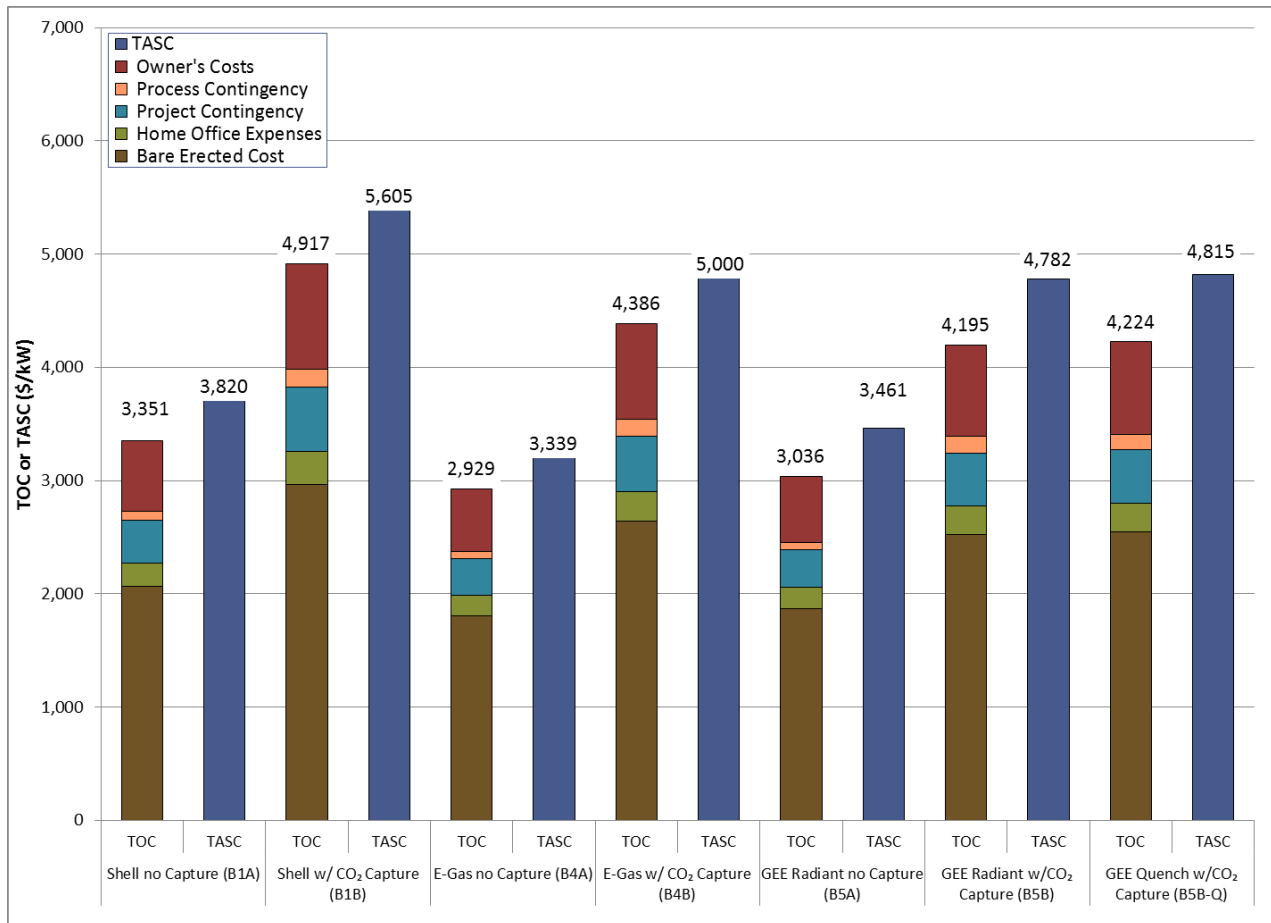
^BBoth the costs of CO₂ captured and avoided were calculated based on the non-capture supercritical pulverized coal (Case B12A) data provided in the "Cost and Performance Baseline for Fossil Energy Plants, Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity, Revision 3" report. (3) Case B12A has a COE of \$82.3/MWh, a CO₂ emission rate of 1,618 lb/MWh-gross, a gross plant output of 579,700 kW, and a net plant output of 550,012 kW.

The following observations can be made regarding plant performance:

- In the non-carbon capture cases the dry fed Shell gasifier has the highest net plant efficiency (42.1 percent), followed by the two-stage E-GasTM slurry fed gasifier (39.7 percent) and the single-stage slurry fed GEE gasifier (39.0 percent). The absolute values of the GEE and E-GasTM gasifiers are close to the reported values per the vendors. (16), (51) The Shell efficiency is slightly lower than reported by the vendor in other presentations. (48)
- The energy penalty associated with adding CO₂ capture is due to steam extraction for use in the SGS reaction, the auxiliary load for the CO₂ separation and compression equipment, and a slight derate of the gas turbine due to the higher moisture content of the flue gas. The reduction in net plant efficiency ranges from 6 to 11 percentage points with the variability being due to the optimum gasifier designs (e.g., slurry vs. dry feed, syngas quench vs. syngas heat recovery) which may vary between the capture and non-capture plant configurations.
- The lowest energy penalty (6 percentage points) corresponds to the GEE Radiant gasifier cases primarily due to the non-capture plant design (slurry feed, water quench) which results in a high moisture content in the syngas and thus a low addition of shift steam for SGS for the capture plant design.
- The highest energy penalty (11 percentage points) corresponds to the Shell gasifier cases. The design uses a dry feed system and, in the non-capture configuration, has relatively high heat recovery in the syngas cooler with no water quench, resulting in very low moisture content in the syngas. For the capture configuration, a water quench is added which increases the moisture content of the syngas for the SGS reaction but decreases the heat recovery in the syngas cooler.
- CB&I E-GasTM has the highest SO₂ emissions (0.085 lb/MWh-gross) of the seven cases because refrigerated MDEA has the lowest H₂S removal efficiency of the AGR technologies.

The components of TOC and the overall TASC of the seven cases are shown in Exhibit 3-125.

Exhibit 3-125 Plant capital cost for all cases



Source: NETL

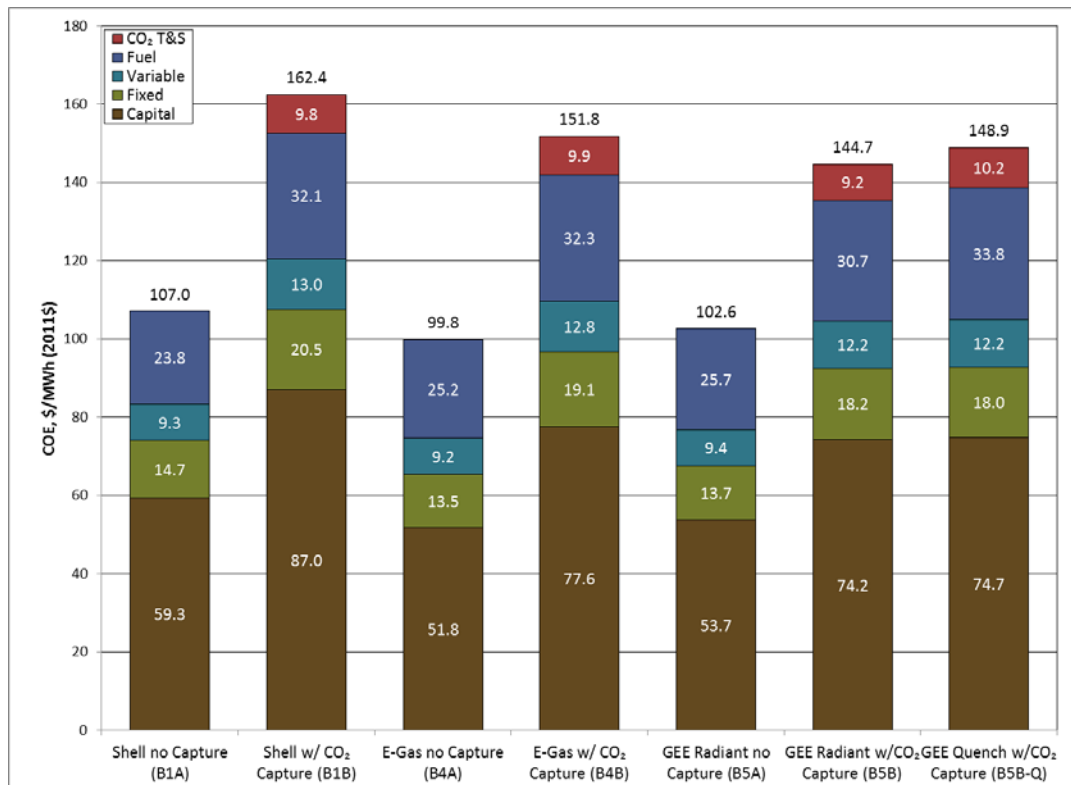
The following TOC observations are made with the caveat that the differences between cases are less than the estimate accuracy (-15 percent/+30 percent). However, all cases are evaluated using a common set of technical and economic assumptions allowing meaningful comparisons among the cases:

- E-GasTM has the lowest TOC cost among the non-capture cases. The E-GasTM technology has several features that lend it to being lower cost, such as:
 - The firetube syngas cooler is much smaller and less expensive than a radiant section. E-GasTM can use a firetube boiler because the two-stage design reduces the gas temperature (slurry quench) into a range where a radiant cooler is not needed.
 - The firetube syngas cooler sits next to the gasifier instead of above or below it, which reduces the height of the main gasifier structure. The E-GasTM proprietary slag removal system – used instead of lock hoppers below the gasifier – also contributes to the lower structure height.
- The normalized TOC of the GEE Radiant gasifier case is approximately 4 percent greater than E-GasTM, and Shell is approximately 14 percent higher.

- The GEE radiant gasifier (GEE quench is less than 1 percent greater than GEE radiant) is the low-cost technology in the CO₂ capture cases with E-GasTM normalized TOC approximately 5 percent higher and Shell approximately 17 percent higher.
- The ASU cost represents, on average, 12 percent of the TOC (range from 10-13 percent). The ASU cost includes oxygen and nitrogen compression, and in the non-capture cases, it also includes the cost of the CT extraction air heat exchanger. With nitrogen dilution used to the maximum extent possible, nitrogen compression costs are significant.
- The normalized TOC premium for adding CO₂ capture averages 43 percent (\$4,430/kW versus \$3,105/kW).

The COE is shown for all seven cases in Exhibit 3-126.

Exhibit 3-126 COE for all cases



Source: NETL

The following observations can be made:

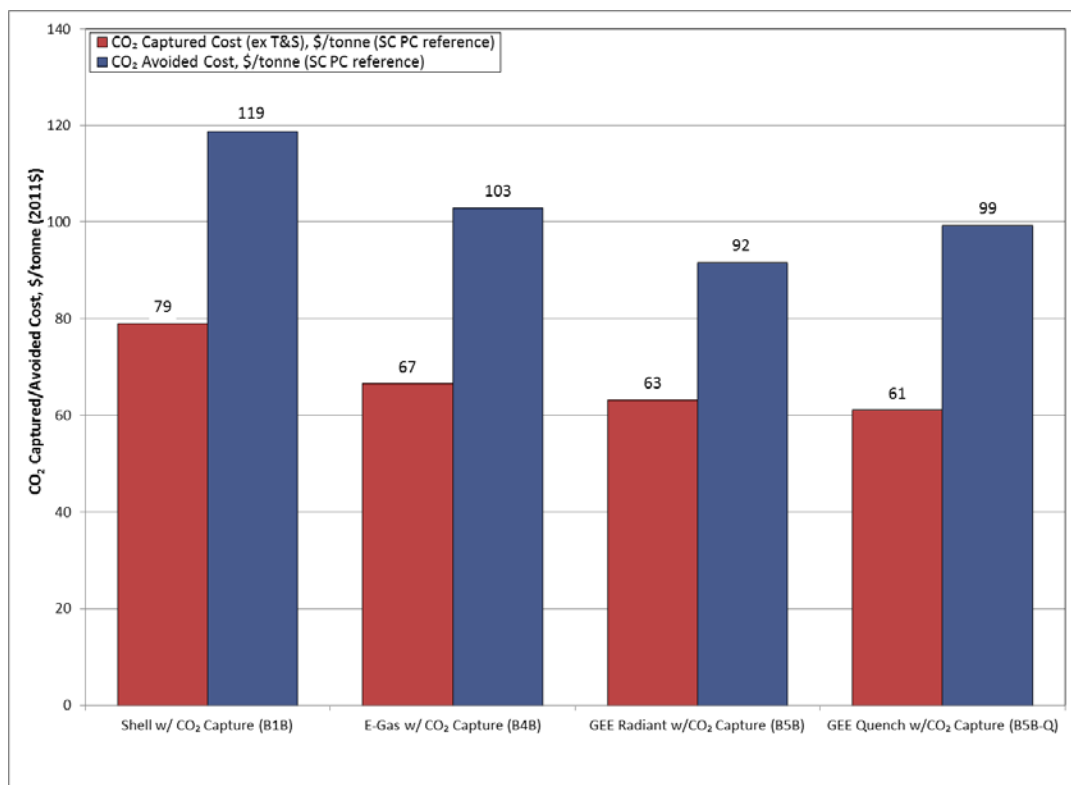
- The COE is dominated by capital costs, and is at least 50 percent of the total in all cases.
- In the non-capture cases the E-GasTM gasifier has the lowest COE, but the differential with Shell is reduced (relative to the normalized TOC comparison) primarily because of the higher efficiency of the Shell gasifier. The Shell COE is 7 percent higher than E-GasTM (compared to 14 percent higher normalized TOC). The GEE gasifier COE is about 3 percent higher than E-GasTM.
- In the capture cases the variation in COE is small; however, the order of the GEE and E-GasTM gasifiers is reversed. The range is from \$144.7/MWh for GEE Radiant to \$162.4/MWh for Shell with GEE Quench and E-GasTM intermediate at \$148.9/MWh and

\$151.8/MWh, respectively. The COE CO₂ capture premium averages 47 percent (range of 41 to 52 percent).

- The CO₂ T&S COE component composes less than 7 percent of the total COE in all capture cases.

As presented in Section 2.6.4, the CO₂ captured and avoided costs were calculated, and the results for the CO₂ capture cases are shown in Exhibit 3-127. The cost of captured CO₂ represents the minimum CO₂ plant gate sales price that will incentivize carbon capture in lieu of a defined reference non-capture plant. The cost of CO₂ avoided represents the minimum CO₂ emissions price, when applied to both the capture and non-capture plant that will incentivize carbon capture in lieu of a defined reference non-capture plant. Both the costs of CO₂ captured and avoided were calculated based on the non-capture supercritical pulverized coal (Case B12A) data provided in the “Cost and Performance Baseline for Fossil Energy Plants, Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity, Revision 3” report. (3) This case is representative of the lowest cost non-capture coal plant option. Case B12A has a COE of \$82.3/MWh, a CO₂ emission rate of 1,618 lb/MWh-gross, a gross plant output of 579,700 kW, and a net plant output of 550,012 kW.

Exhibit 3-127 Cost of CO₂ captured and avoided in all cases

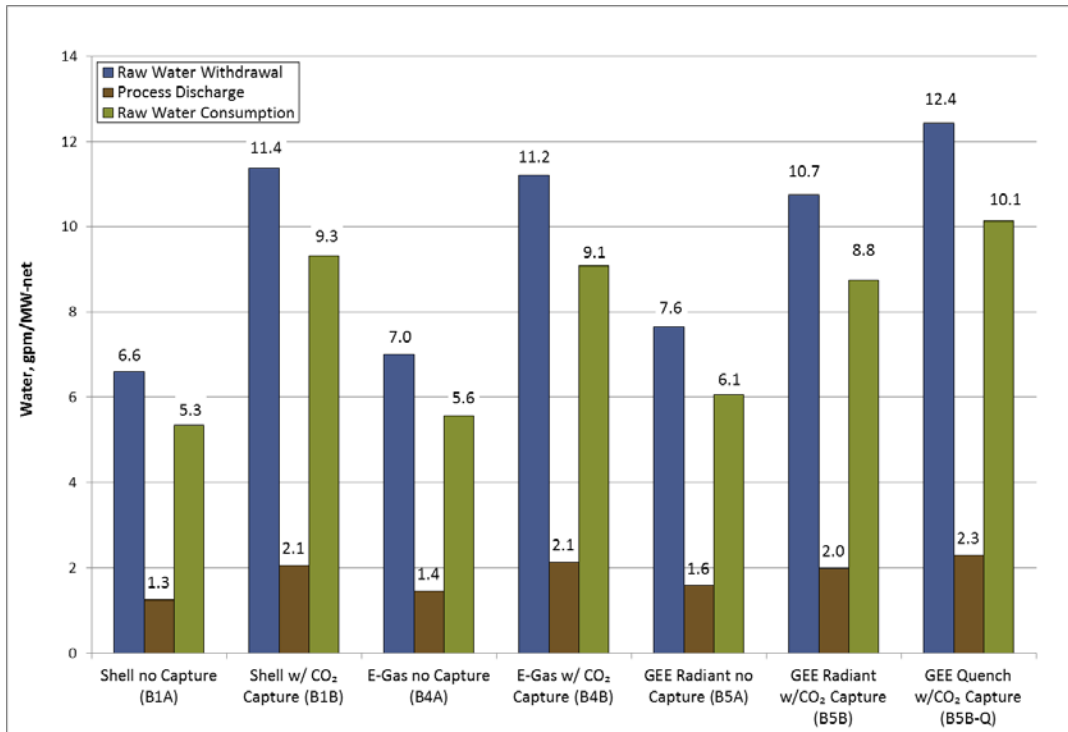


Source: NETL

The cost of CO₂ captured using SC PC as the non-capture reference case averages \$67/tonne (\$61/ton) with a range of \$61-\$79/tonne (\$55-\$72/ton). The cost of CO₂ avoided averages \$103/tonne (\$94/ton) with a range of \$92-\$119/tonne (\$83-\$108/ton).

The normalized water withdrawal, process discharge, and water consumption are presented in Exhibit 3-128.

Exhibit 3-128 Raw water withdrawal and consumption in all cases



Source: NETL

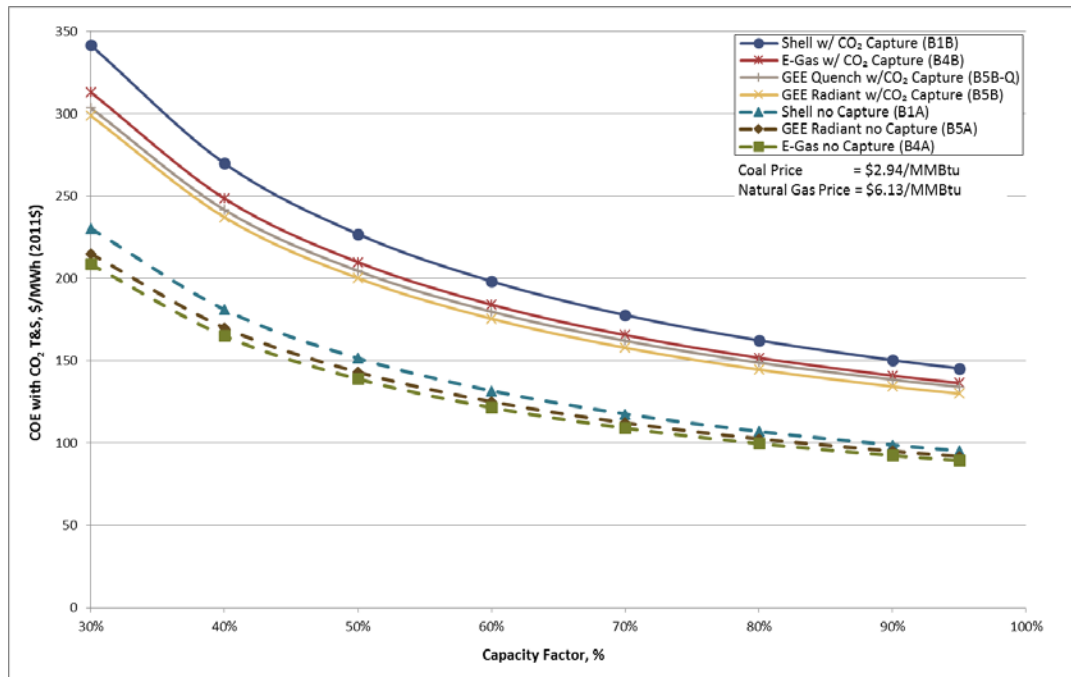
The following observations can be made:

- Raw water consumption for all cases is dominated by cooling tower makeup requirements, which account for 79-87 percent of raw water consumption in non-capture cases and 73-75 percent in CO₂ capture cases.
- Normalized water consumption for the GEE non-capture case is 9 percent higher than the E-GasTM non-capture case and 13 percent higher than the Shell non-capture case primarily because of the large quench water requirement.
- While both the normalized raw water consumption and withdrawal rates are 9 percent greater in the GEE Radiant non-capture case than in the E-GasTM case, the normalized raw water withdrawal and consumption rates are 6 percent and 4 percent greater, respectively, in the E-GasTM non-capture case than the Shell case. The discrepancy between withdrawal and consumption is because very little water is available to recover for internal recycle in the dry-fed Shell system
- The normalized raw water consumption for the four CO₂ capture cases varies by 9 percent from the highest to the lowest. The variation between cases is small because each technology requires approximately the same amount of water in the syngas prior to the shift reactors. The difference in technologies is where and how the water is introduced. Much of the water is introduced in the quench sections of the GEE and Shell cases while steam is added in the E-GasTM case.

3.5.1 Sensitivities

The effect of CF and coal price on COE is shown in Exhibit 3-129 and Exhibit 3-130, respectively. The assumption implicit in Exhibit 3-129 is that each gasifier technology can achieve a CF of up to 95 percent with no additional capital equipment cost.

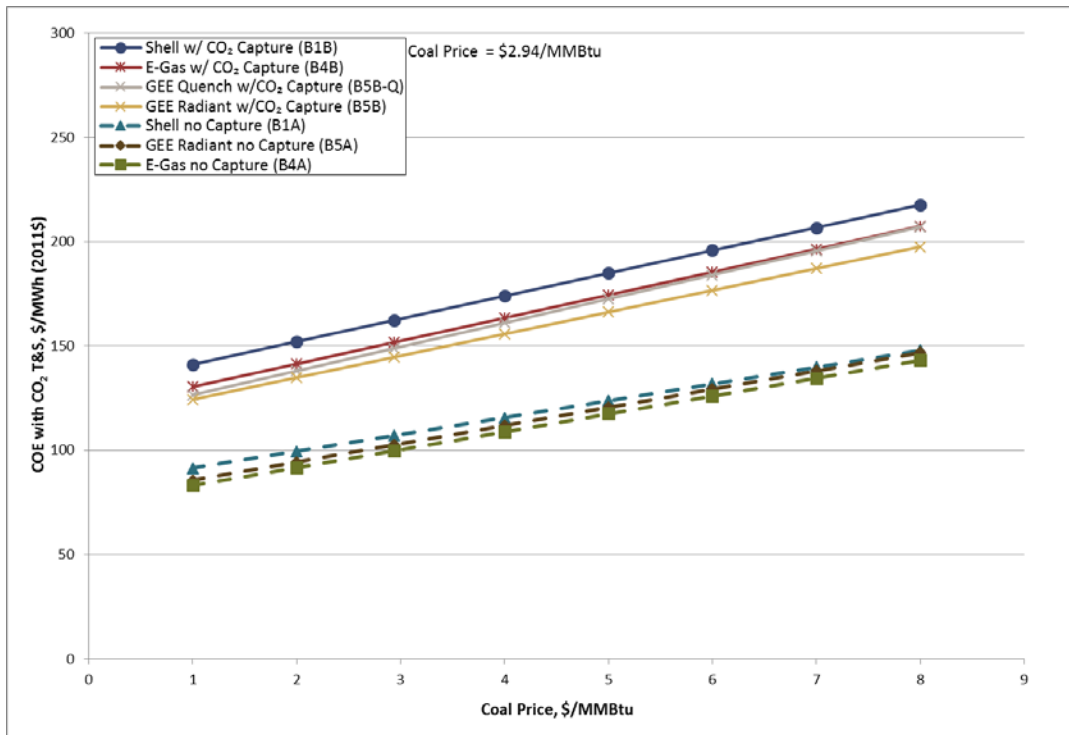
Exhibit 3-129 Capacity factor sensitivity of all cases



Source: NETL

The cost differential between technologies decreases as CF increases. At low CF the capital cost differential is more magnified and the spread between technologies increases slightly.

Exhibit 3-130 Sensitivity of COE to coal price for all cases

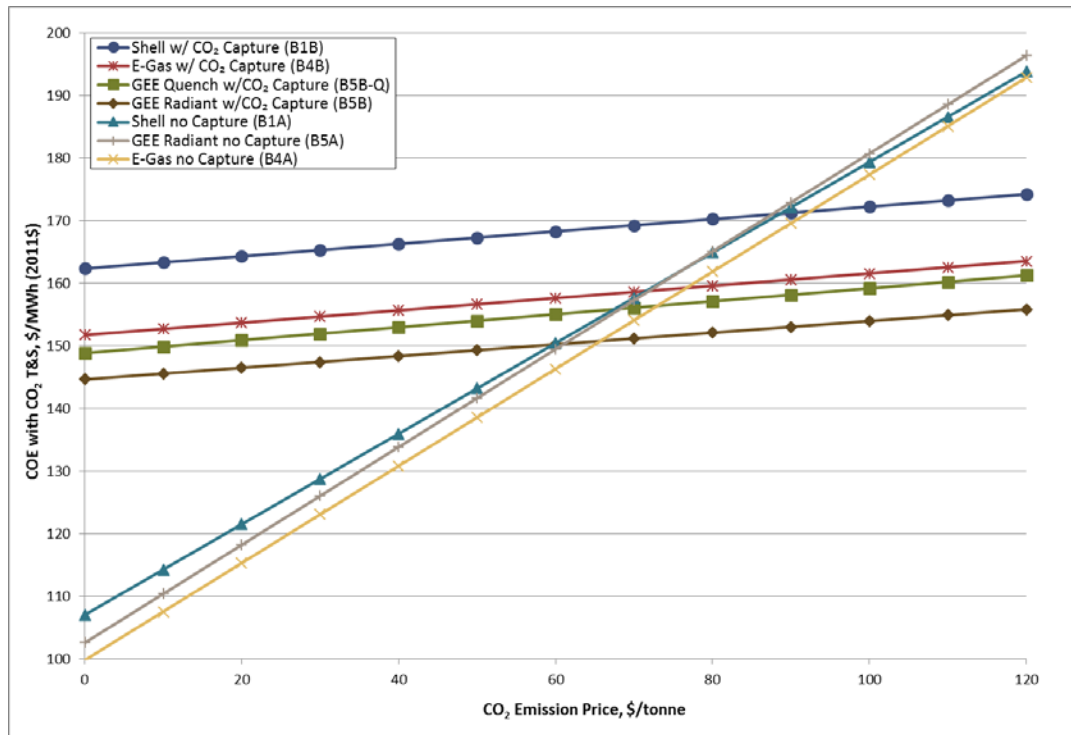


Source: NETL

COE is relatively insensitive to fuel costs; a tripling of coal price from \$1-\$3/MMBtu results in an average COE increase of about 17-20 percent for all cases.

In the event that future legislation assigns a cost to carbon emissions, all of the technologies examined in this report will become more expensive. The technologies without carbon capture will be impacted to a larger extent than those with carbon capture.

The curves in Exhibit 3-131 represent the study design conditions (capacity factor) and fuel price used; namely an 80 percent capacity factor and \$2.78/GJ (\$2.94/MMBtu) for coal.

Exhibit 3-131 Impact of carbon emissions price on study technologies

Source: NETL

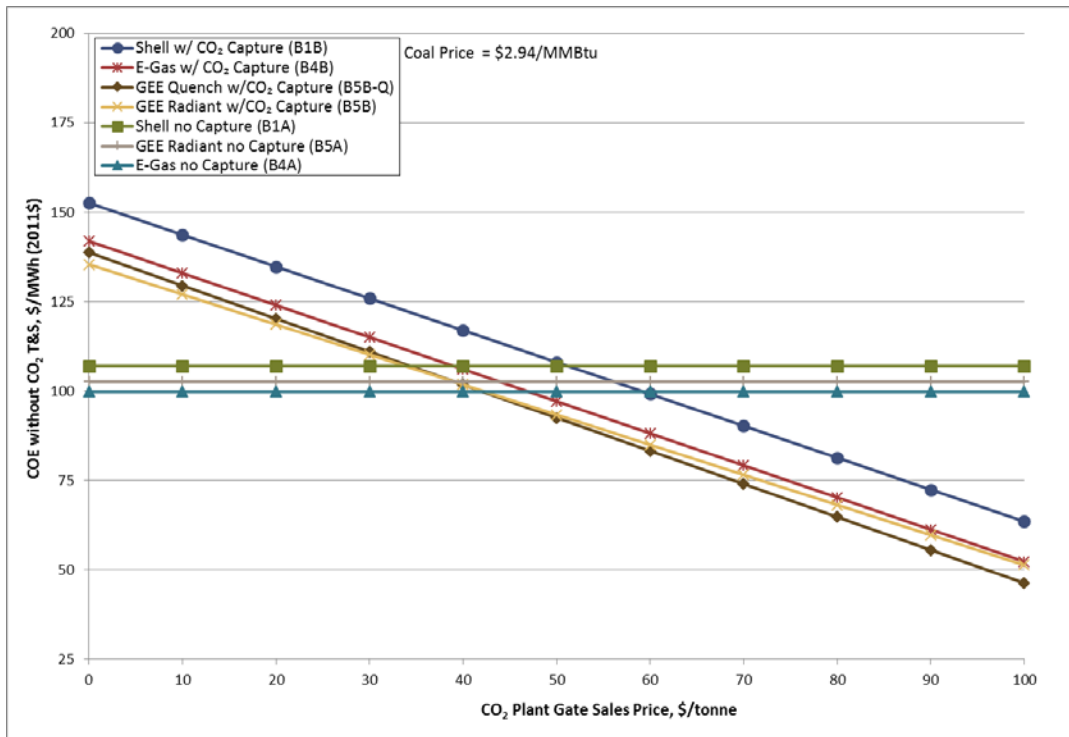
The following conclusions can be drawn from the carbon emissions price graph:

- The impact of an increasing CO₂ emission price has a significantly greater impact on the COE of the non-capture plants than the CO₂ capture plants.
- At \$60/tonne, the COEs of both the Shell and GEE Radiant cases without CO₂ capture are nearly equal to the COE of the GEE Radiant case with CO₂ capture.

Sale of the captured CO₂ for utilization and storage in CO₂ EOR has the potential to provide a revenue stream to the IGCC capture plant configurations. The plant gate CO₂ sales price will ultimately depend on a number of factors including plant location and crude oil prices. The cost of CO₂ captured represents the minimum CO₂ plant gate sales price that will incentivize carbon capture in lieu of a defined reference non-capture plant.

The curves in Exhibit 3-132 represent the study design conditions (capacity factor) and fuel price used; namely an 80 percent capacity factor and \$2.78/GJ (\$2.94/MMBtu) for coal.

Exhibit 3-132 Impact of carbon sales price on study technologies



Source: NETL

The following conclusions can be drawn from the carbon sales price graph:

- At \$51.1/tonne, the COEs of the Shell CO₂ capture and non-capture cases are equal.
- At \$47.0/tonne, the COEs of the E-GasTM CO₂ capture and non-capture cases are equal.
- At \$39.1/tonne, the COEs of both the GEE Radiant and GEE Quench CO₂ capture cases are equal to the GEE Radiant only non-capture case.

4 Revision Control

The initial issue of this report was published in May 2007, and an updated revision was published in November 2010. Subsequent to the reissue date, updates have been made to various report sections. These modifications were made for clarification and aesthetic purposes as well as to bring all costs to 2011 year dollar basis.

Exhibit 4-1 contains information that was added, changed, or deleted in successive revisions.

Exhibit 4-1 Record of revisions

Revision Number	Revision Date	Description of Change	Comments
1	8/23/07	Added disclaimer to Executive Summary and Introduction	Disclaimer involves clarification on extent of participation of technology vendors.
		Removed reference to cases 7 and 8 in Exhibits ES-1 and 1-1 since they no longer exist.	SNG cases moved to Volume 2 of this report as explained in the Executive Summary and Section 1.
		Added Section 2.8	Explains differences in IGCC TPC estimates in this report versus costs reported by other sources.
		Added Exhibit ES-14	Mercury emissions are now shown in a separate exhibit from SO ₂ , NO _x , and PM because of the different y-axis scale.
		Corrected PC and NGCC CO ₂ capture case water balances	The capture process cooling water requirement for the PC and NGCC CO ₂ capture cases was overstated and has been revised.
		Replaced Exhibits ES-4, 3-121, 4-52, and 5-30	The old water usage figures were in gpm (absolute) and in the new figures the water numbers are normalized by net plant output.
		Updated Selexol process description	Text was added to Section 3.1.5 to describe how H ₂ slip was handled in the models.
		Revised PC and NGCC CO ₂ capture case energy balances (Exhibits 4-21, 4-42 and 5-21)	The earlier version of the energy balances improperly accounted for the capture process heat losses. The heat removed from the capture process is rejected to the cooling tower.
		Corrected Exhibit 4-13 and Exhibit 4-27.	Sensible heat for combustion air in the two NGCC cases was for only one of the two combustion turbines – corrected to account for both turbines
2	10/27/10	Updated circulating water flow rate values in Section 3.1.8.	Revision 1 changes to capture system cooling water flow rate were not made in the text in Section 3.1.8 (Circulating Water System).
		Added Supplemental Chapter 6 “Effect of Higher Natural Gas Prices and Dispatch-Based Capacity Factors”	N/A
		Added Supplemental Chapter 7 “Dry and Parallel Cooling”	N/A

Revision Number	Revision Date	Description of Change	Comments
		Added Supplemental Chapter 8 "GEE IGCC in Quench-Only Configuration with CO ₂ Capture"	N/A
		Added Supplemental Chapter 9 "Sensitivity to MEA System Performance and Cost Bituminous Baseline Case B12BA"	N/A
		Updated Aspen models	Major Aspen model updates included: • Converting FORTRAN code-based steam cycles to Aspen blocks • Using the Peng-Robinson property method in the Aspen gasifier section • Modifying the AGR used in the IGCC cases to more closely represent commercially available technology • Increasing the capture efficiency of the E-Gas™ plant with capture to achieve 90 percent • Correcting a steam condition error in the supercritical PC cases with capture
		Updated case performance results	Major updates included: • Revising the water balances to include withdrawal and consumption • CAD-based HMB diagrams were replaced with Visio versions
		Completed updating case economic results	Major updates included: • Adding owner's costs to the total plant costs to generate total overnight cost • Updating fuel costs • Revising the T&S methodology to include the July, 2007 Handy-Whitman Index, pore space acquisition costs, and liability costs • Re-costing of cases based on the updated performance results • Switching to COE as the primary cost metric (as opposed to levelized COE)
		Updated report tables, figures, and text to reflect the revision 2 changes	N/A
2a	9/19/2013	Section 2.7.1 was revised to clarify the text that explains the level of technology maturity reflected in the plant level cost estimates.	N/A

Revision Number	Revision Date	Description of Change	Comments
2b	7/13/2015	Volume 1 has been split into two sub volumes.	Major updates included: • IGCC cases are reported in Volume 1b with a cost-only update (issued as an update to revision 2a) • PC and NGCC cases are reported in Volume 1a with a cost and performance update (issued as revision 3) • Executive summary significantly revised and shortened • Results analysis section added
		Separated Supplemental Chapter 6 “Effect of Higher Natural Gas Prices and Dispatch-Based Capacity Factors” into a stand-alone report.	N/A
		Separated Supplemental Chapter 7 “Impact of Dry and Parallel Cooling Systems on Cost and Performance of Fossil Fuel Power Plants” into a stand-alone report.	N/A
		Incorporated the Supplemental Chapter 8 “GEE IGCC in Quench-Only Configuration with CO ₂ Capture” into the body of the report.	The supplemental chapter was broken down and the information regarding the case describe within it is presented similar to cases B5A and B5B in Section 3.4.12.
		Removed Supplemental Chapter 9 “Sensitivity to MEA System Performance and Cost Bituminous Baseline Case 12A”	N/A
		Updated the environmental targets to current limits published by the EPA and presented in Section 2.3	MATS and NSPS regulate SO ₂ , NO _x , Filterable PM, Hg, and HCl on a lb/MWh-gross basis.
		Updated Section 2.4 covering Capacity Factors	Additional information has been included that supports the assumptions made regarding the capacity factors used for each technology type.
		Removed portions of Section 2.6 concerning cost estimating methodology	Many QGESS documents have been published that detail information generic to a number of studies published by NETL. In an effort to reduce the size of this report, text provided in these QGESS documents has been removed and references have been inserted that provide the QGESS document title and revision notation.
		Added cost of CO ₂ Captured methodology and results.	The Cost of CO ₂ avoided methodology has been moved from the executive summary and combined with the Cost of CO ₂ Captured methodology in Section 2.6.4.
		Removed Section 2.7	N/A

Revision Number	Revision Date	Description of Change	Comments
		Updated Section 3.1.4 to reflect the use of a dual carbon bed.	N/A
		Updated Section 3.1.5	Superfluous information has been removed and the remaining information has been re-organized, especially concerning AGR/Gasifier pairings used in non-capture cases
		Improved the BFD depiction of the HRSGs	N/A
		Updated all cases to a new case naming convention.	Ex. Revision 2's Case 1 is now Case B5A
		Updated performance tables	Major updates include: - Table is split into two sections - Performance summary - Plant power and auxiliary load breakdown - O₂:Coal ratio - Cold gas efficiency - Combustion turbine efficiency - Steam turbine efficiency and heat rate - LHV basis efficiency and heat rate
		Updated case performance results	Major updates included: - Added particle concentration to emissions results - Updated Energy Balance tables by adding Motor Losses and Design Allowances, Non-Condenser cooling tower loads, and ambient losses
		Completed updating case economic results	Major updates included: - Updated to 2011 year dollars - Revised the engineering and construction management costs - Added NG supply line - Updated the T&S costing methodology - Updated fuel prices
		Updated report tables, figures, and text to reflect the Revision 2b changes	N/A

5 References

1. **National Energy Technology Laboratory.** *Quality Guidelines for Energy Systems Studies: Fuel Prices for Selected Feedstocks in NETL Studies.* Office of Program Planning and Analysis. Pittsburgh, Pa : U. S. Department of Energy, 2012. DOE/NETL-3041/11212.
2. —. *Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessments of Power Plant Performance.* Pittsburgh : Department of Energy, 2011.
3. —. *Cost and Performance Baseline for Fossil Energy Plants Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity Revision 3.* Pittsburgh : Department of Energy, 2015. 1723.
4. **AACE International.** *Conducting Technical and Economic Evaluations – As Applied for the Process and Utility Industries; TCM Framework: 3.2 – Asset Planning, 3.3 Investment Decision Making.* s.l. : AACE International, 2003. Recommended Practice 16R-90.
5. —. *Cost Estimate Classification System – As Applied in Engineering, Procurement, and Construction for the Process Industries; TCM Framework 7.3 – Cost Estimating and Budgeting.* s.l. : AACE International, 2005. Recommended Practice No. 18R-97.
6. **National Energy Technology Laboratory.** *QGESS: Cost Estimation Methodology for NETL Assessments of Power Plant Performance.* Pittsburgh, Pa : U.S. Department of Energy, April 2011. DOE/NETL-2011/145.
7. —. *Quality Guidelines for Energy System Studies: Detailed Coal Specifications.* Office of Program Planning and Analysis. Pittsburgh, Pa : U. S. Department of Energy, 2013. DOE/NETL-401/012111.
8. —. *Quality Guidelines for Energy System Studies Specification for Selected Feedstocks.* Office of Program Planning and Analysis. Pittsburgh, Pa : U. S. Department of Energy, 2013. DOE/NETL-341/011812.
9. **Mining Media.** 2004 Keystone Coal Industry Manual. Prairieville, LA : Mining Media publications, 2004.
10. **National Energy Technology Laboratory.** *Major Environmental Aspects of Gasification-Based Power Generation Technologies.* Pittsburgh : Department of Energy, 2002.
11. **U.S. Environmental Protection Agency.** Regulatory Actions: Final Mercury and Air Toxics Standards (MATS) Reconsideration. *EPA.gov*. [Online] 2012. [Cited: February 6, 2012.] <http://www.epa.gov/mats/actions.html>. 2012.
12. —. Fact Sheet: Mercury and Air Toxics Standards for Power Plants. *EPA.gov*. [Online] 2012. [Cited: February 6, 2012.] <http://www.epa.gov/mats/pdfs/20111221MATSummaryfs.pdf>.
13. —. National Emission Standards for HAPs from Coal and Oil-Fired EGUs and Standards of Performance for Fossil-Fuel-Fired Electric Utility, Industrial-Commercial-Institutional, and Small Industrial-Commercial-Institutional Steam Generating. *GPO.gov*. [Online] 2012. [Cited: February 16, 2012.] <http://www.gpo.gov/fdsys/pkg/FR-2012-02-16/pdf/2012-806.pdf>.
14. —. Map of Counties Designated "Nonattainment" for Clean Air Act's National Ambient Air Quality Standards (NAAQS). *Image*. Washington DC : s.n., 2013.

15. —. Electronic Code of Federal Regulations. *U.S. Government Printing Office*. [Online] 2013. [Cited: February 6, 2013.] http://www.ecfr.gov/cgi-bin/text-idx?c=ecfr&tpl=/ecfrbrowse/Title40/40cfr63_main_02.tpl. 40 CFR63, Subpart UUUUU.
16. **Thomas, G.** Delivering the IGCC Solution. s.l. : GE Energy, January 2006.
17. **Morehead, H. and F. Hannemann.** Siemens Technology Improvements Enhance IGCC Plant Economics. San Francisco, CA : Gasification Technologies, October 2005.
18. **Denton, David.** Telephone communication. s.l. : Eastman Chemical Co.
19. **Phillips, J.** IGCC 101. *Gasification Technologies Regulators' Workshop*. Tampa, FL : s.n., 2011.
20. **Higman, Christopher, DellaVilla, Sal, and Steele, Bob.** The Reliability of Integrated Gasification Combined Cycle (IGCC) Power Generation Units. *Achema 2006*. May 2006.
21. **Amick, Phil.** ConocoPhillips Presentation at Gasification Technology Council Workshop for Environmental and Economic Regulators. *Gasification Technologies Regulators' Workshop*. June 2010.
22. **Electric Power Research Institute.** Evaluation of Alternative IGCC Plant Designs for High Availability and Near Zero Emissions: RAM Analysis and Impact of SCR. Palo Alto, CA : s.n., December 2005. 1010461.
23. —. Integrated Gasification Combined Cycle (IGCC) Design Considerations for High Availability, Volume 1: lessons from Existing Operations. Palo Alto, CA : s.n., 2005. 1012226.
24. —. Integrated Gasification Combined Cycle (IGCC) Design Considerations for High Availability, Volume 2: RAM Modeling of Standard Designs. Palo Alto, CA : s.n., 2007. 1014871.
25. —. Gasification Technology Status. December 2012. 1023864.
26. **The Marley Cooling Tower Company.** *Cooling Tower Fundamentals*. [ed.] John C. Hensley. Mission, Kansas : s.n., 1985. Vol. 2nd Edition.
27. **Chemical Engineering.** Chemical Engineering Plant Cost Index. [ed.] Access Intelligence LLC. *Chemical Engineering*. 2013.
28. **Federal Reserve Bank of St. Louis.** Gross Domestic Product: Chain-type Price Index (GDPCTPI). *FRED Economic Data*. [Online] 2013. [Cited: November 20, 2013.] <http://research.stlouisfed.org/fred2/series/GDPCTPI>. BEA Account Code: B191RG3.
29. **National Energy Technology Laboratory.** *Quality Guidelines for Energy System Studies: Carbon Dioxide Transport and Storage Costs in NETL Studies*. Pittsburgh, PA : Department of Energy, 2014. DOE/NETL-2013/1614.
30. **Universal Industrial Gases.** Overview of Cryogenic Air Separation. [Online] n. d. [Cited: August 29, 2013.] <http://www.uigi.com/cryodist.html>.
31. **Geosits, Robert F. and Schmoe, Lee A.** *IGCC – The Challenges of Integration, Proceedings of GT2005 ASME Turbo Expo 2005: Power for Land, Sea, and Air.*, Reno, NV : s.n., June 2005.

32. **Jaeger, Harry.** Plant Design Net Rated 644 MW and 38% HHV on Low Rank Coal. *Gas Turbine World*. March-April 2006.
33. **Jenkins, S.** Real-World Experience with IGCC – Lessons Learned in Design, Permitting, Operations and Maintenance. s.l. : URS Corporation, 2005.
34. **Drnevich, Ray.** Personal communication. *ray_drnevich@praxair.com*. s.l. : Praxair, Inc., November 2002.
35. http://www.calgoncarbon.com/bulletins/Mercury_Removal_Gas_Streams.htm. [Online]
36. <http://www.calgoncarbon.com/bulletins/HGR-P.htm>. [Online]
37. http://www.calgoncarbon.com/bulletins/TYPE_HGR.htm. [Online]
38. **Electric Power Research Institute.** *Process Screening Study of Alternative Gas Treating and Sulfur Removal Systems for IGCC Power Plant Applications*. December 1987. EPRI AP-5505.
39. **CB&I.** E-Gas™ technology. *CB&I*. [Online] 2013. [Cited: November 23, 2013.] <http://www.cbi.com/technologies/gasification-technology>.
40. **National Energy Technology Laboratory.** *Process Screening Analysis of Alternative Gas Treating and Sulfur Removal for Gasification, Revised Final Report*. Pittsburgh, Pa : U.S. Department of Energy, December 2002.
41. **Beychok, Milton.** Wikimedia Commons. [Online] 2014. [Cited: June 11, 2014.] http://en.wikipedia.org/wiki/Claus_process#mediaviewer/File:Claus_Sulfur_Recovery.png.
42. **GE Power Systems.** The 7FB: The Next Evolution of the F Gas Turbine. [Online] 2014. [Cited: November 12, 2014.] http://site.ge-energy.com/prod_serv/products/tech_docs/en/downloads/ger4194.pdf. GER-4194.
43. **Brdar, R. D. and R.M. Jones.** *GE IGCC Technology and Experience with Advanced Gas Turbines*. Schenectady, NY, : GE Power Systems, 10/2000. GER-4207.
44. **GE Power Systems.** *Specification for Fuel Gases for Combustion in Heavy-Duty Turbines*. January 2002. GEI-41040G.
45. —. *Design Considerations for Heated Gas Fuel*. March 2003. GEK-4189.
46. **Electric Power Research Institute.** *Coal Gasification Guidebook: Status, Applications, and Technologies*. s.l. : EPRI, December 1993. TR-102034.
47. **Holthoon, E. Van.** *Shell Coal Gasification – Leading Technology Across Multiple Applications*. s.l. : Gasification Technologies Conference, October 6, 2009.
48. **Van Der Ploeg, H.J., T. Chhoa, and P.L. Zuideveld.** *The Shell Coal Gasification Process for the US Industry*. Washington DC : Gasification Technology Conference, October 2004.
49. **Breton, D. and C. Keeler.** *Comparative IGCC Performance and Costs for Domestic Coals*. San Francisco, CA : Gasification Technologies Council, October 2005.
50. **Rigdon, R. and L. Schmoe.** *The IGCC Reference Plant*. San Francisco, CA : Gasification Technologies Conference 2005, October 2005.

51. **Rigdon, R. and K. Miles.** The Cleaner Coal Option. *Power Engineering International*. July 2006.
52. **National Energy Technology Laboratory.** Power Systems Financial Model Version 6.6 and User's Guide. *Models/Tools* . Pittsburgh, Pa : U. S. Department of Energy, 2012. DOE/NETL-2011/1492 .
53. —. *Updated Costs (June 2011 Basis) for Selected Bituminous baseline Cases*. Pittsburgh : Department of Energy, 2012. 341/082312.
54. **U.S. Environmental Protection Agency.** U.S. Government Printing Office. *Electronic Code of Federal Regulations*. [Online] 2014. [Cited: June 11, 2014.] http://www.ecfr.gov/cgi-bin/text-idx?tpl=/ecfrbrowse/Title40/40cfr60_main_02.tpl.

Tim Fout
timothy.fout@netl.doe.gov

Mark Woods
mark.woods@netl.doe.gov



www.netl.doe.gov
Pittsburgh, PA • Morgantown, WV • Albany, OR • Sugar Land, TX • Anchorage, AK
(800) 553-7681