

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

ARUN K.S. IYENGAR, ALEX A. NORING, RICHARD
A. NEWBY, DALE L. KEAIRNS



April 30, 2022

DOE/NETL-2022-3250

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.

All images in this report were created by NETL, unless otherwise noted.

Authors

Arun K.S. Iyengar^{1,2}: Methodology, Model Development, Formal Analysis, Writing – Original draft, Supervision; **Alex A. Noring**^{1,2}: Model Development, Writing – Review & Editing; **Richard A. Newby**^{1,2}: Model Development, **Dale L. Keairns**^{1,2}: Supervision; **Gregory A. Hackett**^{1*}: Supervision, Review & Editing

¹ National Energy Technology Laboratory (NETL)

² NETL Support Contractor

*Corresponding Contact: Gregory.Hackett@netl.doe.gov, 304.285.5279

Suggested Citation:

A.K.S. Iyengar, A. Noring, R. Newby, D. Keairns, "Techno-economic Analysis of Integrated Gasification Fuel Cell Systems," National Energy Technology Laboratory, Pittsburgh, April 30, 2022.

The authors gratefully acknowledge to contribution from the following:

Gregory Hackett, NETL, Energy Process Analysis Team

Travis Shultz, NETL, Team Supervisor, Energy Process Analysis Team

Shailesh Vora, NETL, Technology Manager, Fuel Cells

Joseph Pierre and **Mark Woods**, NETL Support Contractors

Brian Koeppel, **Jie Bao**, and **Wang Chao**, Pacific Northwest National Laboratory

This page intentionally left blank.

TABLE OF CONTENTS

List of Exhibits	iii
Acronyms and Abbreviations	vi
Executive Summary	1
Discussion of Pathway Results	11
Key Conclusions	24
1 Introduction.....	26
2 Pathway Study Basis	33
2.1 Site Characteristics	33
2.2 Coal Characteristics.....	34
2.3 Natural Gas Characteristics.....	35
2.4 Environmental Targets	36
2.5 Balance of Plant	37
2.6 Plant Capacity.....	39
2.7 Capacity Factor	39
2.8 Sparing Philosophy and Number of Parallel Process Trains	39
2.9 SOFC Power Island Characterization.....	41
2.9.2 SOFC Carbon Deposition Control	43
2.9.3 Estimation of Steam Bottoming Cycle Performance	43
2.9.4 IGFC Power Island	44
2.9.5 VGR Concept	48
2.10 Air Separation Unit	51
2.11 CO ₂ Capture Technology	52
2.12 Raw Water Withdrawal and Consumption	52
2.13 Economic Analysis	53
2.13.1 Cost Estimating Methodology	53
2.13.2 Plant Maturity	54
2.13.3 Estimate Scope	54
2.13.4 Capital Costs.....	55
2.13.5 SOFC Module Costs.....	56
2.13.6 Exclusions.....	58
2.13.7 Contingency	58
2.13.8 Operation and Maintenance.....	59
2.13.9 Operating Labor	59
2.13.10 Maintenance Material and Labor	59
2.13.11 Stack Degradation and Production Costs	59
2.13.12 Administrative and Support Labor	60

2.13.13	Consumables	60
2.13.14	Waste Disposal.....	60
2.13.15	Co-products and Byproducts.....	60
2.13.16	Owner's Costs	61
2.13.17	CO ₂ Transport and Storage	63
2.13.18	Breakeven Cost of CO ₂ Captured	63
3	IGFC Plant Major Process Areas	64
3.1	Descriptions of Process Areas	64
3.1.1	Coal Receiving and Storage Area.....	64
3.1.2	Air Separation Unit	64
3.1.3	Coal Gasification Area	66
3.1.4	Syngas Cleaning Area	73
3.1.5	IGFC Power Island	79
3.1.6	Heat Recovery Steam Generator and Steam Bottoming Cycle.....	84
3.1.7	CO ₂ Compression, Drying, and Purification.....	84
3.1.8	Accessory Electric Plant	85
3.1.9	Instrumentation and Control	85
4	Scenario 1 – IGFC with Atmospheric-Pressure SOFC	86
4.1	Reference Atmospheric IGFC	86
4.2	Case 3: Enhanced Gasifier Atmospheric IGFC	100
4.3	Case 6BV: Advanced Gasifier Atmospheric IGFC with VGR	113
4.4	Pathway 1 Results	128
5	Scenario 2 – IGFC with Pressurized SOFC.....	133
5.1	Reference Pressurized IGFC.....	133
5.2	Case 6: Advanced Gasifier Pressurized IGFC.....	148
5.3	Pathway 2 Results	163
6	Discussion of Pathway Results.....	168
6.1	System Performance and Efficiency	168
6.2	IGFC Plant Costs	171
6.3	Cost of CO ₂ Captured	176
6.4	Water Consumption	176
6.5	Effect of Natural Gas Price	179
7	Conclusions.....	180
8	References	182

LIST OF EXHIBITS

Exhibit ES-1. SOFC Pathway Studies	2
Exhibit ES-2. Reference IGFC System Characteristics	3
Exhibit ES-3. IGFC Pathway Cases without CCS	5
Exhibit ES-4. IGFC Pathway Cases with CCS	6
Exhibit ES-5. Pathway 1 – Atmospheric IGFC Plants with CCS Results Summary	7
Exhibit ES-6. Pathway 1 – Atmospheric IGFC Plants without CCS Results Summary	8
Exhibit ES-7. Pathway 2 – Pressurized IGFC Plants with CCS Results Summary	9
Exhibit ES-8. Pathway 2 – Pressurized IGFC Plants without CCS Results Summary	10
Exhibit ES-9. HHV Efficiency of the Pathway IGFC Plants	12
Exhibit ES-10. Technology Developments and Efficiency Improvements for IGFC Plants without CCS	13
Exhibit ES-11. Technology developments and efficiency improvements for IGFC plants with CCS	13
Exhibit ES-12. Comparison of the TPC of the IGFC Plants	15
Exhibit ES-13. Comparison of the LCOE (without T&S) of the IGFC Plants	16
Exhibit ES-14. LCOE Breakdown (with T&S) of the IGFC Plants	17
Exhibit ES-15. LCOE Breakdown (without T&S) of the IGFC Plants	17
Exhibit ES-16. Progression of Plant LCOE (without T&S) along the IGFC Pathway without CCS	18
Exhibit ES-17. Progression of Plant LCOE (without T&S) along the IGFC Pathway with CCS	19
Exhibit ES-18. Breakeven Sale Price of CO ₂ Captured for the Various IGFC Plants	20
Exhibit ES-19. Water Consumption Variation for the Various IGFC plants	21
Exhibit ES-20. Influence of Gasifier Capital Costs on the LCOE (excl. T&S) of IGFC Plants with CCS	22
Exhibit ES-21. Influence of Natural Gas Price on the LCOE (excl. T&S) of IGFC Plants with CCS	23
Exhibit ES-22. Influence of Natural Gas Price on the LCOE (excl. T&S) of IGFC Plants without CCS	23
Exhibit 1-1. IGFC System Configuration with CCS	27
Exhibit 1-2. SOFC Pathway Studies	28
Exhibit 1-3. Reference IGFC System Characteristics	29
Exhibit 1-4. IGFC Pathway Cases without CCS	31
Exhibit 1-5. IGFC Pathway Cases with CCS	32
Exhibit 2-1. Site Ambient Conditions	33
Exhibit 2-2. Site Characteristics	33
Exhibit 2-3. Design Coal	35
Exhibit 2-4. Natural Gas Composition	36
Exhibit 2-5. Environmental Targets for IGFC cases	37
Exhibit 2-6. Balance of Plant Assumptions	38
Exhibit 2-7. General IGFC Plant Configuration with CCS	40
Exhibit 2-8. General IGFC Plant Configuration without CCS	41
Exhibit 2-9. Overview of the ROM Generation Procedure and Usage	42
Exhibit 2-10. IGFC Module Sizing	45
Exhibit 2-11. IGFC Module Layout	46
Exhibit 2-12. SOFC Section Layout: Components and Dimension	47
Exhibit 2-13. Characteristics of SOFC section	47
Exhibit 2-14. Atmospheric IGFC System Layout	48
Exhibit 2-15. Generalized Configuration of the VGR Concept	49
Exhibit 2-16. Configuration of the VGR Concept without an H ₂ Membrane	50
Exhibit 2-17. Configuration of the Proposed System without both the WGS Reactor and the H ₂ Membrane	51
Exhibit 2-18. Atmospheric SOFC Module Costs	57
Exhibit 2-19. Pressurized SOFC Module Costs	57
Exhibit 2-20. Summary of O&M Costs	61
Exhibit 2-21. Owner's Costs Included in TOC	62
Exhibit 3-1. Coal Gasification Section Assumptions for the CoP E-Gas™ Gasifier	67
Exhibit 3-2. Coal Gasification Section Assumptions for the Enhanced Gasifier	69
Exhibit 3-3. Catalytic Gasifier Coal/Catalyst Processing	71
Exhibit 3-4. Coal Gasification Section Assumptions with Catalytic Gasifier	73
Exhibit 3-5. Syngas Cleaning for IGFC	74
Exhibit 3-6. Gas Cleaning Area Assumptions for the Conventional Gasifier Cases	75
Exhibit 3-7. Gas Cleaning Section Assumptions for the Catalytic Gasifier	75
Exhibit 3-8. IGFC Power Island – Scenario 1 – Atmospheric SOFC	80
Exhibit 3-9. IGFC Power Island – Scenario 2 – Pressurized SOFC	81
Exhibit 3-10. IGFC Power Island Reference Case Assumptions	83
Exhibit 3-11. Auto-Refrigerated CPU Process Developed for the Present Study	85
Exhibit 4-1. IGFC Parameters Used for the Atmospheric Reference Case	86
Exhibit 4-2. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Process Diagram	88
Exhibit 4-3. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Stream Table	89

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-4. Case ACONVIGFC0A – Atmospheric Reference IGFC Plant without CCS – Process Diagram	90
Exhibit 4-5. Case 0 – Atmospheric Reference IGFC Plant with and without CCS – Performance Summary	91
Exhibit 4-6. Case 0 – Atmospheric Reference IGFC Plant with and without CCS – Gross Power Distribution	92
Exhibit 4-7. Case 0 – Atmospheric Reference IGFC Plant with and without CCS – Auxiliary Load Distribution	92
Exhibit 4-8. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Material and Energy Balances	93
Exhibit 4-9. Case ACONVIGFC0B – Reference Plant with CCS – Heat and Material Balance Diagram – ASU, Gasification, and Clean-up ..	94
Exhibit 4-10. Case ACONVIGFC0B – Reference Plant with CCS – Heat and Material Balance Diagram – Power Block.....	95
Exhibit 4-11. Case ACONVIGFC0B – Reference Plant with CCS – Heat and Material Balance Diagram – CO ₂ Dehydration, Purification, and Compression	96
Exhibit 4-12. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Capital Costs.....	97
Exhibit 4-13. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – O&M Costs.....	98
Exhibit 4-14. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Capital Cost Components	99
Exhibit 4-15. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – BOP Cost Distribution.....	100
Exhibit 4-16. Case 0 – Atmospheric reference IGFC plant with and without CCS – levelized cost of electricity	100
Exhibit 4-17. IGFC Parameters Used for the Enhanced Gasifier Case 3.....	101
Exhibit 4-18. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Process Diagram	103
Exhibit 4-19. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Stream Table	104
Exhibit 4-20. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Performance Summary	105
Exhibit 4-21. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Gross Power Distribution	106
Exhibit 4-22. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Auxiliary Load Distribution	106
Exhibit 4-23. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Material and Energy Balances	107
Exhibit 4-24. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC without CCS – Heat and Material Balance Diagram – ASU, Gasification, and Clean-up	108
Exhibit 4-25. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC without CCS – Heat and Material Balance Diagram – Power Block	109
Exhibit 4-26. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC plant without CCS – Capital Costs	110
Exhibit 4-27. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – O&M Costs	111
Exhibit 4-28. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Capital Cost Components	112
Exhibit 4-29. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – BOP Cost Distribution	113
Exhibit 4-30. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Levelized Cost of Electricity	113
Exhibit 4-31. IGFC Parameters Used for the Advanced Gasifier Case 6BV	114
Exhibit 4-32. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Process Diagram	116
Exhibit 4-33. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Stream Table.....	117
Exhibit 4-34. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Performance Summary	118
Exhibit 4-35. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Gross Power Distribution	119
Exhibit 4-36. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Auxiliary Load Distribution	119
Exhibit 4-37. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Material and Energy Balances.....	120
Exhibit 4-38. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Heat and Material Balance Diagram – ASU, Gasifier, and Clean-up	121
Exhibit 4-39. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Heat and Material Balance Diagram – Power Block.....	122
Exhibit 4-40. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Heat and Material Balance Diagram – CO ₂ Dehydration, Purification, and Compression	123
Exhibit 4-41. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC plant with VGR – Capital Costs	124
Exhibit 4-42. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – O&M Costs	125
Exhibit 4-43. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Capital Cost Components.....	126
Exhibit 4-44. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – BOP Cost Distribution	127
Exhibit 4-45. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC plant with VGR – Levelized Cost of Electricity	127
Exhibit 4-46. Pathway 1 – Atmospheric IGFC Plants with CCS Performance Summary	129
Exhibit 4-47. Pathway 1 – Atmospheric IGFC Plants with CCS Cost Summary	130
Exhibit 4-48. Pathway 1 – Atmospheric IGFC Plants without CCS Performance Summary	131
Exhibit 4-49. Pathway 1 – Atmospheric IGFC Plants without CCS Cost Summary	132
Exhibit 5-1. IGFC Parameters Used for the Pressurized Reference Case	133
Exhibit 5-2. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Process Diagram.....	135
Exhibit 5-3. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Stream Table	136
Exhibit 5-4. Case PCONVIGFC0A – Reference Pressurized IGFC Plant without CCS – Process Diagram	137
Exhibit 5-5. Reference Pressurized IGFC plant with and without CCS – Performance Summary.....	138
Exhibit 5-6. Reference Pressurized IGFC Plant with and without CCS – Gross Power Distribution.....	139
Exhibit 5-7. Reference Pressurized IGFC Plant with and without CCS – Auxiliary Load Distribution.....	139
Exhibit 5-8. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Material and Energy Balances	140
Exhibit 5-9. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – ASU, Gasifier, and Clean-up	141
Exhibit 5-10. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – Power Block.....	142
Exhibit 5-11. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – CO ₂ Dehydration, Purification, and Compression	143

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 5-12. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Capital Costs.....	144
Exhibit 5-13. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – O&M Costs.....	145
Exhibit 5-14. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Capital Cost Components	146
Exhibit 5-15. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – BOP Cost Distribution.....	147
Exhibit 5-16. Case 0 – Reference pressurized IGFC plant with and without CCS – levelized cost of electricity	147
Exhibit 5-17. IGFC Parameters Used for the Advanced Gasifier Case 6 BV	148
Exhibit 5-18. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Process Diagram	150
Exhibit 5-19. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Stream Table.....	151
Exhibit 5-20. Advanced Gasifier Pressurized IGFC Plant with and without CCS – Performance Summary	152
Exhibit 5-21. Advanced Gasifier Pressurized IGFC Plant with and without CCS – Gross Power Distribution	153
Exhibit 5-22. Advanced Gasifier Pressurized IGFC Plant with and without CCS – Auxiliary Load Distribution	153
Exhibit 5-23. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Material and Energy Balances.....	154
Exhibit 5-24. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – ASU, Gasifier, and Clean-up	155
Exhibit 5-25. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – Power Block	156
Exhibit 5-26. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – CO ₂ Dehydration, Purification, and Compression	157
Exhibit 5-27. Case PCATIGFC6BV – Advanced Gasifier Pressurized IGFC Plant with CCS – Capital Costs.....	158
Exhibit 5-28. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – O&M Costs	159
Exhibit 5-29. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Capital Cost Components	160
Exhibit 5-30. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – BOP Cost Distribution	161
Exhibit 5-31. Case PCATIGFC6A – Advanced Gasifier Pressurized IGFC Plant without CCS – Capital Cost Components	161
Exhibit 5-32. Case PCATIGFC6A – Advanced Gasifier Pressurized IGFC Plant without CCS – BOP Cost Distribution.....	162
Exhibit 5-33. Case 6 – Advanced Gasifier Pressurized IGFC Plant with and without CCS – Levelized Cost of Electricity	162
Exhibit 5-34. Pathway 2 – Pressurized IGFC Plants with CCS Performance Summary.....	164
Exhibit 5-35. Pathway 2 – Pressurized IGFC Plants with CCS Cost Summary	165
Exhibit 5-36. Pathway 2 – Pressurized IGFC Plants without CCS Performance Summary	166
Exhibit 5-37. Pathway 2 – Pressurized IGFC Plants without CCS Cost Summary	167
Exhibit 6-1. HHV Efficiency of the Pathway IGFC Plants	169
Exhibit 6-2. Technology Developments and Efficiency Improvements for IGFC Plants without CCS	170
Exhibit 6-3. Technology Developments and Efficiency Improvements for IGFC Plants with CCS	170
Exhibit 6-4. Comparison of the TPC of the IGFC Plants.....	172
Exhibit 6-5. Comparison of the LCOE (without T&S) of the IGFC Plants.....	173
Exhibit 6-6. LCOE Breakdown (with T&S) of the IGFC Plants	174
Exhibit 6-7. LCOE Breakdown (without T&S) of the IGFC plants	174
Exhibit 6-8. Progression of Plant LCOE (without T&S) along the IGFC Pathway without CCS.....	175
Exhibit 6-9. Progression of Plant LCOE (without T&S) along the IGFC Pathway with CCS	176
Exhibit 6-10. Breakeven Sale Price of CO ₂ Captured for the Various IGFC Plants	177
Exhibit 6-11. Water Consumption for the Various IGFC Plants.....	178
Exhibit 6-12. Influence of Natural Gas price on the LCOE (excl. T&S) of IGFC Plants with CCS.....	179
Exhibit 6-13. Influence of Natural Gas Price on the LCOE (excl. T&S) of IGFC Plants without CCS	179

ACRONYMS AND ABBREVIATIONS

AACE	Association for the Advancement of Cost Engineering	Cref	Cost of equipment in reference plant area or section
AC	Alternating current	CRT	Cathode ray tube
AFUDC	Allowance for funds used during construction	DC	Direct current
AGR	Acid gas removal	DCS	Distributed control system
Ar	Argon	DG	Distributed generation
Aspen	Aspen Plus®	DGFC	Distributed generation fuel cell
ASR	Area specific resistance	DI	De-ionized
ASU	Air separation unit	DOE	Department of Energy
atm	Atmospheric	DSRP	Direct sulfur recovery process
AUSC	Advanced ultrasupercritical	EAF	Equivalent availability factor
B&V	Black and Veatch	E-Gas™	ConocoPhillips gasifier technology
BACT	Best available control technology	EIA	Energy Information Administration
BB	Bituminous Baseline	EOR	Enhanced oil recovery
BEC	Bare erected cost	EPA	Environmental Protection Agency
BFD	Block flow diagram	EPC	Engineer/Procure/Construct
BFW	Boiler feedwater	EPCM	Engineering/Procurement/Construction Management
BOP	Balance of plant	EPRI	Electric Power Research Institute
Btu	British thermal unit	ESPA	Energy Sector Planning and Analysis
Btu/hr	British thermal unit per hour	F	Capacity of study plant area or section
Btu/kWh	British thermal unit per kilowatt hour	FEED	Front-end engineering design
Btu/lb	British thermal unit per pound	Fref	Capacity of reference plant area or section
Btu/scf	British thermal unit per standard cubic foot	ft	Foot, feet
C	Cost of equipment in study plant area of section	ft ³	Cubic feet
Carb	Number of atoms of carbon in the syngas	gal	Gallon
CCS	Carbon capture and storage	GEE	General Electric Energy
CEPCI	Chemical Engineering Plant Cost Index	GHG	Greenhouse gas
CF	Capacity factor	gpm	Gallons per minute
CH ₄	Methane	h, hr	Hour
cm	Centimeter	H ₂	Hydrogen
CO	Carbon monoxide	H ₂ O	Water
CO ₂	Carbon dioxide	H ₂ S	Hydrogen sulfide
CoP	ConocoPhillips	HCl	Hydrogen chloride
COS	Carbonyl sulfide	Hg	Mercury
CPU	CO ₂ purification unit	HHV	Higher heating value
		hp	Horsepower

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

HP	High pressure	MJ/scm	Megajoule per standard cubic meter
HRSG	Heat recovery steam generator	MMBtu	Million British thermal units (also shown as 10 ⁶ Btu)
HTX, HX	Heat exchanger	MMBtu/hr	Million British thermal units (also shown as 10 ⁶ Btu) per hour
HVAC	Heating, ventilation, and air conditioning	MMkJ	Million kilojoules (also shown as 10 ⁶ kJ)
I&C	Instrumentation and control	MMkJ/hr	Million kilojoules (also shown as 10 ⁶ kJ) per hour
ID	Identification	MP	Multi-physics
IGCC	Integrated gasification combined cycle	MPa	Megapascals
IGFC	Integrated gasification fuel cell	MU	Make up
IOU	Investor-owned utility	mV	Millivolt
IP	Intermediate pressure	MW	Megawatt
IR	Internal reformation	MWe	Megawatt electric
ISO	International Standards Organization	MWh	Megawatt-hour
ITM	Ion transport membrane	N	Number of study plant areas or sections in parallel
kg/GJ	Kilogram per gigajoule	N/A	Not applicable
kg/hr	Kilogram per hour	N ₂	Nitrogen
kJ	Kilojoules	Nref	Number of reference plant areas or sections in parallel
kJ/hr	Kilojoules per hour	NERC	North American Electric Reliability Council
kJ/kg	Kilojoules per kilogram	NETL	National Energy Technology Laboratory
KOH	Potassium hydroxide	NG	Natural gas
kV	Kilovolt	NGCC	Natural gas combined cycle
kW	Kilowatt	NGFC	Natural gas fuel cell
kWe	Kilowatts electric	Nm ³	Normal cubic meter
kWh	Kilowatt-hour	NOx	Oxides of nitrogen
kWt	Kilowatts thermal	NSPS	New Source Performance Standards
lb	Pound	O&M	Operation and maintenance
lb/hr	Pounds per hour	O ₂	Oxygen
lb/MMBtu	Pounds per million British thermal units	OCFn	Category n fixed operating cost for the initial year of operation
lb/MWh	Pounds per megawatt hour	OCR	Oxygen-to-carbon ratio
LCOE	Levelized cost of electricity	OCVnq	Category n variable operating cost for the initial year of operation
LGTI	Louisiana Gasification Technology, Inc.	OSHA	Occupational Safety and Health Administration
LHV	Lower heating value	OTC	Oxygen to carbon (ratio)
LNB	Low NOx burner		
LP	Low pressure		
m	Meter		
m ³ /min	Cubic meter per minute		
mA/cm ²	Milliamps per square cm		
MAF	Moisture and Ash Free		
MDEA	Methyldiethanolamine		
MESA	Mission Execution and Strategic Analysis		

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Oxy	Number of atoms of oxygen in the syngas	SOFC	Solid oxide fuel cell
PC	Pulverized coal	SRU	Sulfur recovery unit
PFD	Process flow diagrams	SWS	Sour water stripper
PM	Particulate matter	Syngas	Synthesis gas
PNNL	Pacific Northwest National Laboratory	T	Temperature
POTW	Publicly owned treatment works	T&S	Transport and storage
ppbv	Parts per billion volume	TASC	Total as-spent cost
ppm	Parts per million	TG	Tail gas
ppmv	Parts per million volume	TGTU	Tail gas treating unit
ppmvd	Parts per million volume, dry	TOC	Total overnight cost
PSA	Pressure swing adsorption	ton	Short ton (2000 lbs)
psia	Pounds per square inch absolute	Tonne	Metric ton (1000 kg)
psid	Pounds per square inch differential	TPC	Total plant cost
psig	Pounds per square inch gauge	TPD	Tons per day
QGESS	Quality Guidelines for Energy System Studies	TRL	Technology Readiness Level
R&D	Research and development	TS&M	Transport, storage, and monitoring
ROM	Reduced order model	U.S.	United States
S	Scaling factor for plant areas or section cost	V	Volt
SC	Supercritical	V-L	Vapor liquid portion of stream (excluding solids)
sCO ₂	Supercritical carbon dioxide	VGR	Vent gas recirculation
SCOT	Shell Claus Off-gas Treating	vol%	Volume percent
SCR	Selective catalytic reduction	WGS	Water gas shift
SG	Specific gravity	WT	Water treatment
SGC	Synthesis gas cooler	wt%	Weight percent
SGS	Sour gas shift	ZnO	Zinc oxide
SNG	Synthetic natural gas	\$/MMBtu	Dollars per million British thermal units
SO ₂	Sulfur dioxide	\$/MMkJ	Dollars per million kilojoules
SOA	State of the art	°C	Degrees Celsius
		°F	Degrees Fahrenheit

This page intentionally left blank.

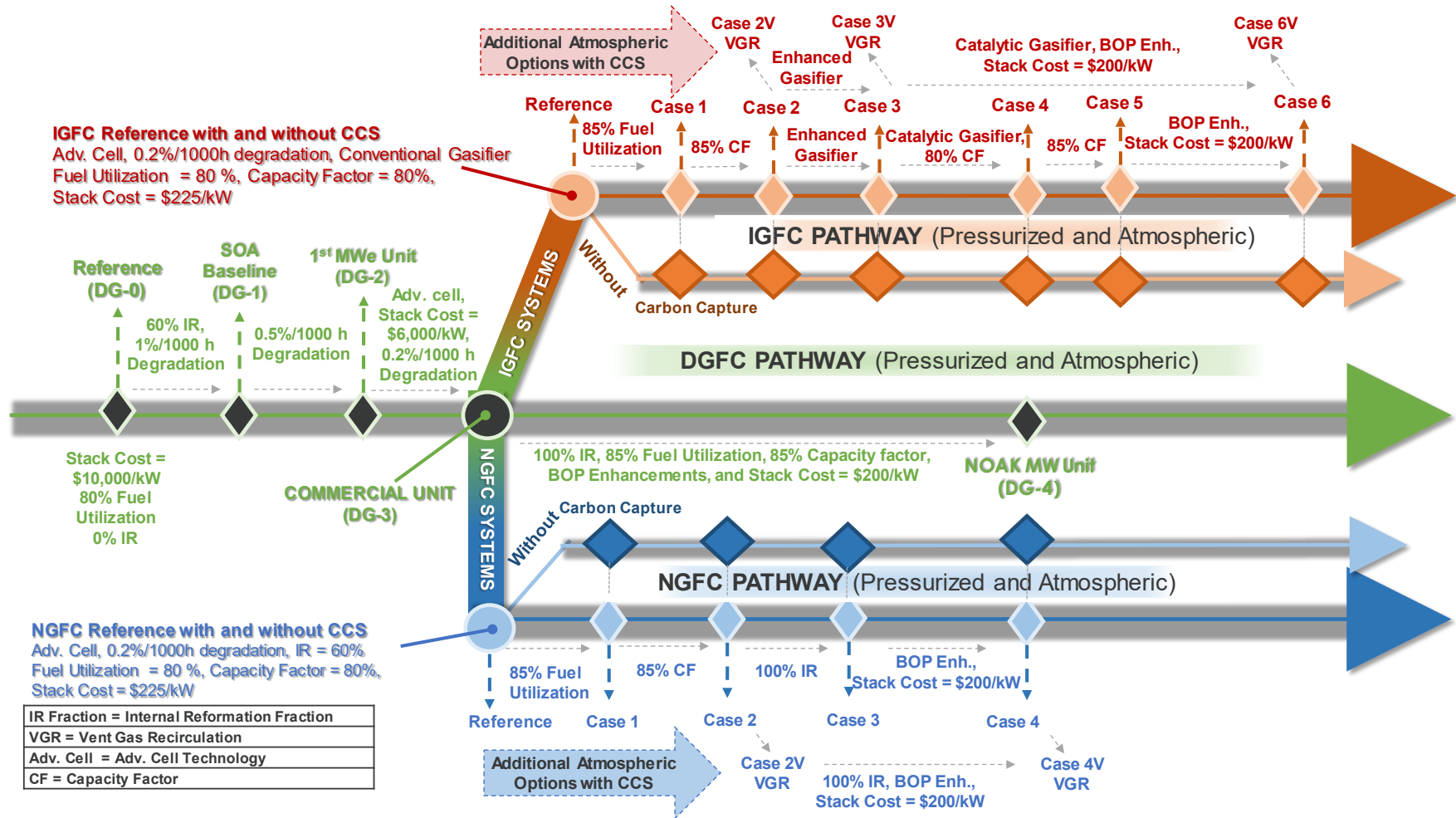
EXECUTIVE SUMMARY

The United States (U.S.) Department of Energy (DOE) Office of Fossil Energy and Carbon Management (FECM) and the National Energy Technology Laboratory (NETL) have been pursuing the development of solid oxide fuel cell (SOFC) technology to enable future power generation systems that are consistent with the cornerstones of the DOE mission—to ensure America’s security and prosperity by addressing its energy and environmental challenges through transformative science and technology solutions. The U.S. DOE FECM SOFC Program is currently focused on the development of low-cost, highly efficient, and reliable SOFC power systems. NETL’s SOFC technology development roadmap is aligned with near-term market opportunities in the distributed generation sector to validate and advance the technology while paving the way for utility-scale (> 50 MW) natural gas (NG)- and coal-derived synthesis gas-fueled applications via progressively larger system demonstrations. The present study describes the results of a pathway study for coal-based integrated gasification fuel cell (IGFC) power systems with and without carbon capture and storage (CCS). The objective of the study is to provide targeted research and development (R&D) guidance to the FECM SOFC Program and SOFC commercial developers.

The IGFC power plant is analogous to an integrated gasification combined cycle (IGCC) power plant, but with the gas turbine power island replaced with a solid oxide fuel cell (SOFC) system. The SOFC provides high electric efficiencies associated with the nearly reversible electrochemical conversion of syngas chemical potential to electric power, as compared to a conventional Carnot-driven heat engine. In addition, the clean oxy-reaction of syngas in the sealed fuel cell system renders itself readily available for CCS with the requirement of only an oxy-combustor downstream of the fuel cell to react the fuel that is not utilized electrochemically. (Electrochemical utilization of fuels typically varies between 75 and 90 percent for current fuel cell technology, due to practical considerations). The heat rejected by the fuel cell system can be recovered further in a combination of Brayton and Rankine cycles, depending on whether the fuel cell system is operating at elevated or atmospheric pressures.

The fuel cell technology is assumed to be a planar configuration with separated anode and cathode off-gas streams for the present pathway studies. The overall framework for the SOFC system R&D pathway considered in this study is depicted in Exhibit ES-1, which takes into account anticipated improvements in SOFC electrochemical performance, gasifier technology, and system configuration supported by the SOFC technology development roadmap, combined with concomitant advances in plant operation and maintenance (O&M) and component technologies. The pathway explores systems with and without CCS. With a commercially viable, small-scale (≈ 1 MWe) distributed generation fuel cell (DGFC) system as a basis, the development of utility-scale systems is envisioned to proceed along two parallel pathways based on the fuel feedstock. The IGFC pathway is based on synthetic gas (syngas) generated via coal gasification, while the Natural Gas Fuel Cell (NGFC) pathway is based on development of natural gas based SOFC systems. The DGFC and NGFC pathways are reported in separate reports, while the results of the IGFC pathway are discussed here.

Exhibit ES-1. Solid Oxide Fuel Cells Pathway Studies



The SOFC technology and system characteristics, listed in Exhibit ES-2, of the established commercial DGFC unit (as shown in the DGFC developmental roadmap¹ in Exhibit ES-1) was assumed as the basis for the reference IGFC system, which assumes the readiness of a conventional gasification technology, and features a gasifier, akin to the E-Gas™ gasifier, with a syngas methane (CH₄) content of roughly six mole percent.

Exhibit ES-2. Reference IGFC System Characteristics

SOFC Parameters	Values
SOFC Performance	Advanced Cell – 50% reduction in Ohmic and Polarization losses relative to the state-of-the-art technology (SOA)
Degradation Rate	0.2% per 1000 h
System Fuel Utilization	80%
Stack Cost	\$225/kWe
Inverter Efficiency	97%
Gasifier Configuration	Conventional
Capacity Factor	80%

Two parallel developmental scenarios—one with the SOFC operating at atmospheric pressure and one based on a pressurized SOFC operation—were considered as part of the IGFC development pathway. An SOFC operating pressure of three atmospheres was assumed in the present calculations based on previous studies, which found insignificant cost benefits for pressures higher than this, and pressurized SOFC systems that have been generally proposed and operated by industrial vendors. Two gasification technology improvements differing in the syngas CH₄ content, which has a significant effect on the IGFC system performance, were considered.

The specific IGFC pathway cases analyzed in this study are listed in Exhibit ES-3 and Exhibit ES-4 for systems without and with CCS, respectively. Components for each plant configuration are described in more detail in the corresponding report sections for each pathway. Case 1 represents an increase in fuel utilization from 80 to 85 percent based on the confidence gained from DGFC systems. Enhanced reliability of balance of plant (BOP) components, combined with optimized O&M is reflected by an increase in capacity factor to 85 percent in Case 2. An enhanced gasifier, representing an incremental adaptation in gasifier technology to generate a syngas with higher methane content relative to the conventional gasifier, is assumed for Case 3. A transformational catalytic gasification technology, which produces a high methane content (≈30 mole percent) syngas with improved IGFC performance benefits, is assumed to be available, albeit initially at a capacity factor of 80 percent, in Case 4. Improvement of the capacity factor of Case 4 system to 85 percent is considered in Case 5, while Case 6 considers potential enhancements in air separation unit (ASU) and inverter technologies, along with a

¹ The DGFC pathway will be reported elsewhere.

reduction in stack cost to \$200/kW. In the case of systems with CCS, an attractive high-efficiency vent gas recirculation (VGR) configuration, which enables fuel utilization values >95 percent, is also explored with systems reflected by the three gasifier technologies, Cases 2, 3, and 6.

The design and cost bases for this evaluation are largely based on National Energy Technology Laboratory's (NETL) Bituminous Baseline (BB) report [2], to facilitate a direct comparison to the baseline results for other fossil fuel power generation technologies. The basis for the design of the SOFC power island components and their cost estimates are described in Section 2 of the report. The IGFC plants are designed for baseload operation with the following key design basis specifications:

- Illinois No. 6 coal
- International Standards Organization (ISO) ambient conditions
- Conventional cryogenic air separation technology
- Conventional dry syngas cleaning and polishing technology
- Cryogenic distillation-based carbon dioxide (CO₂) purification process to meet enhanced oil recovery (EOR) specifications for the CO₂ product stream
- Net plant capacity of 650 MWe
- Coal price is assumed to be \$51.96/ton and the natural gas (NG) price is assumed to be \$4.42/MMBtu

The cost and performance results for the two pathway cases are summarized in Exhibit ES-5 through Exhibit ES-8 and are discussed in the next section.

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit ES-3. IGFC Pathway Cases without CCS

Case Designation	Pathway Parameter	Case ID	SOFC Pressure (atm)	SOFC Technology	Gasifier Technology	Dry Syngas CH ₄ Content (%)	Capacity Factor (%)	Degradation (%/1000 h)	Fuel Utilization (%)	Inverter Efficiency	Stack Cost (\$/kW)				
Reference (Case 0)	Reference Without CCS	ACONVIGFC0A	1	Advanced Cell	Conventional	≈5.8	80	0.2	80	97%	225				
		PCONVIGFC0A	3												
Case 1	85% Fuel Utilization	ACONVIGFC1A	1												
		PCONVIGFC1A	3												
Case 2	85% Capacity Factor	ACONVIGFC2A	1				85								
		PCONVIGFC2A	3												
Case 3	Enhanced Gasifier	AENHIGFC3A	1		Enhanced	≈10.8									
		PENHIGFC3A	3												
Case 4	Catalytic Gasifier, 80% Capacity Factor	ACATIGFC4A	1		Catalytic	≈24.6	80		85						
		PCATIGFC4A	3												
Case 5	85% Capacity Factor	ACATIGFC5A	1				85								
		PCATIGFC5A	3												
Case 6	BOP Enhancements 20% ASU Power Reduction, Inverter Efficiency 98%, Stack Cost Reduction	ACATIGFC6A	1				98%			200					
		PCATIGFC6A	3												

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit ES-4. IGFC Pathway Cases with CCS

Case Designation	Pathway Parameter	Case ID	SOFC Pressure (atm)	SOFC Technology	Gasifier Technology	Dry Syngas CH ₄ Content (%)	Capacity Factor (%)	Degradation (%/1000 h)	Fuel Utilization (%)	Inverter Efficiency	Stack Cost (\$/kW)				
Reference (Case 0)	Reference Without CCS	ACONVIGFC0B	1	Advanced Cell	Conventional	≈5.8	80	0.2	80	97%	225				
		PCONVIGFC0B	3												
Case 1	85% Fuel Utilization	ACONVIGFC1B	1						85						
		PCONVIGFC1B	3												
Case 2	85% Capacity Factor	ACONVIGFC2B	1				85		97.5						
		PCONVIGFC2B	3												
Case 2 V	Case 2 with VGR	ACONVIGFC2BV	1		85		97.5								
Case 3	Enhanced Gasifier	AENHIGFC3B	1				Enhanced		≈10.8			85			
		PENHIGFC3B	3									97.5			
Case 3V	Case 3 with VGR	AENHIGFC3BV	1									97.5			
Case 4	Catalytic Gasifier, 80% Capacity Factor	ACATIGFC4B	1		Catalytic	≈24.6	80		85	98%	200				
		PCATIGFC4B	3				85								
Case 5	85% Capacity Factor	ACATIGFC5B	1												
		PCATIGFC5B	3												
Case 6	BOP Enhancements 20% ASU Power Reduction, Inverter Efficiency 98%, Stack Cost Reduction	ACATIGFC6B	1				85		97.5						
		PCATIGFC6B	3												
Case 6V	Case 6 with VGR	ACATIGFC6BV	1						97.5						

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit ES-5. Pathway 1 – Atmospheric IGFC Plants with CCS Results Summary

Case	Case 0	Case 1	Case 2	Case 2V	Case 3	Case 3V	Case 4	Case 5	Case 6	Case 6V
Gasifier Technology	Conventional				Enhanced		Catalytic			
SOFC Degradation Rate (%/1000 h)	0.2									
Fuel Utilization (%)	80	85		97.5	85	97.5	85			97.5
Capacity Factor (%)	80		85				80	85		
Inverter Efficiency (%)	97								98	
SOFC Stack Cost (\$/kW)	225								200	
Performance										
Gross Power (MWe)	804	799	799	795	790	779	713	713	711	715
Auxiliary Loads (MWe)	154	149	149	145	140	129	62	62	60	65
Net Power (MWe)	650	650	650	650	650	650	650	650	650	650
Coal Flowrate (lb/h)	442,400	432,950	432,950	403,450	417,800	378,800	327,800	327,800	324,075	311,475
Net Electric Efficiency, HHV (%)	43.0	43.9	43.9	47.1	45.5	50.2	58.0	58.0	58.7	61.1
Net Plant Heat Rate, HHV (Btu/kWh)	7,942	7,774	7,774	7,242	7,497	6,800	5,881	5,881	5,815	5,588
CO ₂ Capture Rate (%)	98.8	98.7	98.7	96.1	98.7	95.9	98.6	98.6	98.6	94.3
CO ₂ Captured (tonnes per year)	3,546,910	3,471,017	3,687,955	3,345,812	3,558,637	3,134,435	2,513,636	2,670,739	2,640,389	2,496,945
CO ₂ Emissions (lb/MWhgross)	20.6	20.9	20.9	50.0	20.3	49.0	16.1	16.1	15.9	246.8
CO ₂ Emissions (lb/MWhnet)	25.4	25.6	25.6	61.2	24.6	58.7	17.6	17.6	17.4	271.5
Raw Water Consumption (gpm/MWnet)	2.36	2.06	2.06	1.73	1.74	1.51	1.50	1.50	1.48	1.06
Cost										
Total Plant Cost (TPC) (1000\$)	2,500,963	2,469,563	2,469,563	2,370,701	2,421,613	2,276,869	1,556,032	1,556,032	1,532,178	1,496,126
Total Overnight Cost (TOC) (1000\$)	3,094,997	3,055,481	3,055,364	2,942,472	2,995,692	2,825,595	1,922,481	1,922,408	1,893,174	1,848,016
Total As-Spent Cost (TASC) (1000\$)	3,571,627	3,526,025	3,525,890	3,395,613	3,457,029	3,260,737	2,218,544	2,218,459	2,184,723	2,132,610
Levelized Cost of Electricity (\$/MWh)										
Variable Costs	27.3	26.8	26.4	25.4	25.6	24.1	20.3	20.0	19.4	18.8
Fuel Costs	17.7	17.3	17.3	16.1	16.7	15.1	13.1	13.1	13.0	12.4
Variable O&M Costs	9.6	9.5	9.1	9.2	9.0	9.0	7.2	6.9	6.5	6.4
Fixed O&M Costs	14.8	14.6	13.7	13.2	13.5	12.8	9.7	9.2	9.0	8.9
Capital Costs	55.4	54.8	51.5	49.6	50.5	47.6	34.4	32.4	31.9	31.1
Total LCOE (excluding T&S)	97.5	96.2	91.7	88.2	89.6	84.5	64.4	61.5	60.4	58.8
T&S	7.1	6.9	6.9	6.3	6.7	5.9	5.0	5.0	4.9	4.7
Total LCOE (including T&S)	104.6	103.1	98.6	94.5	96.3	90.4	69.4	66.5	65.3	63.5

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit ES-6. Pathway 1 – Atmospheric IGFC Plants without CCS Results Summary

Case	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology	Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)	0.2						
Fuel Utilization (%)	80	85					
Capacity Factor (%)	80		85		80	85	
Inverter Efficiency (%)	97						98
SOFC Stack Cost (\$/kW)	225						200
Performance							
Gross Power (MWe)	728	728	728	720	681	681	680
Auxiliary Loads (MWe)	79	78	78	70	31	31	29
Net Power (MWe)	650	650	650	650	650	650	650
Coal Flowrate (lb/h)	400,525	394,300	394,300	382,600	315,012	315,012	311,700
Net Electric Efficiency, HHV (%)	47.5	48.2	48.2	49.7	60.4	60.4	61.0
Net Plant Heat Rate, HHV (Btu/kWh)	7,190	7,082	7,082	6,870	5,654	5,654	5,593
CO ₂ Capture Rate (%)	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Captured (tonnes per year)	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Emissions (lb/MWhgross)	1,278.7	1,260.4	1,260.4	1,236.5	1,028.5	1,028.5	1,019.8
CO ₂ Emissions (lb/MWhnet)	1,433.3	1,412.2	1,412.2	1,369.7	1,077.5	1,077.5	1,066.0
Raw Water Consumption (gpm/MWhnet)	1.47	1.23	1.23	0.91	1.07	1.07	1.05
Cost							
Total Plant Cost (TPC) (1000\$)	2,211,045	2,189,424	2,189,424	2,156,627	1,455,125	1,455,125	1,426,953
Total Overnight Cost (TOC) (1000\$)	2,754,017	2,723,013	2,722,897	2,682,740	1,799,159	1,799,089	1,764,708
Total As-Spent Cost (TASC) (1000\$)	3,178,135	3,142,357	3,142,224	3,095,882	2,076,230	2,076,149	2,036,473
Levelized Cost of Electricity (\$/MWh)							
Variable Costs	25.5	25.0	24.7	24.1	19.5	19.2	18.7
Fuel Costs	16.0	15.8	15.8	15.3	12.6	12.6	12.5
Variable O&M Costs	9.5	9.3	8.9	8.8	6.9	6.6	6.2
Fixed O&M Costs	13.2	13.1	12.3	12.2	9.2	8.7	8.5
Capital Costs	49.3	48.8	45.9	45.2	32.2	30.3	29.7
Total LCOE (excluding T&S)	88.1	87.0	82.9	81.5	60.9	58.2	56.9
T&S	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total LCOE (including T&S)	88.1	87.0	82.9	81.5	60.9	58.2	56.9

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit ES-7. Pathway 2 – Pressurized IGFC Plants with CCS Results Summary

Case	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology	Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)	0.2						
Fuel Utilization (%)	80	85					
Capacity Factor (%)	80		85		80	85	
Inverter Efficiency (%)	97						98
SOFC Stack Cost (\$/kW)	225						200
Performance							
Gross Power (MWe)	1,065	1,109	1,109	1,042	784	784	781
Auxiliary Loads (MWe)	415	459	459	393	134	134	131
Net Power (MWe)	650	650	650	649	650	650	650
Coal Flowrate (lb/h)	492,800	492,200	492,200	455,195	332,700	332,700	328,775
Net Electric Efficiency, HHV (%)	38.6	38.6	38.6	41.7	57.1	57.1	57.8
Net Plant Heat Rate, HHV (Btu/kWh)	8,844	8,831	8,831	8,185	5,971	5,971	5,901
CO ₂ Capture Rate (%)	98.8	98.7	98.7	98.7	98.6	98.6	98.6
CO ₂ Captured (tonnes per year)	3,950,953	3,945,823	4,192,437	3,876,712	2,551,459	2,710,925	2,678,833
CO ₂ Emissions (lb/MWhgross)	20.0	20.5	20.5	19.6	15.6	15.6	15.4
CO ₂ Emissions (lb/MWhnet)	32.9	35.0	35.0	31.5	18.7	18.7	18.5
Raw Water Consumption (gpm/MWhnet)	2.15	1.79	1.79	1.51	1.37	1.37	1.35
Cost							
Total Plant Cost (TPC) (1000\$)	2,725,410	2,745,132	2,745,132	2,627,353	1,631,795	1,631,795	1,600,867
Total Overnight Cost (TOC) (1000\$)	3,378,063	3,402,160	3,402,027	3,253,949	2,015,238	2,015,160	1,977,372
Total As-Spent Cost (TASC) (1000\$)	3,898,284	3,926,092	3,925,939	3,755,057	2,325,585	2,325,495	2,281,888
Levelized Cost of Electricity (\$/MWh)							
Variable Costs	29.8	29.8	29.4	27.6	20.5	20.2	19.6
Fuel Costs	19.7	19.7	19.7	18.2	13.3	13.3	13.1
Variable O&M Costs	10.1	10.1	9.7	9.3	7.2	6.9	6.5
Fixed O&M Costs	16.1	16.2	15.2	14.7	10.2	9.6	9.4
Capital Costs	60.8	61.3	57.7	55.3	36.2	34.0	33.4
LCOE (excluding T&S)	106.7	107.3	102.3	97.5	66.8	63.8	62.4
T&S	7.9	7.9	7.9	7.3	5.1	5.1	5.0
LCOE (including T&S)	114.6	115.1	110.1	104.8	71.9	68.8	67.5

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit ES-8. Pathway 2 – Pressurized IGFC Plants without CCS Results Summary

Case	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology	Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)	0.2						
Fuel Utilization (%)	80	85					
Capacity Factor (%)	80		85		80	85	
Inverter Efficiency (%)	97						98
SOFC Stack Cost (\$/kW)	225						200
Performance							
Gross Power (MWe)	902	915	915	887	739	739	737
Auxiliary Loads (MWe)	252	264	264	237	89	89	87
Net Power (MWe)	650	650	650	650	650	650	650
Coal Flowrate (lb/h)	353,060	325,005	325,005	318,400	281,360	281,360	278,470
Net Electric Efficiency, HHV (%)	53.0	52.6	52.6	54.6	67.6	67.6	68.3
Net Plant Heat Rate, HHV (Btu/kWh)	6439	6487	6487	6247	5048	5048	4997
CO ₂ Capture Rate (%)	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Captured (tonnes per year)	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Emissions (lb/MWhgross)	920.1	878.7	878.7	878.9	847.5	847.5	841.1
CO ₂ Emissions (lb/MWhnet)	1277.0	1236.0	1236.0	1198.8	962.9	962.9	953.1
Raw Water Consumption (gpm/MWhnet)	0.72	0.70	0.70	0.55	0.52	0.52	0.52
Cost							
Total Plant Cost (TPC) (1000\$)	2,316,541	2,321,148	2,321,148	2,246,055	1,414,661	1,414,661	1,484,549
Total Overnight Cost (TOC) (1000\$)	2,833,529	2,836,867	2,836,778	2,745,191	1,749,245	1,749,176	1,832,967
Total As-Spent Cost (TASC) (1000\$)	3,269,892	3,273,745	3,273,642	3,167,951	2,018,628	2,018,549	2,115,244
Levelized Cost of Electricity (\$/MWh)							
Variable Costs	21.9	23.2	22.8	21.8	17.8	17.4	17.3
Fuel Costs	14.6	15.9	15.9	15.1	11.2	11.2	11.1
Variable O&M Costs	7.3	7.3	6.9	6.7	6.5	6.2	6.2
Fixed O&M Costs	13.9	14.0	13.1	12.8	9.1	8.5	8.5
Capital Costs	51.2	51.3	48.3	46.7	31.6	29.8	29.6
Total LCOE (excluding T&S)	87.0	88.4	84.2	81.3	58.4	55.7	55.4
T&S	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total LCOE (including T&S)	87.0	88.4	84.2	81.3	58.4	55.7	55.4

DISCUSSION OF PATHWAY RESULTS

In this section, the impact of the technological developments is discussed, along with the cost reduction assumptions considered in the present pathway studies on the performance and cost of an IGFC plant. The results from the various cases are consolidated to provide guidance to DOE-FECM's fuel cell R&D program. Further, the IGFC system performance and costs are compared with conventional heat-engine technologies.

IGFC System Performance and Efficiency

The efficiencies of the various IGFC plants considered in the two pathways are shown in Exhibit ES-9. For an atmospheric IGFC plant without CCS, the higher heating value (HHV) efficiency varies from a value of 47.5 percent for the reference plant to a value of 61.0 percent for the Case 6 advanced plant with a catalytic gasifier. Inclusion of CCS imposes an efficiency penalty of ≈ 3.5 –4.0 percentage points for atmospheric systems. The VGR configuration significantly reduces this penalty to nearly a percent in the case of the conventional gasifier, while the VGR systems with CCS have even a slight efficiency advantage over systems without CCS. Indeed, the highest efficiency of all the CCS systems considered, including pressurized systems, is achieved by the catalytic VGR system. The VGR concept enables operating at a fuel utilization of 97.5 percent, while the in-stack, single-pass utilization is maintained below 60 percent.

Pressurization generally results in an increase in efficiency over the atmospheric IGFC cases (without VGR), except for the conventional and enhanced gasifier cases with CCS, because of excessive process air required (and an increase in the air compression losses) in such cases in the absence of cathode gas recirculation. The pressurized conventional and enhanced gasifier cases without CCS also include an efficiency advantage due to the injection of natural gas into the combustor to enhance cathode expander inlet temperatures.

As expected with the underlying SOFC technology advantages, the IGFC efficiencies are significantly higher than the values for comparable conventional technologies, including the supercritical (SC) pulverized coal (PC) and supercritical CO₂ (sCO₂) indirect coal plants with CCS shown in the Exhibit ES-9. The IGFC plants start to close in on the efficiency of even an H-class NGCC system with enhanced gasification technology, while the catalytic gasifier-based IGFC systems result in efficiencies that are 15–20 percentage points higher relative to conventional systems, depending on the inclusion of CCS.

The catalyst gasifier technology has the highest impact (≈ 9 –15 percentage points) on the plant efficiency with or without CCS, while the VGR concept—which requires minimal advances in technology—is the next highest efficiency booster, resulting in ≈ 2.4 –4.7 percentage points improvement in performance as indicated by the waterfall plots in Exhibit ES-10 and Exhibit ES-11. The BOP enhancements resulted only in modest gains, since the ASU parasitic is lower for the Catalytic Gasifier cases (Cases 4–6 for all the configurations), which, due to higher plant efficiency, have significantly lower O₂ requirements relative to the conventional and enhanced gasifier plants.

Exhibit ES-9. HHV Efficiency of the Pathway IGFC Plants

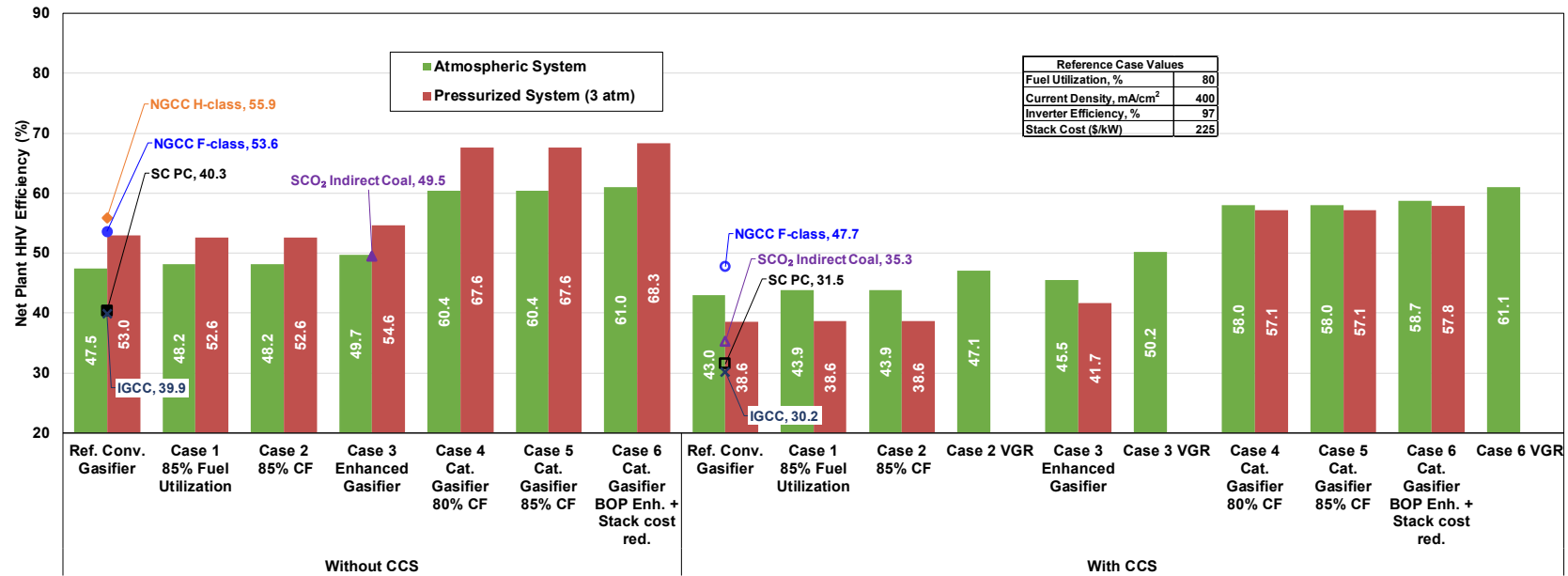


Exhibit ES-10. Technology Developments and Efficiency Improvements for IGFC Plants without CCS

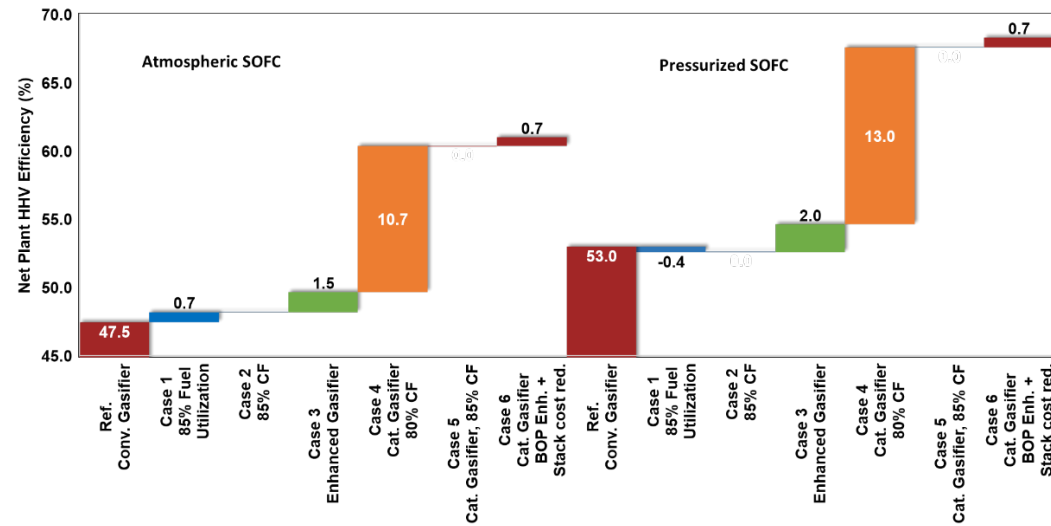
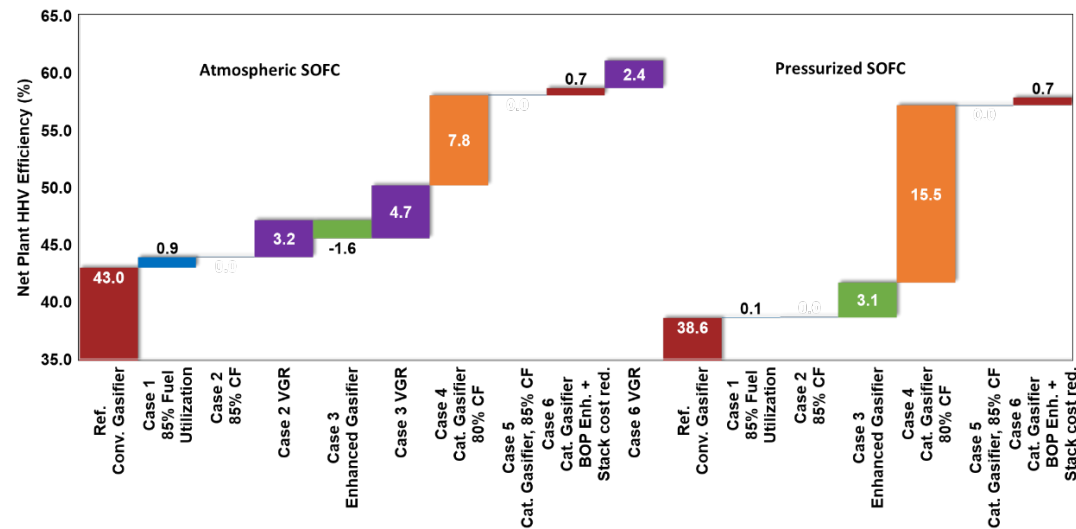


Exhibit ES-11. Technology developments and efficiency improvements for IGFC plants with CCS



IGFC Plant Costs

The catalytic gasifier cases result in the lowest total plant cost (TPC) for the IGFC plants with and without CCS, as shown in Exhibit ES-12, attributable to the high CH₄ content in the syngas, which significantly reduces airflow requirements. The pressurized configurations generally result in a higher TPC relative to the atmospheric systems primarily due to the increased enclosure (pressure vessel) costs and the inclusion of cathode expander costs, which seem to negate any efficiency advantages of pressurization on the capital costs. The TPC of IGFC plants with conventional as well as enhanced gasification technology are generally higher than conventional technologies. A reduction in costs of conventional gasifiers, which dominate the plant costs for these cases, would significantly affect the relative comparison. The reduction in TPC for catalytic gasifier IGFC systems is mainly due to the projected lower costs for these types of gasifiers operating at temperatures lower than the conventional technologies.

The levelized cost of electricity (LCOE) (without transport and storage [T&S]) of the reference atmospheric conventional gasifier IGFC plant with CCS is lower (by \$8/MWh) than the LCOE of SC PC plants with CCS and is comparable to the LCOE of the sCO₂ indirect coal plant, also with CCS, as shown in Exhibit ES-13.^{b, c} All the other IGFC plants with CCS have a lower LCOE than the coal-based conventional technologies with the catalytic gasifier IGFC plants, indicating a significant LCOE advantage even relative to F-class NGCC plants with CCS. In general, the LCOE of pressurized systems are slightly higher than the corresponding atmospheric systems. The VGR configuration substantially mitigates the LCOE penalty of CCS (by nearly half).

On the other hand, all the IGFC plants without CCS have higher costs relative to conventional coal-based technologies, mainly attributable to the higher capital costs of the gasifier. Indeed, the dominance of the capital portion of the plant is reflected in the breakdown of the LCOE with and without T&S shown in Exhibit ES-14 and Exhibit ES-15, respectively. A 10-percent reduction in conventional gasifier costs would reduce the LCOE by nearly five percent for the reference case.

The difference in LCOEs between an IGFC plant with and without CCS is significantly smaller than the corresponding penalties for conventional technologies. The SOFC is, essentially, an oxyfuel reactor and along with the sealed design generally used to separate the air and fuel, it forms a highly effective inherent carbon separator (that produces power); it produces a concentrated CO₂ effluent that is ready for CCS with minimal incremental costs. This underscores the leading role played by SOFC-based systems in meeting NETL's environmental vision.

^b The LCOE for H-class have been estimated to be ≈\$36/MWh in the study by Uysal. [35] However, these were not included on the charts since the costs are based on a 1 GW plant capacity and the financial parameters used are different than the baseline studies.

^c The LCOE for the sCO₂ indirect coal plant from the study of White et al. [36] has been adjusted to use the latest financial parameters.

Exhibit ES-12. Comparison of the TPC of the IGFC Plants

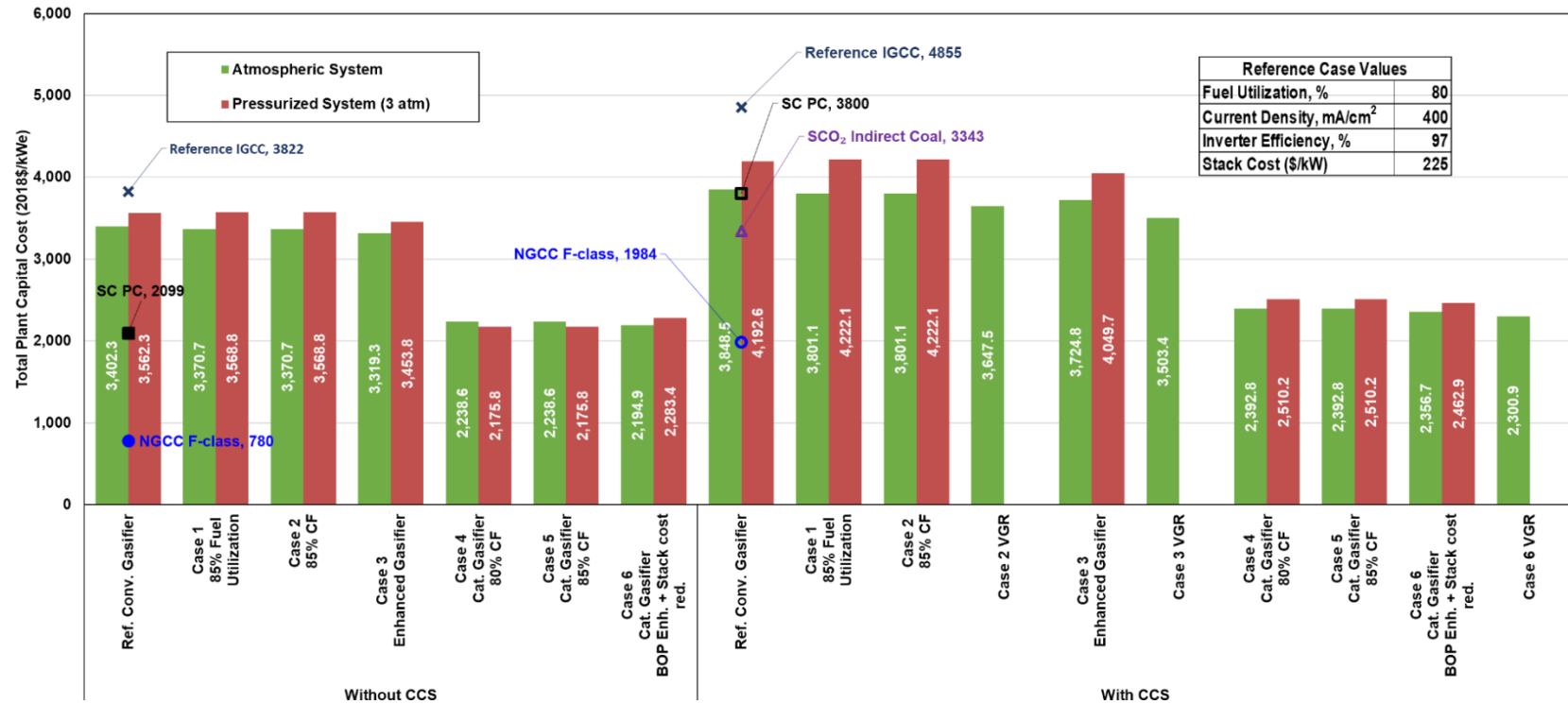
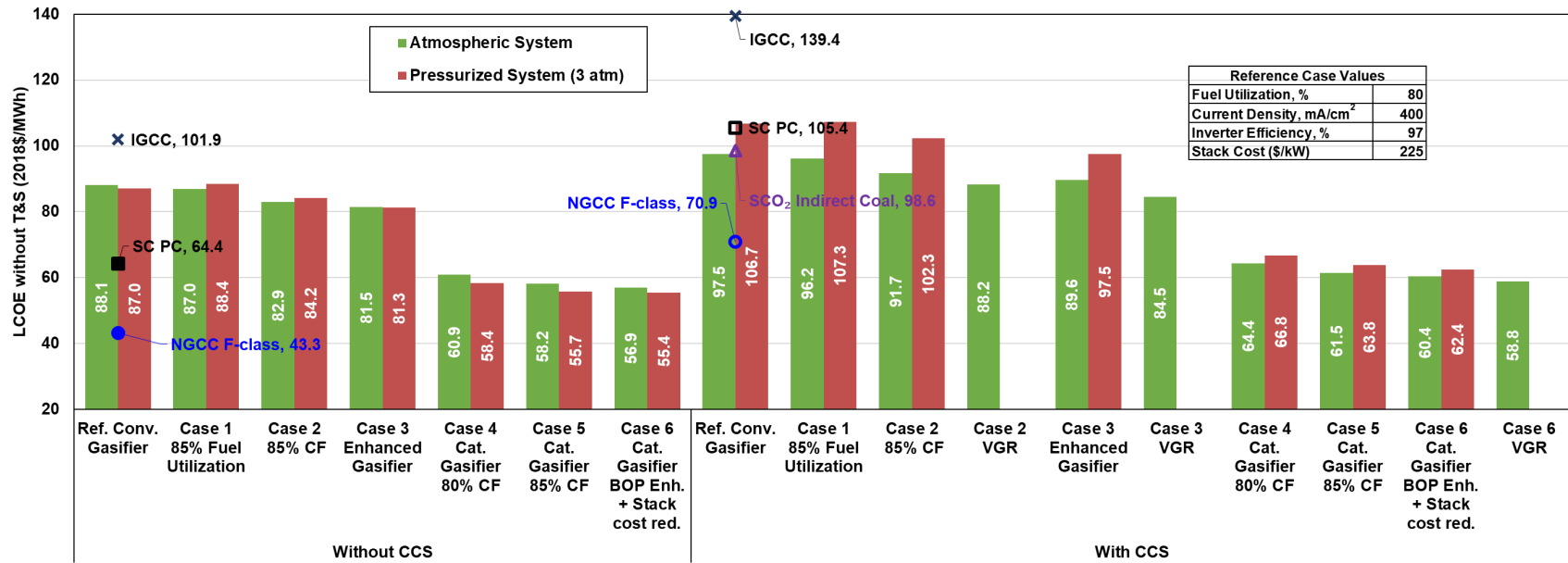


Exhibit ES-13. Comparison of the LCOE (without T&S) of the IGFC Plants



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit ES-14. LCOE Breakdown (with T&S) of the IGFC Plants

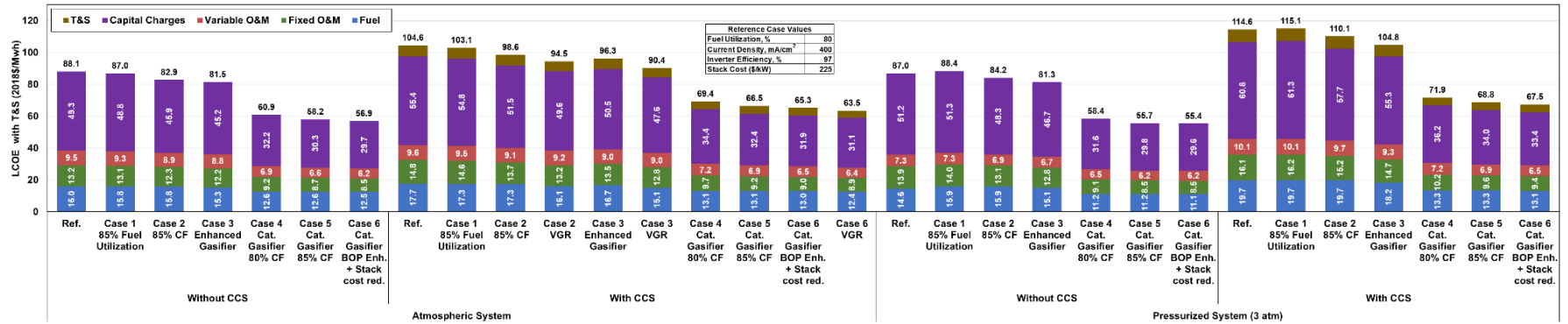
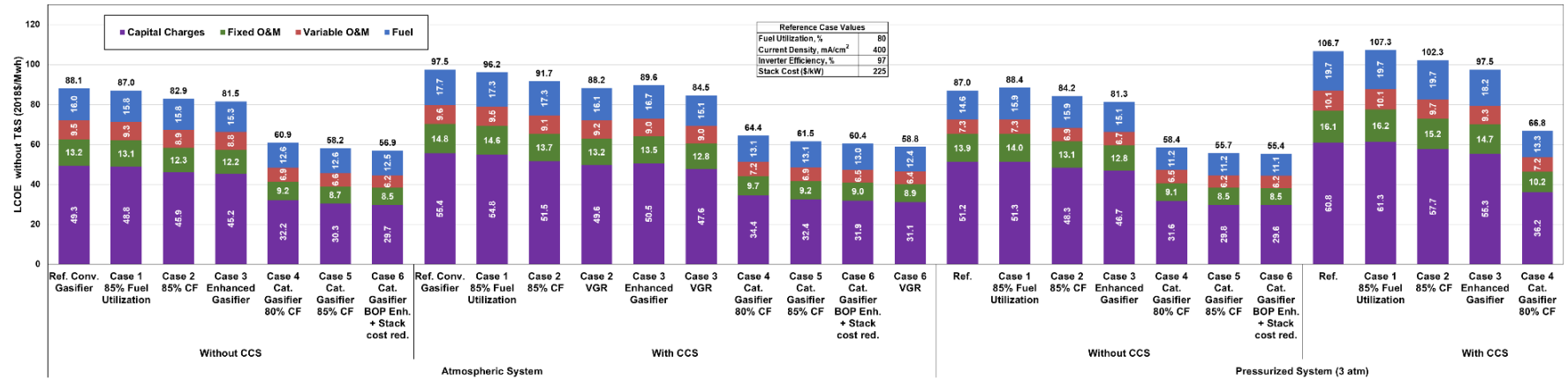


Exhibit ES-15. LCOE Breakdown (without T&S) of the IGFC Plants



The progression of LCOE with pathway technology developments, cost reductions, and increased availability are shown in Exhibit ES-17 for the IGFC plants without CCS and in Exhibit ES-17 for the IGFC plants with CCS, respectively. In the case of IGFC plants with CCS, the largest reduction in LCOE, $\approx \$31/\text{MWh}$, is associated with the introduction of catalytic gasifier technology. An increase in the system capacity factor and the VGR configuration yield the next significant LCOE reductions, ranging between $\approx \$2\text{--}5/\text{MWh}$, depending on the configuration. The introduction of catalytic gasification technology and increased capacity factor are factors that significantly affect the LCOE of the IGFC systems without CCS.

The pathways, which represent practical and realizable steps consistent with the fuel cell program, lead to an IGFC system with capture that has a significantly low LCOE relative to conventional heat-engine based technologies with CCS. The pathways also lead to IGFC plants without CCS that are economically competitive with F-class NGCC plants. Although pressurization does not appear to have a significant advantage in the cases analyzed here, pressurized configurations could be found to be attractive as a hybrid system where operational flexibility aspects may be attractive.

Exhibit ES-16. Progression of Plant LCOE (without T&S) along the IGFC Pathway without CCS

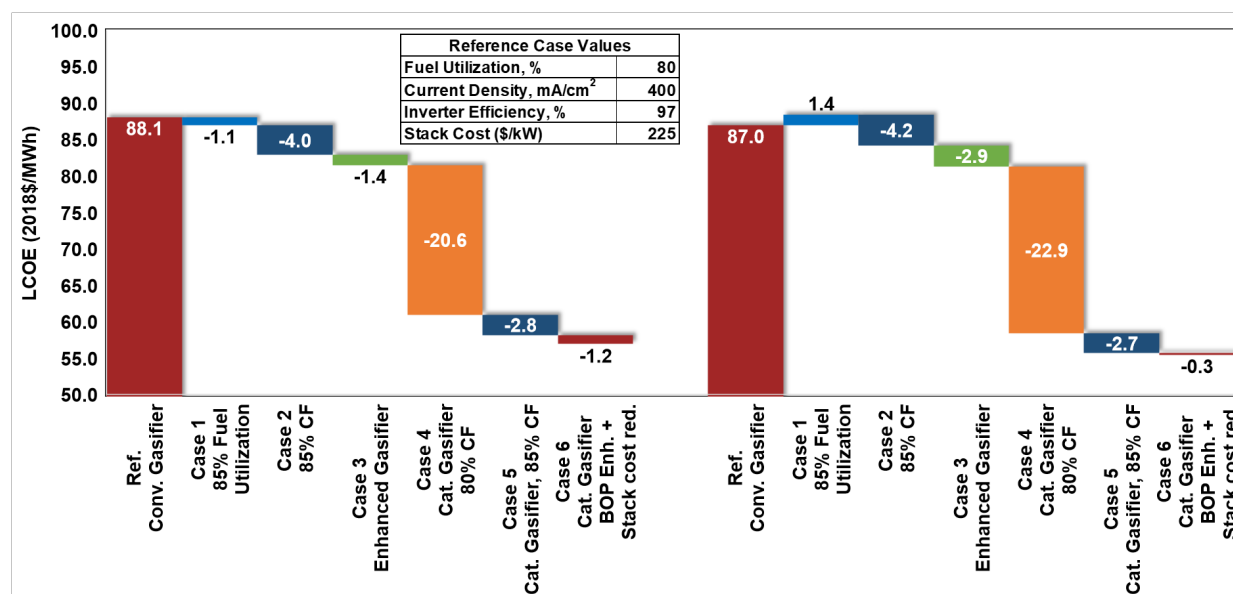
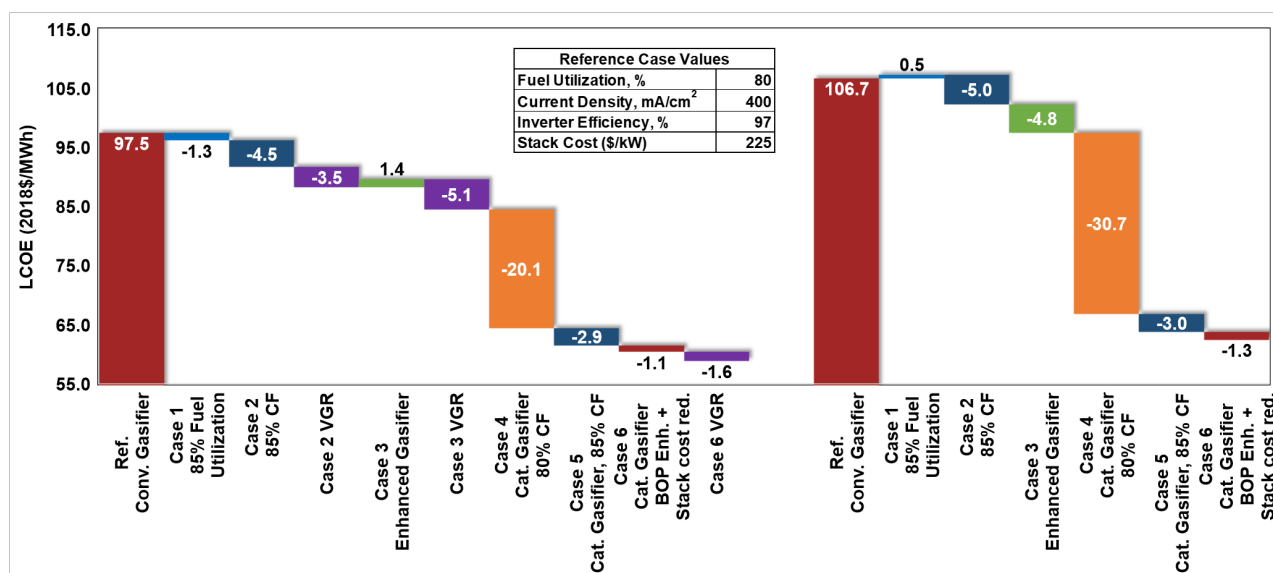


Exhibit ES-17. Progression of Plant LCOE (without T&S) along the IGFC Pathway with CCS

Cost of CO₂ Captured

The breakeven CO₂ sales price (the price of CO₂ that is required to pay for the difference in LCOE between a plant with and without CCS) for the pathway IGFC cases is shown in Exhibit ES-18. All the atmospheric IGFC plants and advanced pressurized plants considered herein have a breakeven CO₂ sales price that is well below the \$40/tonne of CO₂ generally considered to be an achievable selling price for pure CO₂ for EOR purposes. The atmospheric IGFC systems with the VGR and the atmospheric systems with catalytic gasifiers result in breakeven prices that are even below \$10/tonne, which could make these systems highly competitive with other CO₂ sources.

Water Consumption

The IGFC plants, as shown in Exhibit ES-19, result in significantly lower water consumption compared to SC PC and NGCC plants. The IGFC plant water consumption is dominated by the gasifier and the cooling water makeup requirements of the steam cycle.

Exhibit ES-18. Breakeven Sale Price of CO₂ Captured for the Various IGFC Plants

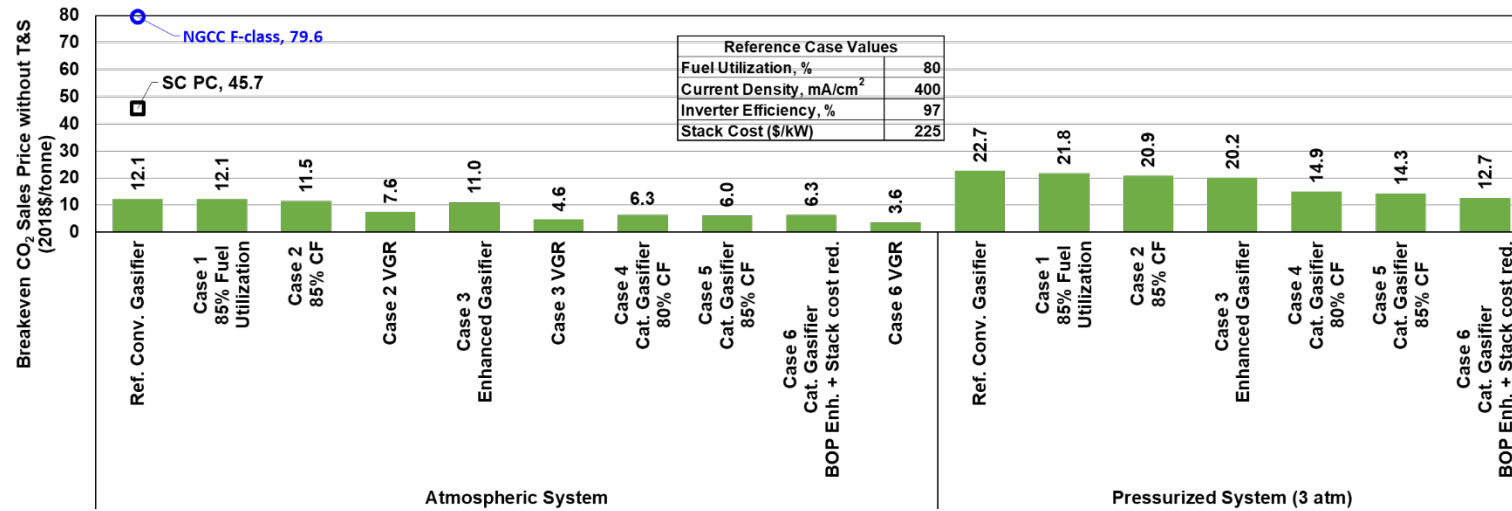
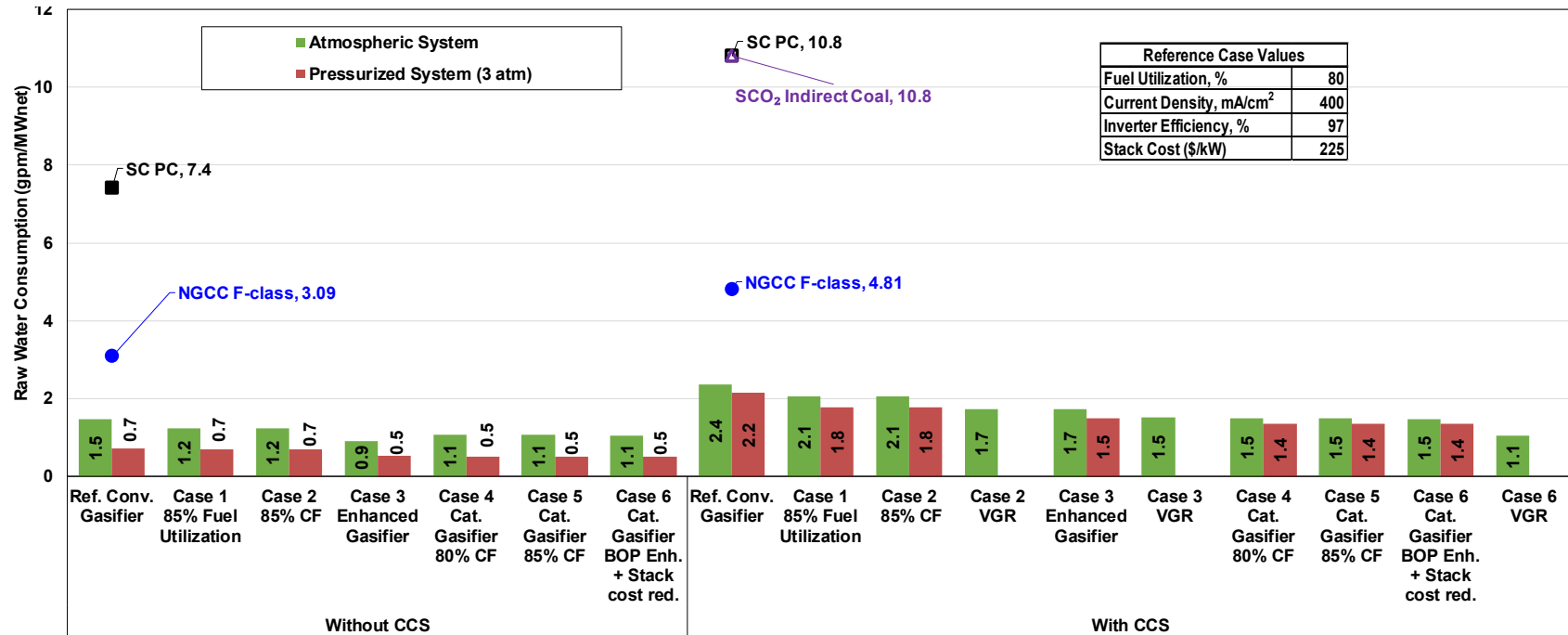


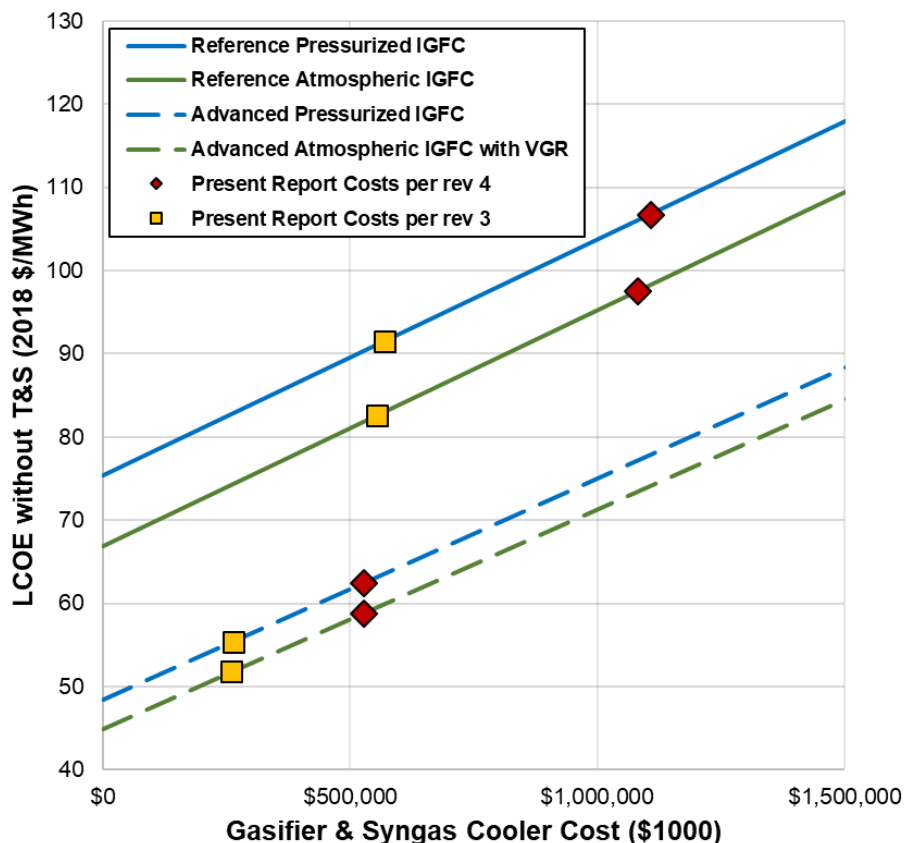
Exhibit ES-19. Water Consumption Variation for the Various IGFC plants



Effect of Gasifier Costs

The LCOE of the reference IGFC plant (Conventional Gasifier Case 0) and the best advanced IGFC plant (Case 6 with VGR) with CCS is shown in Exhibit ES-20 as a function of the gasifier capital costs (gasifier and syngas cooler accounts). The present study IGFC system costs based on Revision 4 and Revision 3 of the baseline report are also shown in the exhibit. The conventional gasifier costs in Revision 4 of the baseline report are nearly double the costs reported in the previous revision, which corresponds to $\approx \$10$ – $\$15$ /MWh increase in LCOE for the IGFC systems.

Exhibit ES-20. Influence of Gasifier Capital Costs on the LCOE (excl. T&S) of IGFC Plants with CCS



Effect of Natural Gas Price

The LCOE of the reference IGFC plant (Atmospheric Conventional Gasifier Case 0) and the best advanced IGFC plant (Case 6 with VGR) with CCS is shown in Exhibit ES-21 as a function of natural gas price. The variation of the LCOE of a conventional F-class NGCC plant with CCS is also shown, along with the LCOE of an SC PC plant, which does not depend directly on the price of natural gas. With CCS, the reference IGFC plant costs become competitive relative to NGCC systems with CCS above a natural gas price of $\approx \$8$ /MMBtu. In plants without CCS, natural gas prices above ≈ 12 /MMBtu are required, as shown in

Exhibit ES-22, to make the IGFC plants competitive relative to conventional NGCC systems.

Exhibit ES-21. Influence of Natural Gas Price on the LCOE (excl. T&S) of IGFC Plants with CCS

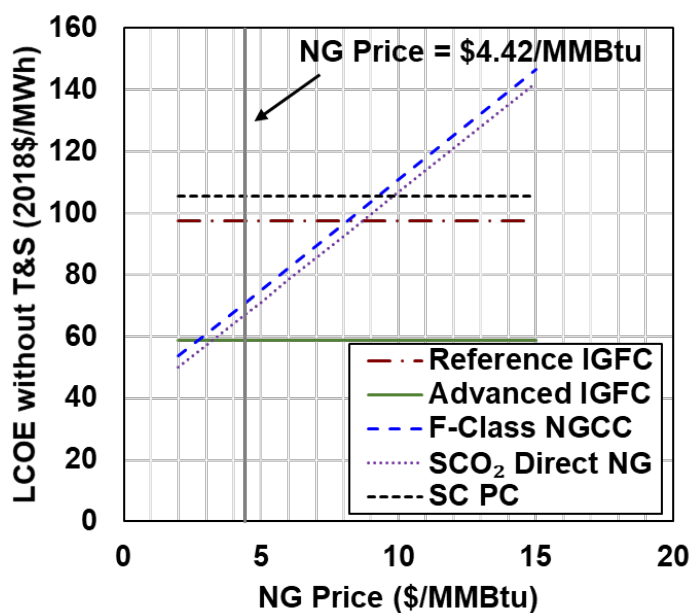
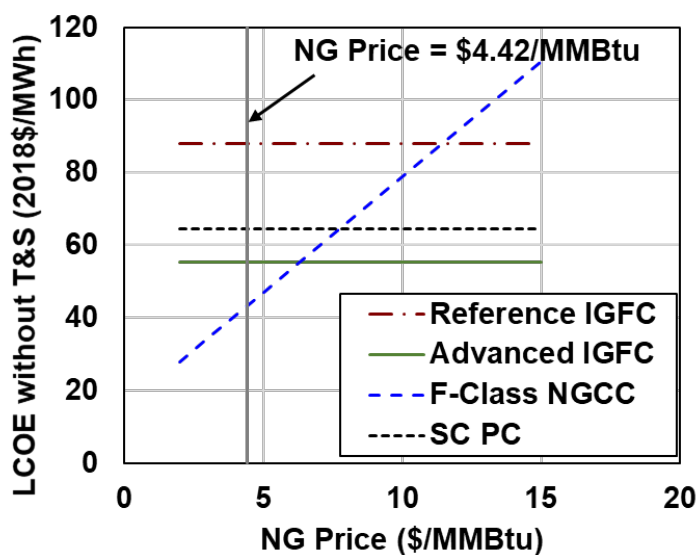


Exhibit ES-22. Influence of Natural Gas Price on the LCOE (excl. T&S) of IGFC Plants without CCS



KEY CONCLUSIONS

The results of a pathway study for coal based, IGFC power systems with and without CCS are presented in this report. Two pathway scenarios are considered, based on the operating pressure of the SOFC stacks. Three gasification technologies, including a conventional technology, differing in the syngas CH_4 content, which has a significant effect on the IGFC system performance, were considered. The performance and cost benefits were estimated for a series of projected gains made through the development of advances in the component technologies or improvements in plant costs and availability. The IGFC systems with CCS included a CO_2 purification unit that met EOR specifications for the product CO_2 stream.

The results indicate that

- IGFC systems have the potential to result in HHV efficiencies that exceed 60 percent even with CCS. The present pathway resulted in a system with efficiency as high as 68.1 percent without capture.
- While the IGFC plants with conventional technology have a significant cost advantage over coal-based conventional technologies, the IGFC economic competitiveness relative to natural gas-based heat engines (at current natural gas prices) are achievable only with significant advances and considerable cost reductions in gasification technology. In particular, the development of gasifiers with higher methane content in the syngas is desirable.
- The difference in LCOE between an IGFC plant with and without CCS is significantly smaller than the corresponding penalties for conventional technologies. The SOFC is, essentially, an oxyfuel reactor and, along with the sealed design generally used to separate the air and fuel, it forms a highly effective inherent carbon separator (that produces power); and it produces a concentrated CO_2 effluent that is ready for CCS with minimal incremental costs. Indeed, breakeven values of CO_2 prices below \$10/tonne are achievable.
- Gasifier cost reductions has a significant impact on the LCOE of the IGFC systems.
- The VGR concept has the potential to significantly reduce the LCOE in the short-term even with conventional technologies.
- Pressurization does not appear to have a significant advantage in the cases analyzed here. However, pressurized configurations could be found to be attractive as a hybrid system where operational flexibility aspects may be attractive.
- The IGFC consumes 50 percent less raw water than even the water economical NGCC system with CCS.
- A pressurized SOFC provides no cost benefit over systems operating with an atmospheric-pressure SOFC. However, the IGFC plant configuration and operating conditions selected for the pressurized SOFC evaluation in this study have not been optimized and, thus, there are opportunities for further benefit. In particular, the cost of

the pressure vessel is an important parameter that needs further refinement. This IGFC configuration requires development of the pressurized-SOFC technology.

The VGR configuration was not optimized in the present study. Additional performance and cost advantages can be realized with this configuration through rigorous optimization. There are other technological innovations that might also benefit the IGFC power plant performance and cost, such as humid gas cleaning, the ion transport membrane technology for oxygen separation incorporating integration with the pressurized SOFC cathode air compressor, and shock-wave CO₂ compression. Further, in the near-term, natural gas injection as a proxy for a catalytic gasifier could greatly increase the performance and cost potential of the IGFC plant using conventional or enhanced-conventional coal gasification. These technology advances will be explored in future IGFC pathway evaluations.

1 INTRODUCTION

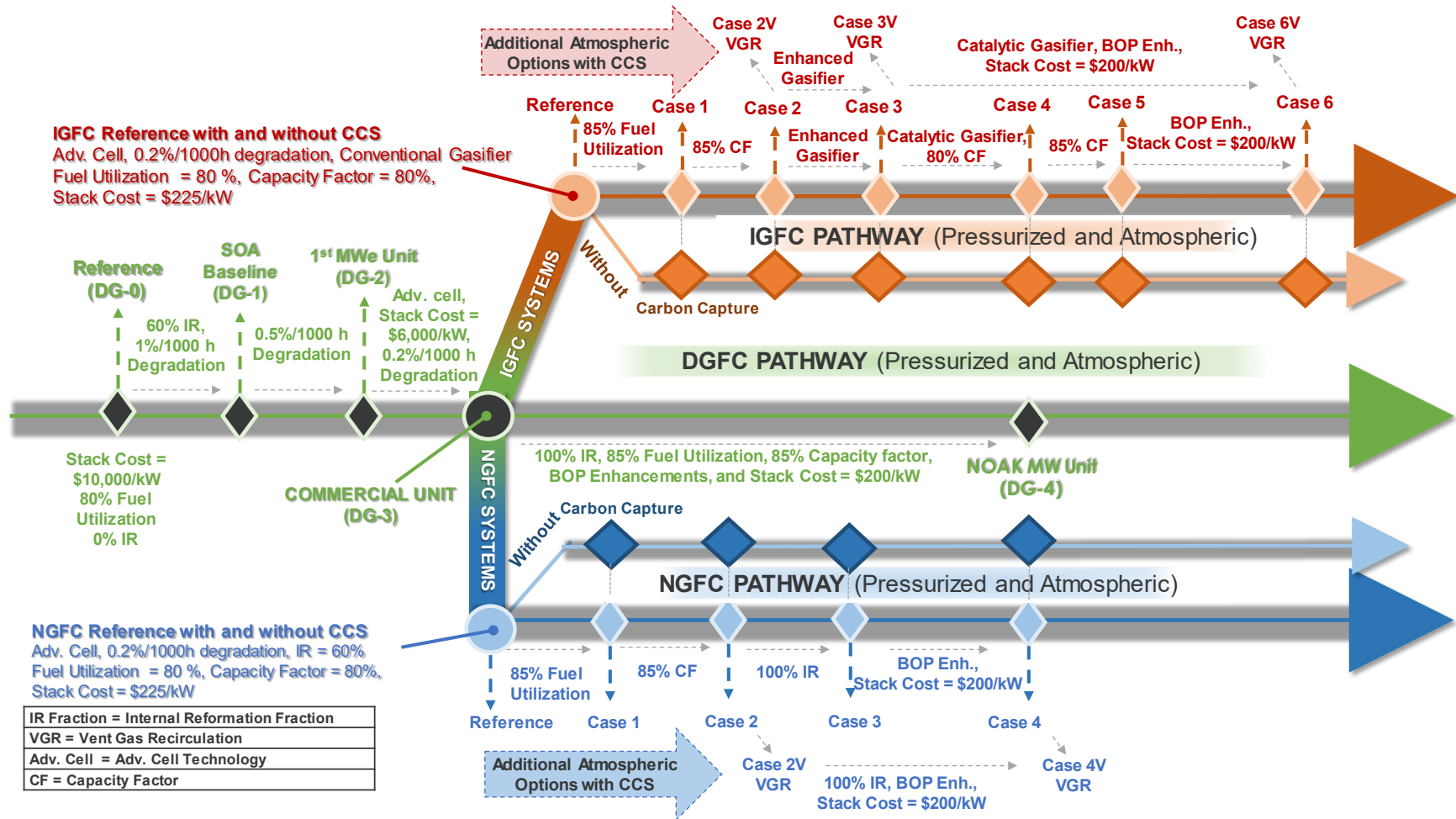
The United States (U.S.) Department of Energy (DOE) Office of Fossil Energy and Carbon Management (FECM) and the National Energy Technology Laboratory (NETL) have been pursuing the development of solid oxide fuel cell (SOFC) technology to enable future power generation systems that are consistent with the cornerstones of the DOE mission—to ensure America's security and prosperity by addressing its energy and environmental challenges through transformative science and technology solutions. The U.S. DOE FECM SOFC Program is currently focused on the development of low-cost, highly efficient, and reliable SOFC power systems. [1]

The present study describes the results of a pathway study for coal-based integrated gasification fuel cell (IGFC) power systems with and without carbon capture and storage (CCS). The objective of the study is to provide targeted research and development (R&D) guidance to the FECM SOFC Program.

The IGFC power plant is analogous to an integrated gasification combined cycle (IGCC) power plant, except the gas turbine power island is replaced with a solid oxide fuel cell (SOFC) system. The SOFC provides for high electric efficiencies associated with the nearly reversible electrochemical conversion of syngas chemical potential to electric power, as compared to a conventional Carnot-driven heat engine. In addition, the clean oxy-reaction of syngas in the sealed fuel cell system renders itself readily available for CCS, requiring only an oxy-combustor downstream of the fuel cell to react the fuel that is not utilized electrochemically (electrochemical utilization of fuels typically varies between ≈ 75 –90 percent for current fuel cell technology, due to practical considerations). The heat rejected by the fuel cell system can be recovered further in a combination of Brayton and Rankine cycles, depending on whether the fuel cell system is operating at elevated or atmospheric pressures.

A general schematic block flow diagram of the IGFC system with carbon capture is shown in Exhibit 1-1. The syngas exiting the gasifier goes through a cleaning step before expansion to SOFC operating pressure. Heat recovered from cooling the raw syngas, the hot oxy-combustor exhaust gas, and the hot vitiated air-exhaust gas from the SOFC system, along with any process heat generated during syngas cleaning and sulfur extraction, is sent to a heat recovery steam generator (HRSG), which, in addition to meeting all the process steam requirements of the system, supplies steam to the bottoming cycle. The cooled anode exhaust from the heat recovery steam generator (HRSG) is further purified to pipeline specifications in a CO₂ purification unit (CPU). In systems featuring a pressurized SOFC generator, a major portion of the compression work needed to supply air at pressure to the generator is recovered by expanding the SOFC cathode air exhaust gas to atmospheric pressure (as shown by the dotted lines in Exhibit 1-1). An air separation unit (ASU) supplies oxygen to the gasifier/reformer, to the sulfur extraction process (IGFC systems), and to the oxy-combustor to enable efficient CO₂ capture.

Exhibit 1-2. Solid Oxide Fuel Cells Pathway Studies



The SOFC technology and system characteristics (Exhibit 1-3) of the established commercial DGFC unit (as shown in the DGFC developmental roadmap¹ in Exhibit 1-2) was assumed as the basis for the reference IGFC system, which assumes the readiness of a conventional gasification technology, and features a gasifier with syngas methane (CH₄) content of roughly 6 mole percent.

Exhibit 1-3. Reference IGFC System Characteristics

SOFC Parameters	Values
SOFC Performance	Advanced Cell – 50% reduction in Ohmic and Polarization losses relative to the state-of-the-art technology (SOA)
Degradation Rate	0.2% per 1000 h
System Fuel Utilization	80%
Stack Cost	\$225/kWe
Inverter Efficiency	97%
Gasifier Configuration	Conventional
Capacity Factor	80%

Two parallel developmental scenarios—one with the SOFC operating at atmospheric pressure and one based on a pressurized SOFC operation—were considered as part of the IGFC pathways. From previous studies, which found insignificant cost benefits for pressures higher than three atmospheres, and pressurized SOFC systems that have been generally proposed and operated by industrial vendors, an SOFC operating pressure of three atmospheres was assumed in the present calculations. Two gasification technology improvements with differing syngas methane (CH₄) content, which has a significant effect on the IGFC system performance, were considered.

The specific IGFC pathway cases analyzed in this study are listed in Exhibit 1-4 and Exhibit 1-5 for systems without and with CCS, respectively. Components for each plant configuration are described in more detail in the corresponding report sections for each pathway. Case 1 represents an increase in fuel utilization from 80 to 85 percent based on the confidence gained from DGFC systems. Enhanced reliability of balance of plant (BOP) components, combined with optimized O&M, is reflected by an increase in capacity factor to 85 percent in Case 2. An enhanced gasifier, representing an incremental adaptation in gasifier technology to generate a syngas with higher methane content (which reduces SOFC process air requirements and has been projected to benefit the performance of the IGFC plant) relative to the conventional gasifier, is assumed for Case 3. A transformational catalytic gasification technology, which is particularly designed to produce a high methane content (≈ 30 mole percent) syngas with improved IGFC performance benefits, is assumed to be available in Case 4, albeit with an initial capacity factor of 80 percent. Improvement of the capacity factor of the Case 4 system to 85 percent is considered in Case 5, while Case 6 considers potential enhancements in ASU and

¹ The DGFC pathway will be reported elsewhere.

inverter technologies, along with a reduction in stack cost to \$200/kW. In the case of systems with CCS, an attractive high-efficiency vent gas recirculation (VGR) configuration, which enables fuel utilization values >95 percent, is also explored, with systems reflected by three gasifier technologies—Cases 2, 3, and 6.

Each IGFC case is designated by a numbering system—“XGIGFCYZ*” —where X = A or P depending on whether the SOFC is operating under atmospheric or pressurized conditions respectively, G = CONV/ENH/CAT depending on whether a conventional, enhanced, or catalytic gasifier technology is utilized, Y represents the case number, Z = B or A depending on whether the systems features CCS, and * is a wildcard letter used to designate special configurations such as the VGR concept.

The design and cost bases for this pathway study closely follow the bases applied in the NETL Bituminous Baseline (BB) report so that direct performance and cost comparisons can be made with the conventional fossil-fuel power plant results estimated in that report. [2] The results of the study represent the potential future benefits of IGFC technology development.

Performance and cost projections for the IGFC plants are compared with conventional heat-engine based technologies.

The balance of this report is organized as follows:

- Section 2 provides the basis for the technical and cost evaluations.
- Section 3 describes the major plant components applied throughout the case studies.
- Section 4 describes the plant simulations and presents the results for the atmospheric pathway, discussing atmospheric-pressure IGFC cases with and without CCS and their corresponding pathway parameters.
- Section 6 describes the pressurized IGFC plant simulations with and without CCS and considers their corresponding pathway parameters.
- Section 6 provides a discussion of all the pathway results and provides DOE with a basis to select the most appropriate development path for IGFC, and to measure and prioritize the contribution of its R&D program to future power systems technology.
- Section 7 presents major conclusions from the study.

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 1-4. IGFC Pathway Cases without CCS

Case Designation	Pathway Parameter	Case ID	SOFC Pressure (atm)	SOFC Technology	Gasifier Technology	Dry Syngas CH ₄ Content (%)	Capacity Factor (%)	Degradation (%/1000 h)	Fuel Utilization (%)	Inverter Efficiency	Stack Cost (\$/kW)								
Reference (Case 0)	Reference Without CCS	ACONVIGFC0A	1	Advanced Cell	Conventional	≈5.8	80	0.2	80	97%	225								
		PCONVIGFC0A	3						85										
Case 1	85% Fuel Utilization	ACONVIGFC1A	1				85												
		PCONVIGFC1A	3																
Case 2	85% Capacity Factor	ACONVIGFC2A	1				Enhanced					≈10.8	85						
		PCONVIGFC2A	3																
Case 3	Enhanced Gasifier	AENHIGFC3A	1		Catalytic	≈24.6	80												
		PENHIGFC3A	3																
Case 4	Catalytic Gasifier, 80% Capacity Factor	ACATIGFC4A	1				80		85										
		PCATIGFC4A	3																
Case 5	85% Capacity Factor	ACATIGFC5A	1				85												
		PCATIGFC5A	3																
Case 6	BOP Enhancements 20% ASU Power Reduction, Inverter Efficiency 98%, Stack Cost Reduction	ACATIGFC6A	1				85			98%	200								
		PCATIGFC6A	3																

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 1-5. IGFC Pathway Cases with CCS

Case Designation	Pathway Parameter	Case ID	SOFC Pressure (atm)	SOFC Technology	Gasifier Technology	Dry Syngas CH ₄ Content (%)	Capacity Factor (%)	Degradation (%/1000 h)	Fuel Utilization (%)	Inverter Efficiency	Stack Cost (\$/kW)				
Reference (Case 0)	Reference Without CCS	ACONVIGFC0B	1	Advanced Cell	Conventional	≈5.8	80	0.2	80	97%	225				
		PCONVIGFC0B	3												
Case 1	85% Fuel Utilization	ACONVIGFC1B	1						85						
		PCONVIGFC1B	3												
Case 2	85% Capacity Factor	ACONVIGFC2B	1				85		97.5						
		PCONVIGFC2B	3												
Case 2 V	Case 2 with VGR	ACONVIGFC2BV	1						97.5						
Case 3	Enhanced Gasifier	AENHIGFC3B	1		Enhanced	≈10.8	85		85						
		PENHIGFC3B	3						97.5						
Case 3V	Case 3 with VGR	AENHIGFC3BV	1						97.5						
Case 4	Catalytic Gasifier, 80% Capacity Factor	ACATIGFC4B	1		Catalytic	≈24.6	80		85	98%	200				
		PCATIGFC4B	3												
Case 5	85% Capacity Factor	ACATIGFC5B	1				85								
		PCATIGFC5B	3												
Case 6	BOP Enhancements 20% ASU Power Reduction, Inverter Efficiency 98%, Stack Cost Reduction	ACATIGFC6B	1												
		PCATIGFC6B	3												
Case 6V	Case 6 with VGR	ACATIGFC6BV	1				97.5								

2 PATHWAY STUDY BASIS

Systems models were developed under the Aspen Plus® (Aspen) platform to simulate the IGFC process configurations. The major equipment characterizations were used to generate capital and operating cost estimates for the IGFC plants. 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 as described in the BB report and the previous IGFC pathway report. [2, 3]

Capital and operating costs for most of the conventional equipment items were scaled based on the updated BB cost estimates. [2] Costs of the components were estimated also based on the previous IGFC pathway report, as appropriate. [3] All the costs are reported in 2018 dollars, and the LCOE, calculated based on the methods used in the BB report, is presented as the revenue requirement figure-of-merit for each of the cases.

The design basis for the pathway study, which is largely based on the BB study, [2] is reported in this section, along with the environmental targets and cost assumptions.

2.1 SITE CHARACTERISTICS

All plants in this study are assumed to be located at a generic plant site in the midwestern U.S., with ambient conditions and site characteristics as presented in Exhibit 2-1 and Exhibit 2-2. The ambient conditions are the same as International Standards Organization (ISO) conditions.

Exhibit 2-1. Site Ambient Conditions

Elevation, m (ft)	0
Barometric Pressure, MPa (psia)	0.10 (14.696)
Design Ambient Temperature, Dry Bulb, °C (°F)	15 (59)
Design Ambient Temperature, Wet Bulb, °C (°F)	11 (51.5)
Design Ambient Relative Humidity, %	60

Exhibit 2-2. Site Characteristics

Location	Greenfield, Midwestern U.S.
Topography	Level
Size, acres	150
Transportation	Rail
Ash/Slag Disposal	Off Site
Water	Municipal (50%) / Groundwater (50%)
Access	Land locked, having access by train and highway
CO ₂ Storage	Compressed to 15.3 MPa (2,215 psia), transported 80 kilometers (50 miles) and sequestered in a saline formation at a depth of 1,239 meters (4,055 feet)

The IGFC cases assume that 15 acres of land area are required for the main plant along with an extra area amount that provides a buffer of approximately 0.25 miles to the fence line, which could also provide for a rail loop, if required. The steam turbine, unlike the gasifier and the SOFC stack units, is assumed to be enclosed in a building.

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

2.2 COAL CHARACTERISTICS

The design coal is Illinois No. 6 with characteristics presented in Exhibit 2-3 as per National Energy Technology Laboratory (NETL) Quality Guidelines for Energy System Studies (QGESS). [4] The first-year cost of coal used in this study is \$2.11/MMkJ (\$2.22/MMBtu) in 2018 dollars, in accordance with the values in the updated BB report. [2] The coal mercury content for this study was assumed to be 0.15 ppm (dry), which is consistent with the mercury content estimated and applied in the BB report. It was also assumed that all the coal mercury enters the gas phase, and none leaves with the bottom ash or slag.

Exhibit 2-3. Design Coal

Rank	Bituminous	
Seam	Illinois No. 6 (Herrin)	
Source	Old Ben Mine	
Proximate Analysis (weight %) ^		
	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	27,113	30,506
HHV, Btu/lb	11,666	13,126
LHV, kJ/kg	26,151	29,544
LHV, Btu/lb	11,252	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

2.3 NATURAL GAS CHARACTERISTICS

Natural gas is utilized as secondary fuel in one case (Case 1-6), and its composition is presented in Exhibit 2-4. [4]

Exhibit 2-4. Natural Gas Composition

Component		Volume Percentage
Methane	CH ₄	93.1
Ethane	C ₂ H ₆	3.2
Propane	C ₃ H ₈	0.7
n-Butane	C ₄ H ₁₀	0.4
Carbon Dioxide	CO ₂	1.0
Nitrogen	N ₂	1.6
Total		100.0
Unit	LHV	HHV
kJ/kg	47,454	52,581
MJ/scm	34.71	38.46
Btu/lb	20,410	22,600
Btu/scf	932	1,032

Note: Fuel composition is normalized, and heating values are calculated

The first-year cost of natural gas used in this study is \$4.13/MMkJ (\$4.42/MMBtu) as per the BB report. [5]

2.4 ENVIRONMENTAL TARGETS

The IGFC plant emissions are projected to be highly limited because:

1. The total sulfur content in the clean syngas must be maintained at less than 100 ppbv to protect the critical fuel cell materials;
2. Ninety-five percent of the mercury and other trace components are removed from the syngas;
3. The oxy-combustor is a low NO_x combustor; and
4. Most of the coal syngas contaminant species remaining after syngas cleaning are sequestered with the CO₂ product.

Only the plant solid waste streams, coal ash, spent sorbents, and waste catalysts might be of environmental concern. Accordingly, the emissions from the IGFC plants are expected to be much lower than the typical environmental targets identified in the BB report [2] for the conventional fossil fuel-based power plants.

Environmental targets for IGFC plants were based on the emissions limits for IGCC plants established in the BB report, which are shown in Exhibit 2-5, along with a summary of the control technology utilized in the IGFCs to satisfy the limits. [2]

Exhibit 2-5. Environmental Targets for IGFC cases

Pollutant	(lb/MWh-gross)	Control Technology
SO ₂	0.40	Selexol, and ZnO polishing
NO _x	0.70	Low NO _x oxy-combustors
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, sodium hydroxide treatment, SWS

The SO₂ limit is met via an AGR process, which captures sulfur as H₂S before it is oxidized to SO₂ later in the process train. An acid gas removal (AGR) process with sulfur capture efficiency of ≈99.9 percent is required to meet the environmental target for sulfur dioxide (SO₂) emissions, which corresponds to ≈10 ppmv sulfur in the sweet syngas. However, an additional sulfur polishing step is required for the IGFC systems to reduce the sulfur concentration in the syngas to less than 100 ppbv and prevent poisoning of the SOFCs.

Most of the coal ash is removed from the gasifier as slag or bottom ash. The ash that remains entrained in the syngas is captured in the downstream equipment, including the syngas scrubber and a cyclone, along with either ceramic or metallic candle filters, to meet the environmental targets for filterable particulates.

The Eastman Chemical Company plant, where syngas from a GEP gasifier is treated, achieved a Hg 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). Hg removal between 90 and 95 percent has been reported with a bed life of 18–24 months. Removal efficiencies may be even higher, but the measurement precision limit was reached at 95 percent. Eastman has yet to experience any Hg contamination in its product. [6] As an Hg removal efficiency of up to 97 percent is required to meet the Hg emissions limit, a dual sulfur-impregnated carbon bed system (i.e., two beds in series) capable of achieving greater than 99 percent Hg removal is required. It was assumed that the Hg removal efficiency is linearly related to bed depth; therefore, the bed depths were chosen to meet the emission limit.

2.5 BALANCE OF PLANT

The balance of plant assumptions are common to all cases and are presented in Exhibit 2-6.

Exhibit 2-6. Balance of Plant Assumptions

<u>Cooling system</u>	Recirculating Wet Cooling Tower
<u>Fuel and Other Storage</u>	
Coal	30 days
Slag/ash	30 days
Sulfur	30 days
Sorbent/catalyst	30 days
<u>Plant Distribution Voltage</u>	
Motors below 1 hp	110/220 volts
Motors between 1 hp and 250 hp	480 volts
Motors between 250 hp and 5,000 hp	4,160 volts
Motors above 5,000 hp	13,800 volts
Steam and Gas Turbine Generators	24,000 volts
Grid Interconnection Voltage	345 kV
<u>Water and Wastewater</u>	
Makeup Water	The water supply is 50 percent from a local Publicly Owned Treatment Works 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.

2.6 PLANT CAPACITY

The IGFC plant's net generating capacity is fixed at 650 MWe in this pathway study. The coal feed rate varies over a broad range from 126,312–223,530 kg/h (278,470–492,800 lb/h), as received over all the IGFC cases evaluated.

The pressurized pathway cases without CCS use a small amount of natural gas, ranging from $\approx$ 0–1400 kg/h (0–3100 lb/h) to boost the bottoming Brayton cycle.

2.7 CAPACITY FACTOR

The capacity factor for the IGFC reference plant is assumed to be 80 percent based on the IGCC cases used in the BB report. [7] The plant processing sections are designed for 100 percent capacity, with no excess capacity provided for any component other than the SOFC stack. An increase in capacity factor to 85 percent was explored as a pathway case to represent commercial maturity of both SOFC and gasifier technology.

The capacity factor and plant availability are equal since the present study assumes that each new plant would be dispatched any time it is available and would be capable of generating maximum capacity when online. The North American Electric Reliability Council (NERC) defines an equivalent availability factor (EAF), which is essentially a measure of the plant capacity factor assuming there is always a demand for the output. The EAF accounts for planned and scheduled derated hours as well as seasonal derated hours. As such, the EAF matches this study's definition of capacity factor.

2.8 SPARING PHILOSOPHY AND NUMBER OF PARALLEL PROCESS TRAINS

There is no redundancy provided in the case evaluations, other than normal sparing of rotating equipment. In practice, degradation of SOFC performance is mitigated by providing additional capacity in the form of extra SOFC surface area, coupled with operational strategies to maintain a constant power output as discussed in Section 2.13.11.

Apart from the SOFC section, the typical IGFC plant with CCS analyzed here consists of the ASU unit, the gasifier area, the oxy-combustor system, and the steam-cycle power island, as illustrated in Exhibit 2-7, where the captured CO₂ is dehydrated, compressed, and purified in the CPU. The corresponding IGFC plant without CCS is shown in Exhibit 2-8. Syngas expansion and cathode gas expansion, an additional component that is included for pressurized SOFC operation, are also assumed to interact with the SOFC power island that is highly modular, as explained in Section 2.9.4, and is designed for the shutdown and bypass of individual modules to enable turndown. Accordingly, the IGFC plant part-load performance is limited primarily by the syngas-supply component (gasifier, ASU) and the heat recovery component (oxy-combustor, HRSG, steam cycle) turndown limitations, and the turndown capability should be comparable to an IGCC with CCS. The load-follow capability of the IGFC, on the other hand, is primarily limited by SOFC heat-up rate constraints. The number of parallel processing trains of components utilized in the IGFC plant depends on the flow capacities for each case and were

based on the IGCC plants in the BB report, and as in the previous pathway study, [3] generally consisted of the following:

- Two parallel air separation units (2×100 percent)
- Two train gasification section, including gasifier, synthesis gas cooler, quench and scrubber (2×100 percent)
- Two parallel train syngas clean-up section (2×100 percent)
- Two parallel trains Selexol acid gas removal (2×100 percent), and two Claus-based sulfur recovery units (1×100 percent)
- Two oxy-combustor/HRSG trains (2×100 percent)
- One steam turbine system (1×100 percent)
- One CO₂ drying, compression, and purification unit (CPU)

Exhibit 2-7. General IGFC Plant Configuration with CCS

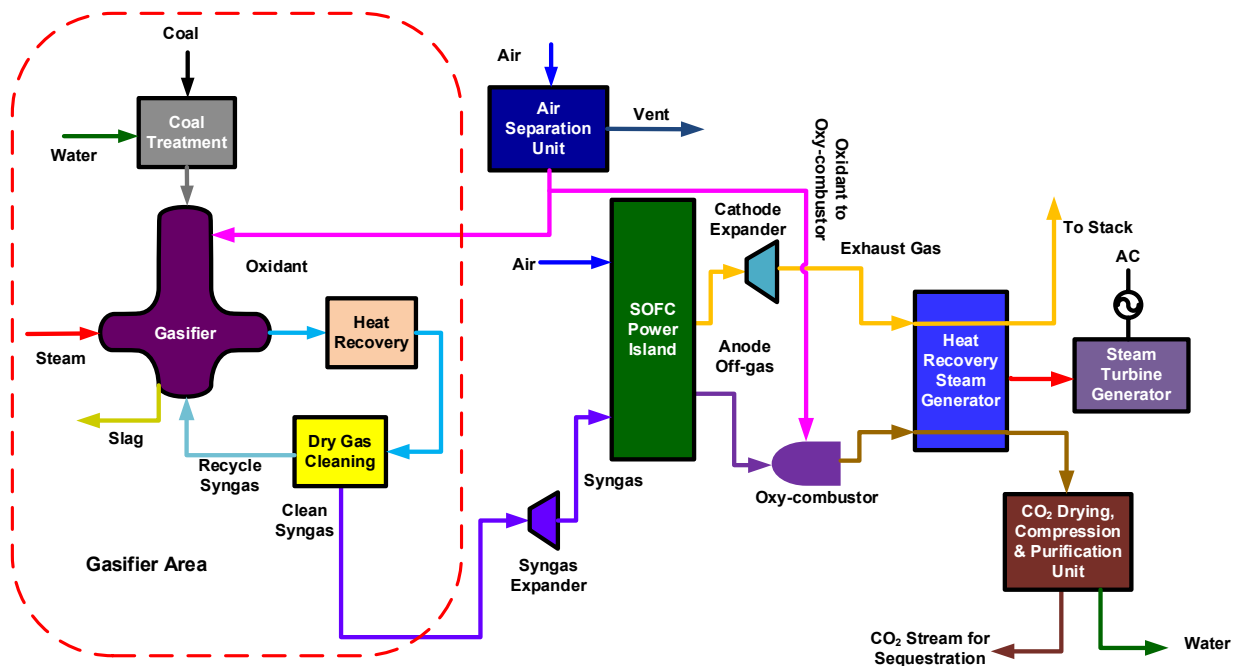
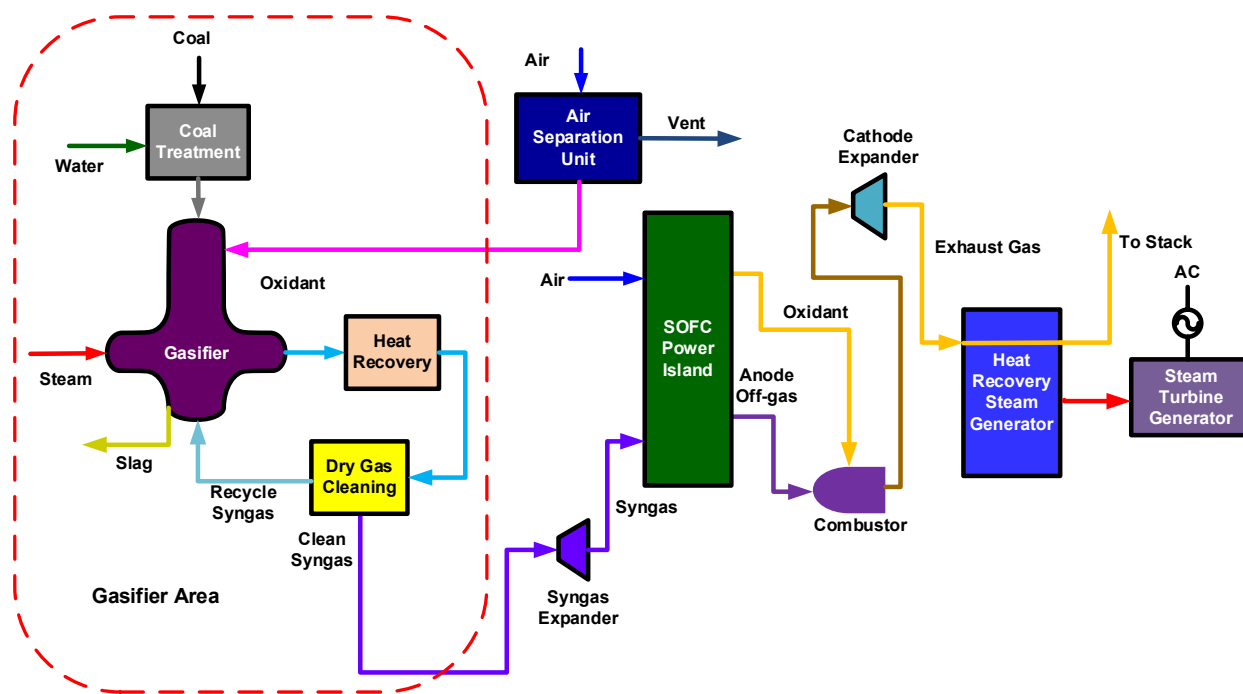


Exhibit 2-8. General IGFC Plant Configuration without CCS

2.9 SOFC POWER ISLAND CHARACTERIZATION

The SOFC simulations represent the expected operating conditions and performance capabilities of planar fuel cell technology, having split cathode and anode off-gas streams and operating at both atmospheric-pressure and elevated-pressure conditions. Several assumptions were applied to estimate the performance of the IGFC power island. These assumptions were generally based on SOFC test data and internal vendor reports.

2.9.1.1 Estimation of SOFC Operating Voltage

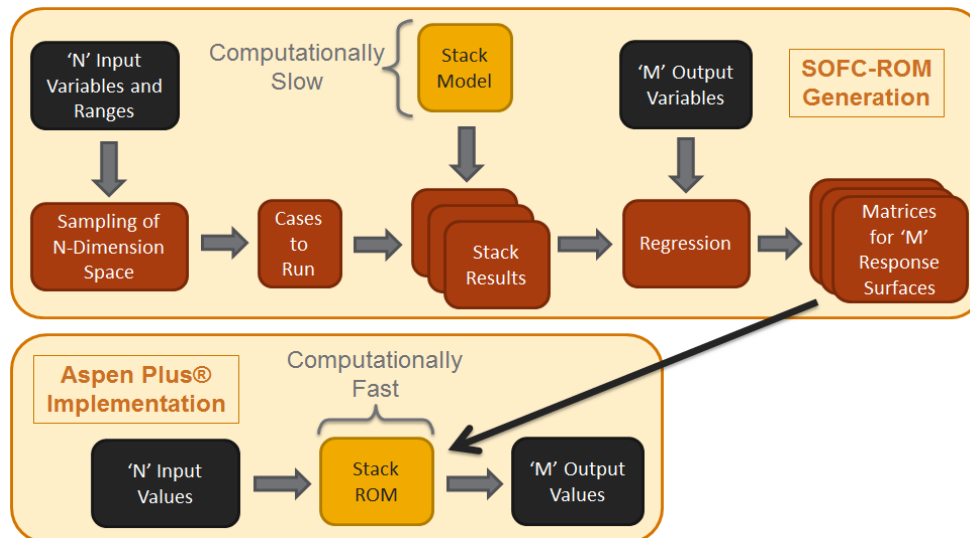
NETL's previous IGFC pathway study [3] was based on simplified representations of the SOFC performance at only a specific design point. While a variety of detailed computational models can be employed to predict the electrical performance, along with the thermal and flow fields of SOFCs, these models are generally computationally intensive for use in the system-level thermodynamic and material balance models typically used in techno-economic analyses. Through a partnership with the Pacific Northwest National Laboratory (PNNL), NETL has developed a reduced order model (ROM) based on detailed computational models of the SOFC stack. This model, which is referred to as the PNNL-ROM model here, is not computationally intensive, and allows for effective use in system-level models. The PNNL-ROM model, in addition to its ability to predict the stack performance over the entire range of operating conditions, is designed to provide other desirable information about the state of the stack, such as internal temperature gradients, which can be critical constraints in system optimization studies.

The PNNL-ROM model was used to predict the SOFC performance, along with the cell thermal fields, for a given airflow and fuel flow for the IGFC studies. The PNNL-ROM model is based on

the response surface methodology [8], with detailed SOFC stack model results to create a computationally efficient ROM for the stack that retains desirable information about its internal state. The response surface approach attempts to describe the behavior of a dependent or “response” variable of interest based on the independent or “explanatory” variables that are inputs to a given process or model. For the SOFC, the input variables could constitute operational parameters such as fuel/oxidant compositions, temperatures, flow rates, and utilizations. Response variables could include electrical performance characteristics such as stack voltage, power output, and efficiency. Peak cell/stack temperature, cell temperature gradient, or maximum local current density are among other response variables that may be of interest due to their influence on cell/stack structural stability and performance degradation.

The developed numerical process for ROM generation is shown schematically by the flow chart in Exhibit 2-9. The number of different model input variables and their value ranges are assigned based on expected SOFC operating conditions. This defines the multi-dimensional design space for the problem, and random sampling is performed to define sets of parameters. Each sample set defines a modeling case to be run. The detailed stack model then runs each of the defined cases and collects the results. Based on the expected usage of the ROM, the output parameters of interest are also defined. The corresponding values for these parameters are then extracted from the case results, and regression is performed to a selected fitting function. The regression results provide a mathematical relationship describing the predicted response surface for each response variable as a function of the input variables. These response surfaces are then exported in a suitable format for integrating the ROM into the system model. This ROM then provides a computationally efficient predictor of a given stack’s performance without integrating the entire detailed model directly into the system model, thereby adding further nonlinear convergence iterations to that solution. Further details of the PNNL-ROM model, the regression methodology, and its validation can be found in the Koepfel et al. study. [9]

Exhibit 2-9. Overview of the ROM Generation Procedure and Usage



Although the ROM procedure is applicable to comprehensive 3-D models, the stack is presently modeled using PNNL’s SOFC-multi-physics (MP) software [10] that solves for the steady-state

cell distributions of species, temperature, and current density. The underlying 1-D channel cell model represents a 550 cm² active area anode-supported cell with metal interconnects in a counter-flow configuration. The model assumes the water-gas shift reaction is in equilibrium but uses a first-order kinetic expression for the slower on-cell steam reforming of methane with a Ni/YSZ anode. [10]

The voltage-current (V-I) relationship for the cell within the stack model is defined by a user-defined function that returns a voltage based on the local temperature, species concentrations, and current density at each location of the cell's active area. The stack model iterates on the current density distribution until the cell voltage is converged. The cell voltage is calculated by subtracting expressions for the activation, ohmic, and concentration polarization losses from the open circuit voltage as given by the equation:

$$V_{cell} = V_{OC} - \eta_{act} - i * R - \eta_{conc}^{anod} - \eta_{conc}^{cath}$$

Where V_{cell} is the cell voltage, V_{OC} is the open circuit potential, η_{act} is the voltage loss due to activation polarization, R is the overall ohmic resistance, and η_{conc}^{anod} and η_{conc}^{cath} are voltage losses due to diffusion polarization at the anode and cathode, respectively. The coefficients used in the polarization equations were selected to provide a V-I response representative of high performance state-of-the-art (SOA) planar cells operating under atmospheric conditions. The selected cell performance provides 0.8 V at 400 mA/cm² operating at a temperature of 750 °C for a wet hydrogen (H₂) fuel (97 percent H₂, 3 percent water (H₂O)) with 75 percent fuel utilization and 12.5 percent air utilization (see [9] for the full V-I characteristics).

2.9.2 SOFC Carbon Deposition Control

The SOFC stack inlet anode gas composition can induce the formation of solid carbon deposits, which can disrupt the normal performance of the stack. [11] Anode gas recirculation is used to control the anode gas inlet conditions to maintain an atomic oxygen to carbon (OTC) ratio greater than 2.1, which is a generally used criterion to prevent carbon deposition anywhere in the SOFC fuel flow domain. [12] Anode gas recirculation is accomplished using hot gas blowers for the atmospheric cases, and jet-pumps for the pressurized cases, which also serve to keep the fuel utilization within the stack below acceptable limits from a flow distribution perspective.

2.9.3 Estimation of Steam Bottoming Cycle Performance

The anode off-gas stream is combusted with oxygen for the cases with carbon capture, while a portion of the SOFC cathode exhaust is utilized for combustion in cases without carbon capture. In both cases, the hot stream from the combustor exchanges heat in an HRSG system to produce steam, which generates power in a subcritical steam bottoming cycle after satisfying process steam requirements. The steam cycle provides a relatively small proportion of the total plant generation output, with steam conditions and capacity varying greatly across study cases. In some cases, the heat recovery temperature available is relatively low, resulting in poor steam superheat conditions. A simple subcritical steam cycle was included in the conventional coal gasifier case simulations. Conversely, the steam cycle performance for the catalytic gasifier cases was estimated using a nominal steam cycle efficiency of 38.1 percent, based on simulations of steam cycles in select pathway cases. [3] For conventional IGFC cases, this

approach has been shown to result in efficiencies of up to 1.5 percentage points lower or higher than fully simulated steam cycle model values depending on whether the SOFC is operating at atmospheric or elevated pressures, respectively. These uncertainties must be kept in mind, especially while interpreting the catalytic gasifier case results. The auxiliaries for the steam cycle were scaled from the BB report [2] based on the steam cycle power output.

2.9.4 IGFC Power Island

The IGFC power island (Exhibit 2-7 and Exhibit 2-8) consists of a syngas expander that expands the syngas from its reformer outlet condition down to the operating pressure of the fuel cell unit; the SOFC power island with direct current (DC)-alternating current (AC) inverters; an anode off-gas oxy-combustor; an HRSG that captures heat from the combusted anode off-gas; and a steam bottoming cycle.

The SOFC power island configuration is modular and is made up of a parallel train of “sections” that are essentially assemblies of basic building blocks—the SOFC modules. SOFC modules are independently manufacturable units built on the commercially viable 1 MWe class DGFC units that include planar SOFCs stacked and connected electrically in a convenient fashion. Each module is distinguished by an enclosure, which serves as a pressure and material boundary and houses the insulation system, the fuel and air distribution systems, the current collection systems, and the instrumentation and control (I&C) system, in addition to the SOFC stacks.

In the present study, the size of the modules was determined based on a nominal cell operating at a voltage of 0.8 V at a current density of 400 mA/cm², corresponding to a power density of 400 mW/cm². The module is comprised of several SOFC stacks, each consisting of 100 cells, each measuring 25 × 25 cm in cross-section, with an effective area of 550 cm². A 64-stack module was found to be suitable based on a variety of factors, including the stack arrangement, module total current, module voltage rating, and shipping constraints. The 64-stack module, whose dimensions are listed in Exhibit 2-10 along with the salient assumptions, also aligns well with the development of the 1 MWe class systems currently being pursued under the fuel cell program.

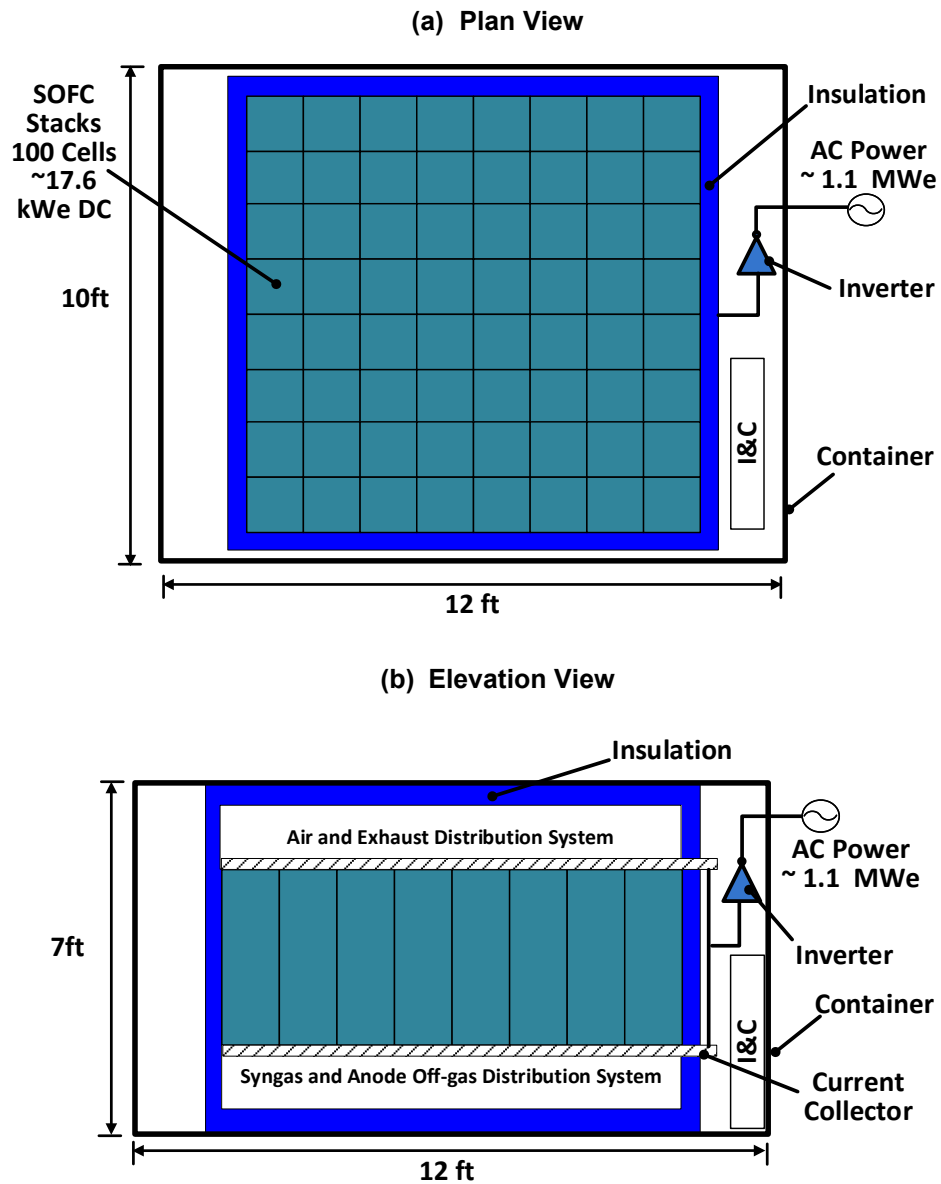
The enclosing pressure and material boundary for atmospheric and pressurized SOFC modules are referred to as a “container” and “pressure” vessel, respectively.

Exhibit 2-10. IFGC Module Sizing

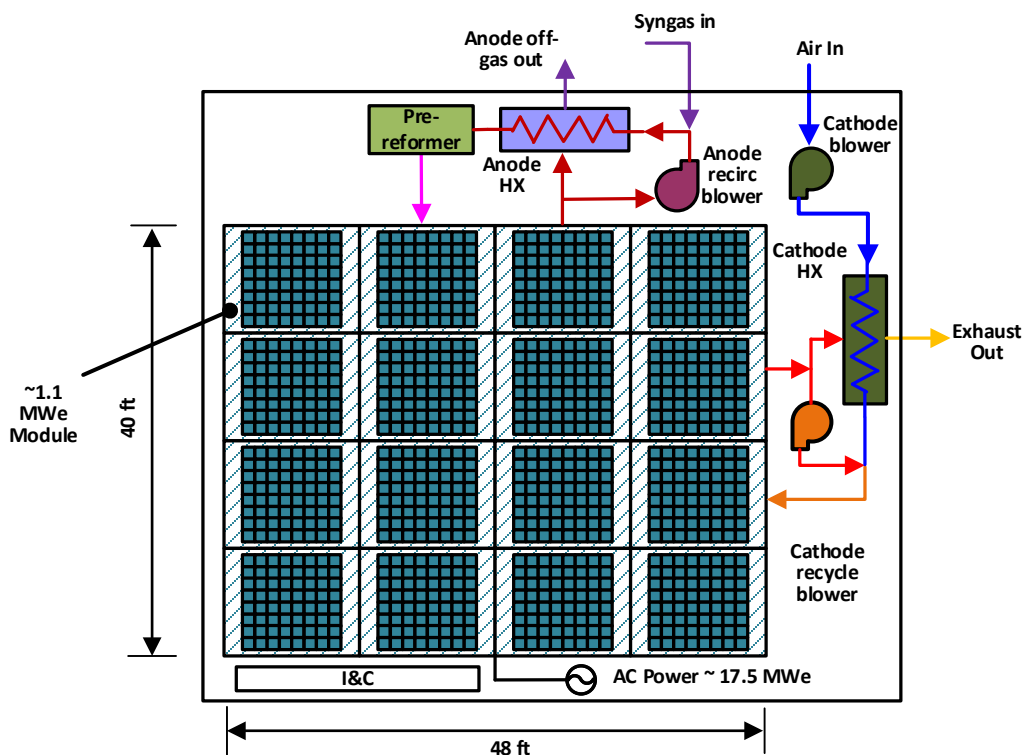
Nominal Cell Voltage (V)	0.8
Nominal Current Density (mA/cm ²)	400
Nominal Power Density (mW/cm ²)	320
Active Cell Area (cm ²)	550
Nominal Cell Current (A)	220
Cell Width in (cm)	9.84 (25)
Cell Length in (cm)	9.84 (25)
Cell Thickness in (mm)	0.315 (8)
Number of Cells per Stack	100
Nominal Stack Voltage (V)	80
Nominal Stack Current (A)	220
Nominal Stack Power (kWe DC)	17.6
Stack Height (in)	31.5
Stack Pitch in Module Spanwise (x) Direction (in)	12.0
Stack Pitch in Module Lengthwise (z) Direction (in)	12.0
Insulation Thickness (in)	5
Additional spacing allowance for sub-systems	24
Spacing above and below stack for flow distribution	36
Number of Stacks per Module	64
Number of Stacks in Module Spanwise (x) Direction	8
Number of Stacks in Module Lengthwise (z) direction	8
Number of Cells per Module	6400
Total Current per Module (A)	440
Total Voltage per Module (V)	2560
Atmospheric SOFC Module rating (MWe DC)	1.126
Container/Vessel Head	Flat
Atmospheric SOFC Container Dimensions (ft)	Width x Length x Height = 10 x 12 x 7
Pressurized SOFC Vessel Dimensions (ft)	Diameter x Length = 10 x 12
Pressurized SOFC Vessel Diameter ft (in)	9.9 (118.8)

Plan and elevation views of the finalized SOFC module layout are shown schematically in Exhibit 2-11 (a) and (b), which also depict the major components considered to be within the module boundary for the present study.

Exhibit 2-11. IGFC Module Layout



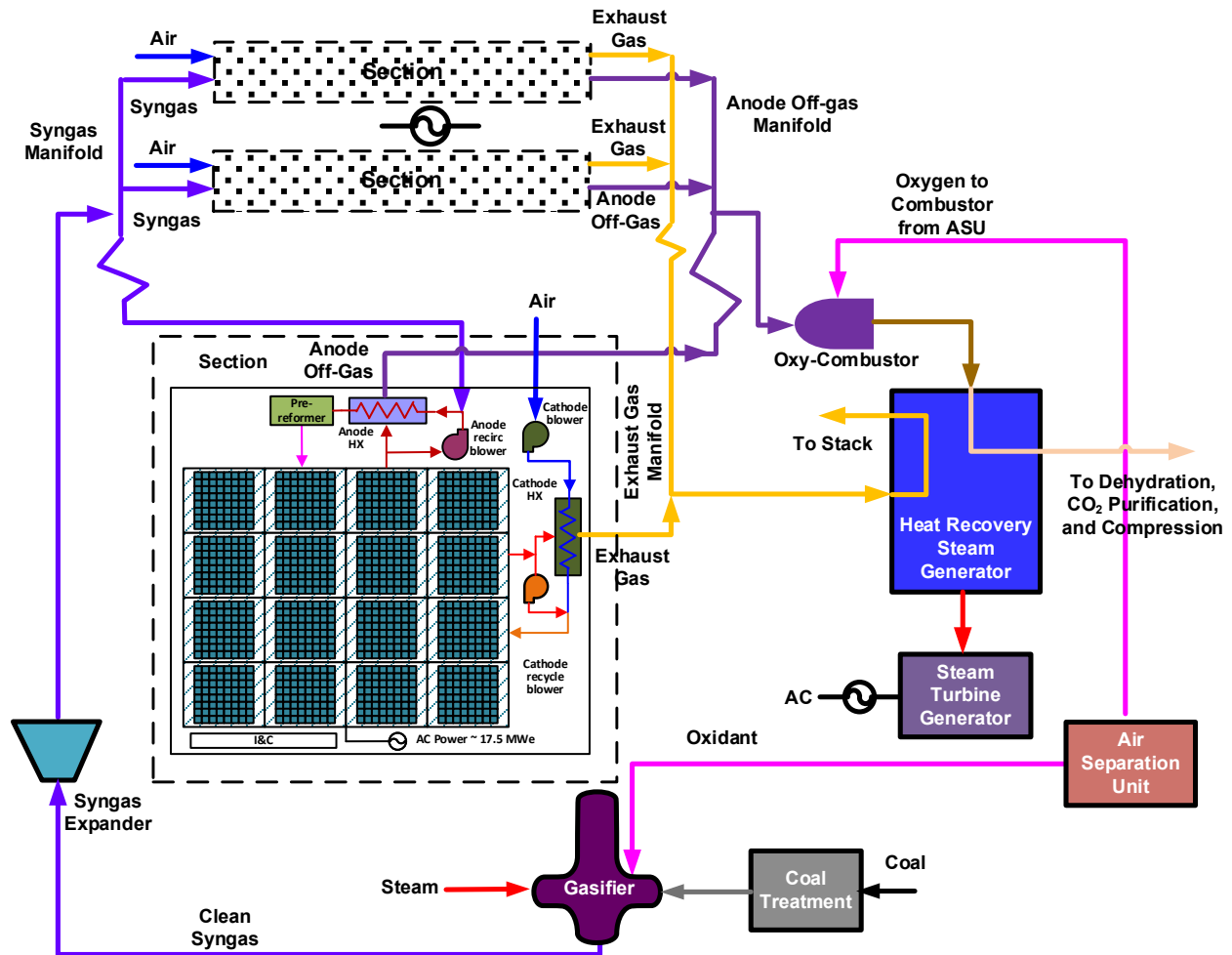
Several SOFC modules are assembled into sections, which also houses SOFC BOP equipment such as heat exchangers, blowers, sectional I&C, and the power collection system for ease of thermal, flow, and power management. A section that consisted of 16 modules, as depicted by the layout in Exhibit 2-12, was deemed to be nominal in the present study from flow distribution and current collection perspectives. The power rating of the resulting section is ≈ 19 MWe, and its salient electrical characteristics are listed in Exhibit 2-13. Trade-offs between the number of modules per section and the total system cost will be explored as a separate study.

Exhibit 2-12. SOFC Section Layout: Components and Dimension**Exhibit 2-13. Characteristics of SOFC section**

Number of modules in parallel	4	
Number of modules in series	4	
Total current per section	1760	A
Total voltage per section	10240	V
Total power per section	17.6	MWe
Total current per section	1760	A

The overall system is a collection of sections, as shown in Exhibit 2-14, connected electrically in parallel and pneumatically to produce the total rated power. A network of insulated pipes and ducts distribute the syngas from the central syngas expander to each section, while the anode and cathode exhaust gases are collected and sent to a central oxy-combustor or combustor for plants with and without CCS, respectively. The pressurized SOFC system is nearly identical to the atmospheric system layout and features, in addition, a central cathode gas expander and a central oxy-combustor expander (for systems without CCS).

Exhibit 2-14. Atmospheric IGFC System Layout



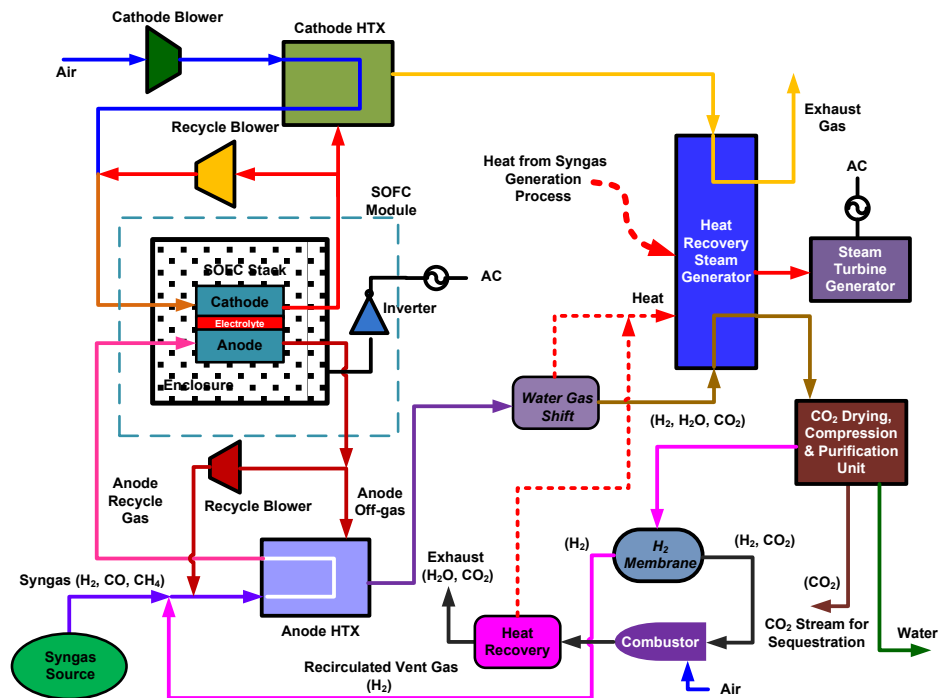
2.9.5 VGR Concept

The process diagram of the VGR in its most general form is shown in Exhibit 2-15. Unlike the typical SOFC-based systems, there is no oxy-combustor; the anode exhaust from the fuel cell flows through a water gas shift (WGS) reactor, which converts most of the CO to CO₂, before being fed to the CPU unit subsequent to heat recovery and dehydration. The vent gas stream remaining after the CO₂ separation in the CPU is rich in fuel containing mainly H₂ and CO₂. The H₂ in this mixture is separated using an H₂ membrane and fed back to mix with the fuel from the syngas source before entering the SOFC module. Depending on the CO₂ capture rate and the H₂ capture rate, exhausting a small portion of the CPU vent stream may be necessary due to mass balance considerations. However, the fuel energy in the exhausted residual fuel can be recovered in a normal air blown combustor. A conventional pressure swing adsorption process could also be used in place of the H₂ membrane.

The proposed system has several advantages:

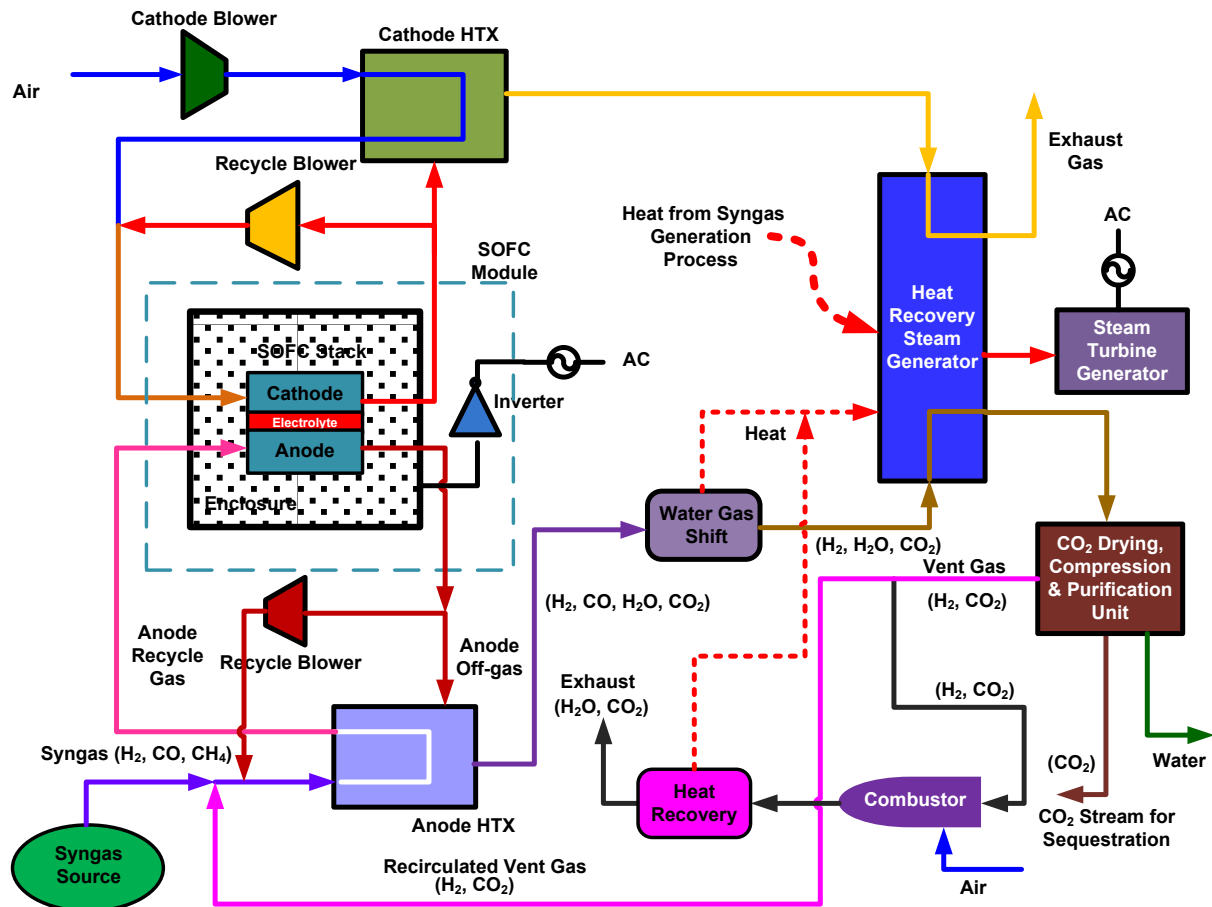
1. The single-pass fuel consumption across the fuel cell can be sufficiently low, even though an overall system fuel utilization of nearly 100 percent can be achieved by controlling the rate of VGR. This has a direct benefit on the system efficiency.
2. VGR also increases the chemical potential of the fuel entering the SOFC, thereby increasing the electrochemical (Nernst) potential and the system performance.
3. The oxy-combustor (along with its O_2 supply) is eliminated, lowering the ASU O_2 demand, and consequently, lowering the ASU parasitic load. In the case of an IGFC system with complete internal reformation, the ASU system is eliminated.
4. A higher VGR rate also makes the fuel cell system robust to fuel flow maldistributions and eases the design of fuel distribution systems.
5. At the same current load, the system allows operation at a cell voltage higher than conventional systems, alleviating performance degradation concerns.
6. The airflow rate to the SOFC system, which constitutes a major parasitic load, is also reduced due to the VGR serving as an additional coolant for removing the SOFC waste heat.
7. No additional technological advancements are needed to enable the system. It can be used with current technology.

Exhibit 2-15. Generalized Configuration of the VGR Concept



A more attractive version of the system without the need for a H₂ membrane is shown in Exhibit 2-16, where the CPU vent stream is directly fed back into the SOFC module fuel inlet without any separation of the H₂. The elimination of the feed stream compression required for the H₂ membrane, coupled with fuel losses due its <100 percent effectiveness, are generally expected to offset adverse effects of the remnant CO₂ in the vent gas stream on the performance through dilution of the fuel entering the SOFC module. Further, elimination of the H₂ membrane, which is still in a developmental stage, results in improved cost savings and reliability.

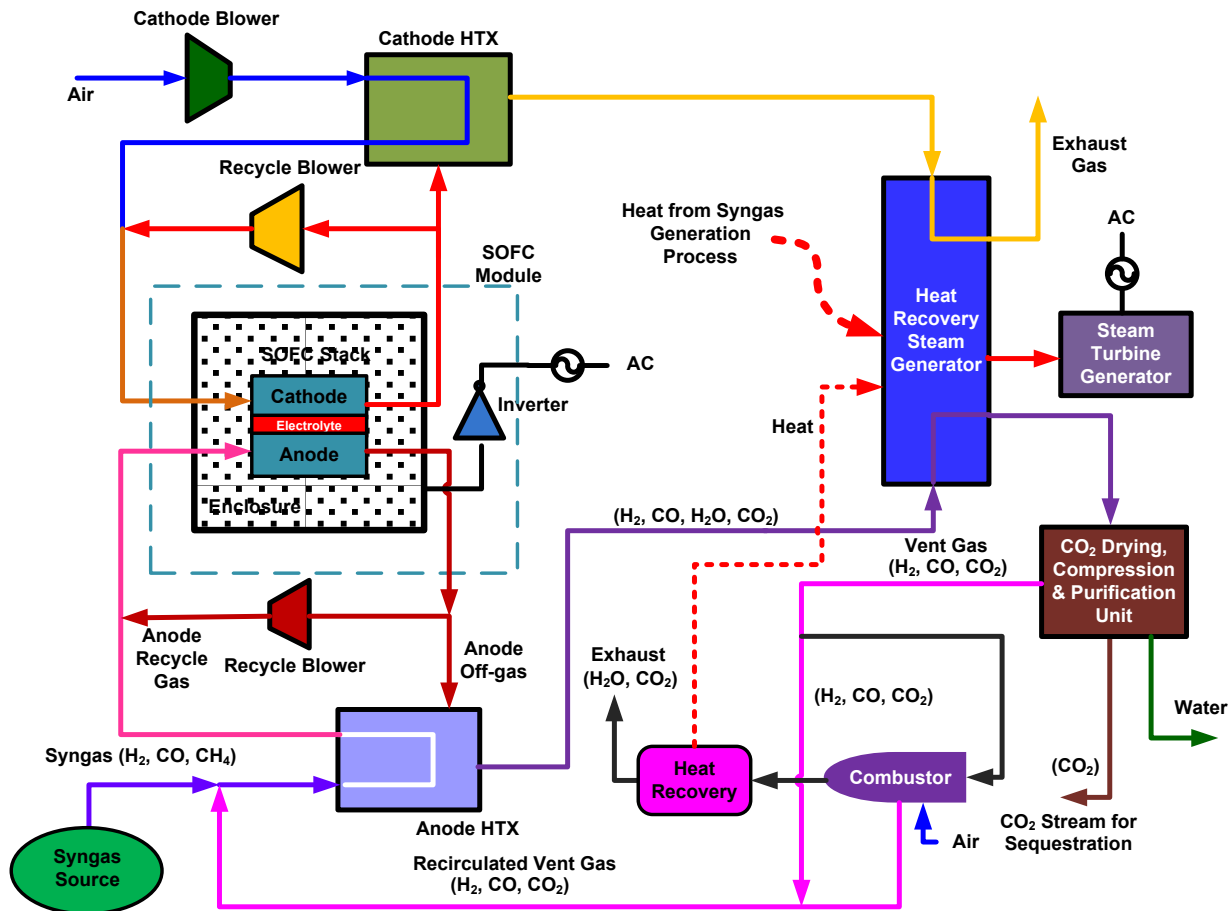
Exhibit 2-16. Configuration of the VGR Concept without an H₂ Membrane



In this case, the recirculated vent gas contains CO and CO₂ and will need steam to maintain an oxygen-to-carbon ratio (OCR) of 2.1 to prevent carbon formation at the high temperatures within the SOFC module. The internal anode off-gas recirculation loop is accordingly an essential part of this system to avoid the need for external steam injection, which will have an adverse impact on the system efficiency and cost, to avoid carbon deposition due to any CO/CO₂ that mixes with the VGR. The combustor is located downstream of the CPU to maximize the rate of CO₂ capture and minimize any steam losses. Further, any valve needed to regulate the vent gas recirculation is in a cold region and can be easily controlled.

Exhibit 2-17 shows an alternate configuration of the proposed system, which eliminates the WGS reactor in addition to the H₂ membrane. This configuration is economically attractive and formed the basis for the VGR cases in the present study. Similar to the configuration with the H₂ membrane, exhausting a small portion of the CPU vent stream may be necessary due to mass balance considerations in the variants shown in Exhibit 2-17 and Exhibit 2-16, which also serves to limit the accumulation of inert species such as Nitrogen (N₂) and Argon (Ar) in the recirculating vent gas.

Exhibit 2-17. Configuration of the Proposed System without both the WGS Reactor and the H₂ Membrane



2.10 AIR SEPARATION UNIT

The conventional cryogenic ASU generates oxidant for use in two sections of the IGFC plant: the gasifier and the anode off-gas oxy-combustor. In this report, the ASU main air compressor discharge pressure was set at 0.5 MPa (79 psia), providing O₂ product at sufficient pressure—0.16 MPa (23 psia)—to operate the oxy-combustor for the atmospheric pressure SOFC applications. The ASU is designed to generate 99.5 percent pure O₂ for IGFC applications to maintain the sequestered CO₂ stream with low nitrogen and argon content. There is no opportunity for ASU air-side integration in the IGFC plant like there is in IGCC plants, and there is no need or benefit from nitrogen dilution of syngas in the IGFC. In this report, the ASU

nitrogen product is used only for inert gas needs, with the remainder vented. The air compressor providing air to the process is powered by an electric motor.

2.11 CO₂ CAPTURE TECHNOLOGY

The anode off-gas is combusted using 99.5 percent O₂ in an advanced oxy-combustor, with excess O₂ limited to 1 mole percent. The combusted anode gas, consisting of CO₂, water vapor, excess O₂, and minor traces of contaminants (sulfur species and oxides of nitrogen [NO_x]), is sent to a CPU for compression, drying, and purification to pipeline/ EOR levels.⁵ [13] In the CPU, the CO₂ is purified by liquefaction, followed by flash separation and distillation. The cooling for the liquefaction is accomplished by an auto-refrigeration process. The CPU was designed to yield an O₂ content of 10 ppm in the CO₂ product stream.⁶ The CPU cost was scaled based on an available quote.

2.12 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. Internal recycle water available from various sources, including boiler feedwater blowdown and condensate from syngas cleaning or from CO₂ gas compression, was used to offset the water demand. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is the water removed from the ground or diverted from a surface-water source for use in the plant. Raw water consumption is also accounted for as the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products or otherwise not returned to its source.

Raw water makeup was assumed to be 50 percent from 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, boiler feedwater makeup, slurry preparation makeup, ash handling makeup, and quench system makeup. Consumption represents the withdrawn water permanently lost from the water source.

Boiler feedwater 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 sour water stripper (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 and evaporative cooling tower, and all process blowdown streams were assumed to be treated and recycled to the cooling tower.

⁵ The QGESS reference [39] recommends a CO₂ purity of at least 95% as a conceptual design basis for EOR purposes; however, distillation methods used to meet the more stringent O₂ concentration requirement of 100 ppmv or less generally result in 99.9% + CO₂ purity.

⁶ An O₂ concentration of 10 ppmv in the CO₂ product stream was selected as the basis for conceptual design since it represents the lower limit of the range of values recommended in literature for EOR applications. [39] The number of distillation stages can be reduced slightly to design to the upper limit of 100 ppmv for O₂ concentration. [39] However, the impact of the associated small decrement in distillation cost on the overall cost is expected to be insignificant.

A cooling water temperature of 16°C (60°F) was assumed based on Exhibit 2-1, along with 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 [14] assumptions:

- 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 were calculated as follows:
- Blowdown Losses = Evaporative Losses / (Cycles of Concentration - 1)
 - Where cycles of concentration are a measure of water quality, and a mid-range value of 4 was chosen for this study

The water balances presented in subsequent sections include the water demand of the major water consumers within the process, the amount of process water returned to the source, and the raw water consumption.

2.13 ECONOMIC ANALYSIS

Capital and production cost estimates follow the economic basis applied in the BB report, which provides factored estimates developed for each plant section for conventional fossil fuel plants. [2] This report scales those costs for comparable plant sections that appear in the IGFC plants. Costs were factored using operating variables and scaling exponents appropriate for each system account. Costs for unique equipment in the IGFC plants were estimated using available generalized cost correlations or using cost estimates for comparable equipment reported in other power plant studies. In the case of the SOFC stack components, the estimated capital costs were based on current NETL technology development cost goals and SOFC vendor projections.

2.13.1 Cost Estimating Methodology

Following the basis used in the BB report, [2] the capital costs at the total overnight cost (TOC) level include equipment, materials, labor, indirect construction costs, engineering, owner's costs, and contingencies. Where applicable, the cost of major conventional plant sections in the study case plants are based on scaled estimates from costs presented in the BB report, applying the general cost-scaling equation:

$$C = N * (C_{ref} / N_{ref}) * [(F / N) / (F_{ref} / N_{ref})]^S$$

where;

C is the cost of the study case plant section,

N is the number of parallel sections in the study case plant,

C_{ref} is the cost of the reference plant section,

N_{ref} is the number of parallel sections in the reference plant,

F is the capacity of the study case plant section,

F_{ref} is the capacity for the reference plant section, and

S is the scaling factor characteristic of the plant section equipment (a fraction typically between 0.5 and 0.8).

The TPC and operation and maintenance (O&M) costs for each of the cases in the study were estimated using data generated by Black and Veatch (B&V) for the BB report. [2] The estimates carry an accuracy of ± 30 percent, consistent with the screening study level of information available for the various study power technologies.

All capital costs are presented as “overnight costs” expressed in June 2018 dollars.

Capital costs at the TPC level includes:

- Equipment (complete with initial chemical and catalyst loadings),
- Materials,
- Labor (direct and indirect),
- Engineering and construction management, and
- Contingencies (process and project).

The total as-spent cost (TASC) is then calculated by adding the estimated financing costs to the TOC, which is a sum of the owner’s costs and the TPC. The current-dollar, levelized cost of electricity is calculated using TOC.

2.13.2 Plant Maturity

The case estimates provided include technologies at different commercial maturity levels, and the overall IGFC plants represent highly advanced, immature technologies. Although the commercial components in the IGFC plants are based on data from commercial IGCC offerings, there have been very limited sales of these units so far.

The SOFC and oxy-combustion technologies for the IGFC cases are also immature and are unproven at commercial scale in power generation applications. The developing SOFC technology performance and cost has been estimated through scaling to commercial levels by the SOFC developers. While commercial pre-combustion CO₂ removal technology could be applied in place of the oxy-combustion based CO₂ removal, the oxy-combustion technology merits additional development due to its large advantage over pre-combustion CO₂ removal. The catalytic gasification technology is based on prior extensive development work conducted for a similar coal gasification technology by Exxon in the 1970s for the purpose of synthetic natural gas (SNG) production. The specific catalytic gasifier simulated for application to IGFC has not been tested and represents a conceptual processing step in the pathway evaluation.

2.13.3 Estimate Scope

The estimates represent a complete power plant facility on a generic site. Site-specific considerations such as unusual soil conditions, special seismic zone requirements, unique local conditions, or local regulatory requirements are not considered in the estimates.

The estimate 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. The single exception to the fence line limit is in the CO₂ capture cases where costs are included for T&S of the sequestered CO₂.

2.13.4 Capital Costs

B&V developed the capital cost estimates for IGCC plants in the BB report [2] using the company’s in-house database and conceptual estimating models for each of the specific technologies. The estimating models are based on a reference bottoms-up estimate for each major component. This provides a basis for subsequent comparisons and easy modification when comparing between specific case-by-case variations.

Some equipment costs for the cases were calibrated to reflect recent quotations and/or purchase orders for other ongoing in-house power or process projects. These include, but are not limited to, the following equipment:

- Steam Turbine Generators
- Circulating Water Pumps and Drivers
- Cooling Towers
- Condensers
- Air Separation Units (partial)
- Main Transformers

Other key estimate considerations include the following:

- Labor costs are based on Midwest Merit Shop. Costs would need to be reevaluated for projects at different locations or for projects employing union labor.
- 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 (five 10-hour days). No additional incentives such as per diem 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. Current indications are that regional craft shortages are likely over the next several years. The types and amounts of incentives will vary based on project location and timing relative to other work. The cost impact resulting from an inadequate local work force can be significant.
- 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.

- Costs are limited to within the “fence line,” terminating at the high voltage side of the main power transformers, except for costs included for T&S of the sequestered CO₂ in all capture cases.
- Engineering and Construction Management were estimated as a percentage of bare erected cost based on the BB report. 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.
- All capital costs are presented as “Overnight Costs” in June 2018 dollars. Escalation to period-of-performance is specifically excluded.

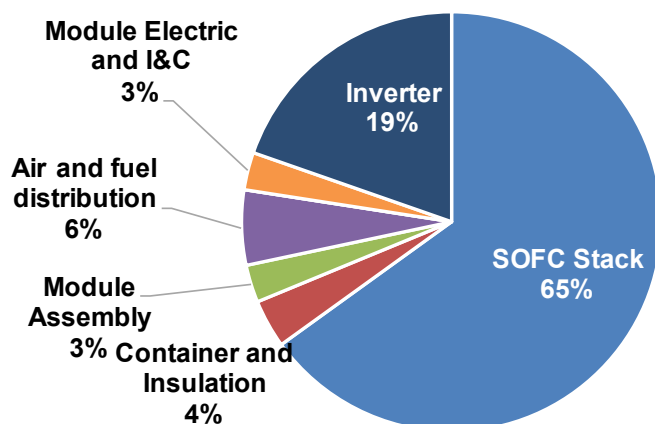
The current-dollar LCOE was calculated for each case using economic parameters in the BB report. [2] The capital component of LCOE was calculated using the TASC at a capital charge factor of 0.077 along with a TASC/TOC ratio of 1.093.

2.13.5 SOFC Module Costs

The NETL goal of \$225/kWe in 2018\$ for the stack cost forms the basis for the SOFC cost calculations. The atmospheric SOFC module cost is itemized in Exhibit 2-18 according to the major components within the module. Wherever available, costs were scaled based on the references based on the references indicated. The cost of the container for atmospheric systems was estimated from commercial shipping container costs while costs for air distribution, fuel distribution, and current collectors were estimated based on authors’ previous industrial experience. The SOFC stacks and inverter account for nearly 85 percent of the module costs, as shown by the cost distribution pie chart.

Exhibit 2-18. Atmospheric SOFC Module Costs

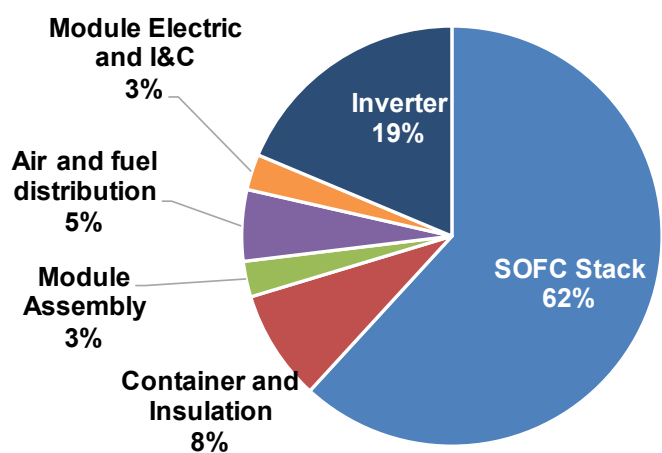
Module costs (2018\$/kWe AC) Atmospheric		
SOFC Stack	225	
Container	9	[15]
Insulation	4	[16]
Module Assembly	10	[16]
Air distribution	10	Estimate
Fuel distribution	10	
Module Current Collectors	5	
Module I&C	5	
Inverter	68	[17]
Total Module	356	



The corresponding costs for the pressurized SOFC module are shown in Exhibit 2-19 and differ only by the enclosure cost, which was estimated using Aspen Activated Analyses and reflects the significantly higher cost for the pressure vessels required for pressurized SOFC operation.

Exhibit 2-19. Pressurized SOFC Module Costs

Module costs (2018\$/kWe AC) Pressurized		
SOFC Stack	225	
Pressure Vessel	27	Estimated from Aspen Activated Analysis
Insulation	4	[16]
Module Assembly	10	[16]
Air distribution	10	Estimate
Fuel distribution	10	
Module Current Collectors	5	
Module I&C	5	
Inverter	68	[17]
Total Module	374	



2.13.6 Exclusions

The capital cost estimate includes all anticipated costs for equipment and materials, installation labor, professional services (Engineering and Construction Management), contingency, and owner's costs. The following items are excluded from the capital costs:

- Site-specific considerations – including, but not limited to, seismic zone, accessibility, local regulatory requirements, excessive rock, piles, laydown space, etc.
- Labor incentives in excess of a 5-day/10-hour work week
- Additional premiums associated with an engineer/procure/construct (EPC) contracting approach

2.13.7 Contingency

Both the project contingency and process contingency costs represent costs that are expected to be spent in the development and execution of the project that are not yet fully reflected in the design. It is industry practice to include project contingency in the TPC to cover project uncertainty, and the cost of any additional equipment that would result during detailed design. Likewise, the estimates include process contingency to cover the cost of any additional equipment that would be required because of continued technology development, and the project and process contingencies applied are consistent with those used in the BB report.

Based on the Association for the Advancement of Cost Engineering (AACE) international contingency guidelines, as presented in NETL's QGESS, it would appear that the overall project contingencies for the subject cases should be in the range of 30 to 40 percent. [18] However, such contingencies are believed to be too high when the basis for the cost numbers is considered. The costs have been extrapolated from an extensive data base of project costs (estimated, quoted, and actual), based on both conceptual and detailed designs for the various technologies. This information has been used to calibrate the costs in the current studies, thus improving the quality of the overall estimates. As such, the overall project contingencies should be more in the range of 15 to 20 percent based on the specific technology.

No project contingency has been applied to the SOFC stack unit cost; these contingencies already being incorporated by vendor estimates for the SOFC stack unit. A 15 percent project contingency has been applied to the ancillary components in the SOFC power island.

Process contingency is intended to compensate for uncertainties arising because of the state of technology development. No process contingency was placed on the SOFC stack unit cost, with the IGFC plant cost sensitivity to variations in the SOFC stack unit cost to be separately examined. Process contingencies have been applied to the estimates as follows:

- Slurry Prep and Feed – 5 percent on CoP IGFC cases – systems are operating at a high as 800 psia as compared to 600 psia in IGCC experience
- Gasifiers and Syngas Coolers – 15 percent on all IGFC cases – next-generation commercial offering and integration with the power island

- Trace Element Removal – 5 percent – minimal commercial scale experience in IGCC applications
- SOFC power island ancillary components – 15 percent

2.13.8 Operation and Maintenance

The production costs or operating costs and related maintenance expenses 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 byproduct credit (that is, a negative cost for any byproducts 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. The approach followed in estimating these costs is consistent with that applied in the BB report.

2.13.9 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 \$38.50 per hour. The associated labor burden is estimated at 30 percent of the base labor rate. Six operators per shift are assumed in all cases.

2.13.10 Maintenance Material and Labor

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

2.13.11 Stack Degradation and Production Costs

SOFCs have the potential to operate over a long period of time as demonstrated in laboratory scale tests, where operation for over five years has been demonstrated without appreciable loss of performance. [19] However, with current planar stack technologies, stack performance has been observed to decline over its lifetime, generally due to an increase in the apparent electrical resistance of the stack associated with a variety of material and design related factors, which limits the permissible current at the same voltage (in a constant voltage operation mode). Performance degradation limits the operating lifetime of the capital intensive SOFC stack and

forms a significant component of the production costs. The present study adopts the methodology developed in the previous pathway study [20] to estimate the production costs.

2.13.12 Administrative and Support Labor

Labor administration and overhead charges are assessed at rate of 25 percent of the burdened operation and maintenance labor.

2.13.13 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 based on the quantity required using reference data.

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

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.

2.13.14 Waste Disposal

Waste quantities and disposal costs were estimated similarly to the consumables. In this study slag/ash and sorbents from the IGFC cases are considered a waste with a disposal cost of \$17.98/tonne (\$16.31/ton). The carbon used for trace element control in the IGFC cases is considered a hazardous waste with disposal cost of \$931/tonne (\$844/ton).

2.13.15 Co-products and Byproducts

Byproduct quantities were also determined similarly to the consumables. However, due to the variable marketability of these byproducts, specifically sulfur, no credit was taken for their potential salable value. Nor were any of the technologies penalized for their potential disposal cost. That is, for this evaluation, it is assumed that the byproduct or co-product value simply offset disposal costs, for a net zero in operating costs.

The different components of the O&M costs used in the present study are summarized in Exhibit 2-20 for convenience.

Exhibit 2-20. Summary of O&M Costs

Component	Specific Cost, 2018\$
Operating Labor Rate (\$/h)	38.5
Stack Replacement O&M (\$/h)	18.3
Water (\$/1000 gal)	1.90
Chemicals	
MU & WT Chemicals (\$/lb)	0.28
Carbon (Trace Removal) (\$/lb)	1.64
Selexol Solution (\$/gal)	38
Claus/DSRP Catalyst (\$/ft ³)	48
ZnO polishing sorbent (\$/lb)	1.51
KOH Coal Catalyst makeup (\$/lb)	0.19
Lime (\$/lb)	0.12
Waste Disposal	
Spent Trace Catalyst (\$/lb)	0.42
Ash + HCl Sorbent (\$/ton)	16.31
Spent sorbents (\$/lb)	0.42
Fuel	
Coal (\$/tons)	51.96
Natural Gas (\$/MMBtu)	4.42

2.13.16 Owner's Costs

The owner's costs included in the TOC estimate are shown in Exhibit 2-21.

Exhibit 2-21. Owner's Costs Included in TOC

Owner's Cost	Comprising
Preproduction Costs	• 6 months operating, maintenance, and administrative and support labor • 1-month maintenance materials • 1-month non-fuel consumables • 1 month of waste disposal costs • 25% of one month's fuel cost @ 100% capacity factor • 2% of TPC
Inventory Capital	• 60-day supply of fuel and consumables @ 100% capacity factor • 0.5% of TPC (spare parts)
Land	\$3,000/acre (300 acres for greenfield IGCC and PC)
Financing Costs	2.7% of TPC
Other Owner's Costs	15% of TPC
Initial Cost for Catalyst and Chemicals	All initial fills not included in BEC
Prepaid Royalties	Not included in owner's costs (included with BEC)
Property Taxes and Insurance	2% of TPC (Fixed O&M cost)
AFUDC and Escalation	TASC = TOC * 1.093

The category labeled "Other Owner's Costs" includes the following:

- Preliminary feasibility studies, including a Front-End Engineering Design (FEED) study
- Economic development (costs for incentivizing local collaboration and support)
- Construction and/or improvement of roads and/or railroad spurs outside of site boundary
- Legal fees
- Permitting costs
- Owner's engineering (staff paid by owner to give third-party advice and to help the owner oversee/evaluate the work of the EPC contractor and other contractors)
- Owner's contingency: sometimes called "management reserve," these are funds to cover costs relating to delayed startup, fluctuations in equipment costs, unplanned labor incentives in excess of a five-day/10-hours-per-day work week, etc.

Cost items excluded from "Other Owner's Costs" include:

- EPC Risk Premiums: Costs estimates are based on an engineering/procurement/construction management (EPCM) approach utilizing multiple subcontracts, in which the owner assumes project risks for performance, schedule, and

cost. This approach provides the owner with greater control of the project, while minimizing, if not eliminating, most of the risk premiums typically included in a lump-sum, “turnkey” EPC contract, under which the EPC contractor assumes some or all of the project risks. The EPCM approach used as the basis for the estimates here is anticipated to be the most cost-effective approach for the owner.

- Transmission interconnection: the cost of interconnecting with power transmission infrastructure beyond the plant busbar.
- Taxes on capital costs: all capital costs are assumed to be exempt from state and local taxes.
- Unusual site improvements: normal costs associated with improvements to the plant site are included in the bare erected cost, assuming that the site is level and requires no environmental remediation. Unusual costs associated with the following design parameters are excluded: floodplain considerations, existing soil/site conditions, water discharges and reuse, rainfall/snowfall criteria, seismic design, buildings/enclosures, fire protection, local code height requirements, noise regulations.

2.13.17 CO₂ Transport and Storage

The CO₂ T&S costs were calculated as \$10/tonne based on the Four-Basin study update. [21]

2.13.18 Breakeven Cost of CO₂ Captured

The breakeven cost of CO₂ captured was computed using the equation,

$$\text{Cost of CO}_2 \text{ Captured } (\$/\text{tonne CO}_2) = \frac{LCOE_{IGFC \text{ with CCS}} - LCOE_{IGFC \text{ without CCS}}}{CO_2 \text{ captured (tonnes/MWh)}}$$

where,

$LCOE_{IGFC \text{ with CCS}}$ is the LCOE of the IGFC system with CCS under consideration, and $LCOE_{IGFC \text{ without CCS}}$ is the LCOE of the comparable IGFC system without CCS.

3 IGFC PLANT MAJOR PROCESS AREAS

Two IGFC power plant scenarios with a series of pathway parameters, including evolving gasification technologies, were evaluated in the study. The Scenario 1 plant configuration uses the SOFC operated at atmospheric pressure. The Scenario 2 configuration utilizes advantages of SOFC operation at elevated pressures. The steam bottoming cycle represents a much smaller portion of the overall plant power generation relative to conventional fossil power plants, such as PC, IGCC, and NGCC power plants. The major process areas are described in detail in this section.

3.1 DESCRIPTIONS OF PROCESS AREAS

The IGFC plant, like the IGCC plant, consists of several integrated process areas, the primary ones being the coal receiving and storage area, the air separation unit, the gasification area, the gas cleaning area, the power island, and the CO₂ dehydration and compression area.

Descriptions of these areas and their selected technologies are presented in this report section; many of these plant areas have descriptions analogous to those used for IGCC in the BB report. [2] Additional case-specific information is presented in the relevant case sections.

3.1.1 Coal Receiving and Storage Area

The function of the Coal Receiving and Storage system is to unload, convey, prepare, and store the coal delivered to the plant. The boundary 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 for all the IGFC cases; however, coal preparation and feed are gasifier specific.

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 × 0 (3" × 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 transfers 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 on to the reclaim pile.

The reclaimers load the coal into two vibratory feeders located in the reclaim hopper under the pile. The feeders transfer the coal onto a conveyor belt that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3 cm × 0 (1¼" × 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.

3.1.2 Air Separation Unit

A cryogenic ASU is assumed to generate the total oxidant for use in three sections of the IGFC plant: the coal gasifier, the Claus sulfur recovery process, and the anode gas oxy-combustor. In this study, the ASU main air compressor discharge pressure was set to 0.5 MPa (79 psia),

providing oxygen product at sufficient pressure, 0.16 MPa (23 psia), to operate the oxy-combustor for the atmospheric pressure SOFC applications. The ASU is designed to generate 99.5 percent pure oxygen for IGFC applications to keep the nitrogen and argon content in the sequestered CO₂ stream low. Unlike IGCC plants, there is no opportunity for ASU air-side integration, as there is neither a requirement nor a benefit from syngas nitrogen dilution in the IGFC plant. In this study, the ASU nitrogen product was used only to provide an inert atmosphere, wherever applicable in the plant, and for transporting solids, with the remainder vented.

An air compressor providing air to the ASU is powered by an electric motor. Air to this 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 between each stage.

Air from the compressor is cooled and fed to an adsorbent-based pre-purifier system. The adsorbent removes water, carbon dioxide, 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 countercurrent to the normal airflow. The nitrogen is heated against extraction steam (1.7 MPa [250 psia]) in a shell and tube heat exchanger. 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 the 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, and 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 process steam at 0.3 MPa (50 psia) to drive the absorption refrigeration cycle. The remaining 5 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.

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 air feed before being warmed to ambient temperature. The gaseous oxygen exits the cold box and a portion is fed to the power island's oxy-combustor and Claus plant. The remainder of the oxygen is fed to the centrifugal compressor, with intercooling between each stage of compression. This compressed oxygen is then fed to the gasification unit.

3.1.3 Coal Gasification Area

Three gasification technologies were considered for this pathway: a conventional, entrained coal gasification technology; a conceptual, near-term, enhanced coal gasifier; and an advanced catalytic gasifier. The conventional coal gasifier technology selected for use in the IGFC plant is similar to the Conoco Phillips (CoP) E-Gas™ gasifier and can produce a syngas having a moderate CH₄ content of about 5.9 mole percent, which is higher than the CH₄ content of the syngas produced by either the GEE or Shell gasifier. The two-stage design allows for improved cold gas efficiency and lower oxygen consumption, but the quenched second stage retains some CH₄. The syngas CH₄ concentration exiting the gasifier is 5.9 volume percent (dry gas), compared to 0.10 volume percent for the GEE and 0.001 volume percent for the Shell gasifier. The presence of CH₄ in the syngas is expected to be beneficial to the IGFC plant performance because it reduces the excess cathode air flow needed for SOFC stack temperature control by providing local cooling of the SOFC stack through the endothermic reformation reaction.

A conceptual enhanced coal gasifier, having design features similar to the commercial CoP gasifier but operated to achieve a higher syngas methane content of about 10 mole percent was also considered to determine the potential benefits of developing and applying such a gasifier as part of the IGFC pathway.

3.1.3.1 Conventional Gasifier

The conventional, entrained, CoP E-Gas™ gasification technology represents one of the best conventional coal gasifier technologies for use in IGFC, with its ability to generate a syngas with a moderate methane content of approximately 6 mole percent. The design basis and performance estimates for the CoP gasifier were based on the BB report [2] and the assumptions are shown in Exhibit 3-1. The cold gas efficiency of the gasifier was estimated to be 81 percent (HHV).

The E-Gas™ two-stage coal gasification technology features an oxygen-blown, entrained-flow, refractory-lined gasifier with continuous slag removal. The Scenario 1 plant cases in this study utilize two parallel gasification trains to process Illinois No. 6 coal with both the gasifiers operating at maximum capacity.

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.

Exhibit 3-1. Coal Gasification Section Assumptions for the CoP E-Gas™ Gasifier

Gasifier	Specification/Assumptions
Technology	CoP 2-stage coal-water slurry
Number in parallel	2
Dried coal moisture, wt%	11.0 (as-received)
Coal feed type	coal-water slurry pumps
Oxygen-to-coal feed ratio	0.68
Slurry coal content, wt%	71
Steam-to-coal ratio	0.33
Steam temperature, °C (°F)	288 (550) saturated
Recycle gas-to-coal ratio	0.31
Recycle gas compressor eff., % (adiabatic)	85
Exit temperature, °C (°F)	999 (1830)
Exit pressure, MPa (psia)	3.10 (450)
Carbon loss with ash, % of coal carbon	0.8
Raw syngas composition basis	Equilibrium approach
Syngas methane content, vol% (dry)	5.9
Raw Syngas Cooler	
Technology	Fire-tube boiler
Number in parallel	2
Outlet temperature, °C (°F)	316 (600)

About 78 percent of the total slurry feed is fed to the first (or bottom) stage of the gasifier. 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 and devolatilization reactions in this stage reduce the final gas temperature to about 999°C (1,830°F).

The coal ash in the first stage 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. Char is produced in the second gasifier stage and is captured and recycled to the hotter first stage to be gasified.

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 a proprietary pressure letdown device.

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 study 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 transport the unloading station underneath the hopper and remove a quantity of slag for disposal. While the slag is suitable for use as a component of road paving mixtures, it was assumed in this study that the slag would be landfilled at a specified cost.

3.1.3.2 Enhanced, Conventional Gasifier

The estimated operating parameters of the enhanced conventional gasifier, which represents a conceptual extrapolation of the CoP gasifier based on gasifier enhancement activities [22], are listed in Exhibit 3-2. The cold gas efficiency is estimated to be 82.5 percent (HHV) for this gasifier, whose other features, including its stage-one characteristics, are expected to be very similar to the CoP E-Gas™ gasifier.

Exhibit 3-2. Coal Gasification Section Assumptions for the Enhanced Gasifier

	Specification/Assumptions
Gasifier	
Technology	2-stage coal-water slurry
Number in parallel	2
Dried coal moisture, wt%	11.0 (as-received)
Coal feed type	coal-water slurry pumps
Oxygen-to-coal feed ratio	0.61
Slurry coal content, wt%	71
Steam-to-coal ratio	0.33
Steam temperature, °C (°F)	288 (550) saturated
Recycle gas-to-coal ratio	0.31
Recycle gas compressor eff., % (adiabatic)	85
Exit temperature, °C (°F)	935 (1715)
Exit pressure, MPa (psia)	4.82 (700)
Carbon loss with ash, % of coal carbon	0.8
Raw syngas composition basis	Equilibrium approach
Syngas methane content, vol% (dry)	10.8
Raw Syngas Cooler	
Technology	Fire-tube boiler
Number in parallel	2
Outlet temperature, °C (°F)	316 (600)

3.1.3.3 Catalytic Gasifier

The partial-combustion of the coal and the loss of carbon with the ash constitute significant coal energy losses in the gasification of coal into the syngas. Catalytic coal gasification promotes efficient gasification of coal at a relatively low temperature where oxygen consumption is minimized and carbon conversion remains acceptably high, resulting in a gasifier cold gas efficiency that is high compared to conventional gasifiers. Under these conditions, especially if operated at high pressure, the methane content of the product syngas is also high, making it desirable for use with SOFC.

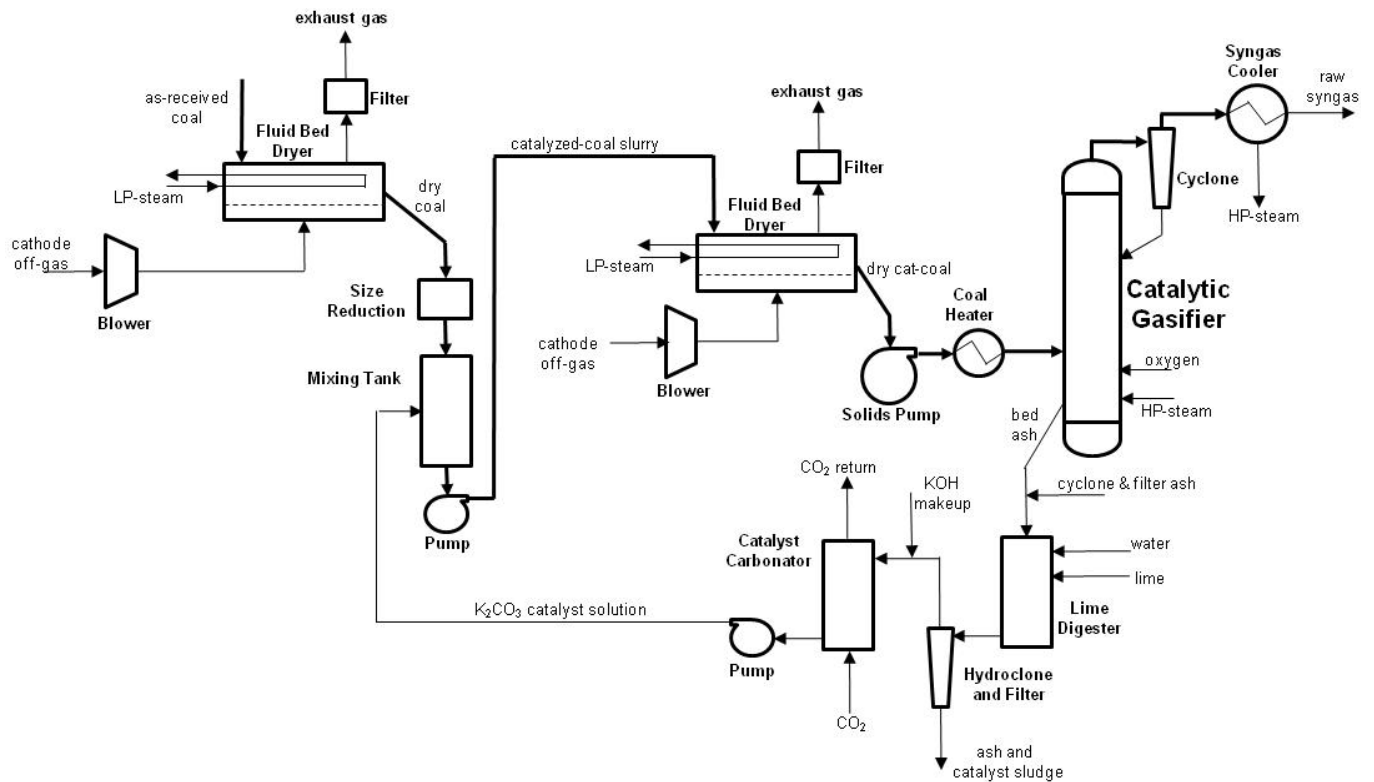
Catalytic coal gasification has not been tested beyond early developmental stages. It is assumed that the catalytic gasifier can be successfully developed for operation at the selected conditions, and with the performance estimated in this evaluation. While there is currently no ongoing development effort for this type of coal gasifier, an objective of this analysis is to assess the benefits of the catalytic gasification technology to justify future investments.

While a number of gasifier catalysts have been tested in laboratory studies, it has been found that the catalyst applied by Exxon (K_2CO_3 with KOH makeup) in their prior development program is very effective, but relatively expensive when compared to other, less effective catalysts. [23] The catalyst material, K_2CO_3 , is used as the primary catalyst in this evaluation, with KOH being the catalyst makeup form because of its lower cost.

The catalytic coal gasifier, assumed to use fluid bed contacting with steam and oxygen injection, was selected for the IGFC application because of its theoretical capability to efficiently generate a syngas with high methane content (approximately 30 mole percent). High-methane syngas is expected to promote more effective fuel cell cooling performance through internal SOFC methane reforming, leading to enhanced total plant efficiency due to a reduction in the needed cathode air rate that results. The demonstration of this enhanced capability using high-methane fuel in SOFC has not yet been completed.

Prior catalytic coal gasifier development by the Exxon Corporation was applied to a different fluid bed concept that used steam injection and recycle of a high-temperature stream of hydrogen-rich syngas, with the industrial application being synthetic natural gas production. [24] The design basis for the steam-oxygen catalytic gasifier applied in this evaluation was generated from thermodynamic equilibrium estimates for the gasifier operating at high pressure (975 psia exit pressure) and moderate temperature 704°C ($1,300^\circ\text{F}$), as well as from Exxon catalytic gasifier design assumptions for the coal-catalyst treatment and catalyst recovery processes. The estimated performance for this gasifier is supported by Exxon catalytic gasifier data, assuming a carbon loss of 5 weight percent of the coal feed carbon.

A general process diagram for the catalytic coal gasifier and its associated coal-catalyst treatment and catalyst recovery equipment is shown in Exhibit 3-3.

Exhibit 3-3. Catalytic Gasifier Coal/Catalyst Processing


As-received coal is first dried in a fluid bed dryer with warm cathode off-gas from the power island for fluidization, and with low-pressure steam for additional in-bed heating. [25] The dried coal is reduced in size and mixed with a K_2CO_3 catalyst solution. This slurry is then dried in a second fluid bed dryer similar to the first, again using warm cathode off-gas for fluidization, and LP-steam for in-bed heating. The processed coal is preheated to 149°C (300°F) using low-pressure steam or cathode off-gas indirect heating and is pressurized in a dry coal pump to the catalytic gasifier coal feed nozzles.

The gasifier ash and overhead fines are collected and are treated in a lime digester to release the catalyst from the ash constituents. The ash and catalyst sludge are separated from the slurry, and the catalyst solution is mixed with makeup catalyst (KOH). The catalyst solution is carbonated using a small portion of the plant CO_2 product. This step completes the recovery of the K_2CO_3 catalyst solution.

Details of the coal-catalyst processing steps assumed are as follows:

- Coal Catalyst Treatment:
 - Coal is crushed to -8 mesh (-2,380 microns or 0.0937 inch)
 - Coal is mixed with recycled catalyst solution (37 weight percent K_2CO_3)
 - The coal-catalyst solution is dried in fluid bed dryer at 54 °C (130 °F) using cathode off-gas and LP-steam heat source

- The process results in a coal catalyst loading of 15 weight percent K_2CO_3 (dry coal)
- Catalyst Recovery Factors:
 - First step is $Ca(OH)_2$ digestion plus water washing, operated at $149^\circ C$ ($300^\circ F$) with a mass ratio for Ca/K of 0.7 lb/lb
 - Soluble K recovery is 90 percent of the solids content to the digester
 - Solid/liquid separation is conducted using hydrocyclones
 - Overall catalyst recovery is 87 percent of the total loading
 - Catalyst makeup rate is 13 percent of the total catalyst feed rate
 - The makeup catalyst form is KOH
 - The recovered catalyst solution has 37 weight percent K_2CO_3 equivalent
- Gasifier Catalyst Reactions:
 - It is estimated that some of the K_2CO_3 catalyst decomposes in the gasifier, releasing CO_2
 - K_2O reacts with the char and ash, producing water-soluble and insoluble forms
 - An equivalent stream of CO_2 is recycled from the plant CO_2 product stream to the makeup catalyst carbonator vessel

The assumptions for the coal gasifier and the raw syngas cooler are listed Exhibit 3-4. It has been assumed in this study that the ash and catalyst mixture from the catalytic gasifier cases can be landfilled at the same per ton cost as the slag from the conventional coal gasifier.

Exhibit 3-4. Coal Gasification Section Assumptions with Catalytic Gasifier

Specification/Assumptions	
Gasifier	
Technology	Advanced steam-O ₂ catalytic
Number in parallel	1
Dried coal-catalyst moisture, wt%	5.5
Coal feed technology	Advanced dry feed pump
Coal-catalyst preheat temperature, °C (°F)	149 (300)
Oxygen-to-coal feed ratio	0.19
Steam-to-coal ratio	1.445
Steam temperature, °C (°F)	538 (1000)
Recycle gas-to-coal ratio	0
Exit temperature, °C (°F)	704 (1300)
Exit pressure, MPa (psia)	6.72 (975)
Carbon loss with ash, wt% of coal carbon	5
Raw syngas composition basis	Equilibrium
Syngas methane content, vol% (dry)	31.3
Raw Syngas Cooler	
Technology	Fire-tube boiler
Number in parallel	1
Outlet temperature, °C (°F)	427 (800)

The catalytic coal gasifier is a fluidized bed reactor contained within a cylindrical, refractory-lined pressure vessel. It is assumed to operate with a superficial velocity of 1.2 ft/s. The gas residence time is very long at about 100 seconds, resulting in a very deep bed. Coal, oxygen, and steam are introduced into the vessel with mixing conditions to avoid the creation of hot spots within the fluidized bed.

3.1.4 Syngas Cleaning Area

The function of the Gas Cleaning Area is to remove contaminants from the gasifier raw syngas to protect downstream equipment from damage, in addition to satisfying the environmental emission requirements. In IGFC systems, the acceptable levels of contaminants in the syngas for SOFC operation are more stringent than the environmental requirements.

All of the IGFC plant configuration cases utilize conventional, dry gas cleaning technology. A single-stage Selexol AGR technology, which is expected to generate a cleaner syngas than alternatives such as amine-based AGR, is employed in all the IGFC cases. The Selexol AGR is

preceded by carbonyl sulfide (COS) hydrolysis and by low-temperature, activated-carbon beds to remove mercury and other trace elements. The syngas from the Selexol AGR step is reheated to support a conventional zinc oxide (ZnO) polishing step to further remove the sulfur to levels (less than 100 ppbv of total sulfur) that are acceptable to the SOFC. The clean syngas is expanded to the required pressure and fed to the fuel cell as its anode feed gas.

A high-temperature barrier filter, a water scrubbing system, a COS hydrolysis unit, a low-temperature syngas cooling system, a trace element removal system, a Selexol single-stage acid gas removal process, a syngas reheat unit, and a ZnO fixed-bed sulfur-polishing unit comprises the Gas Cleaning Area illustrated schematically in Exhibit 3-5. The configuration of the Gas Cleaning Area is nearly identical for both the conventional and catalytic gasifier pathways, with minor differences in operating conditions. The reheating of the syngas for the subsequent ZnO polishing step is accomplished through high pressure (HP) steam indirect heating in the conventional gasifier cases, unlike the recuperative gas-to-gas heat exchange employed for the catalytic gasifier cases shown in Exhibit 3-5.

Exhibit 3-5. Syngas Cleaning for IGFC

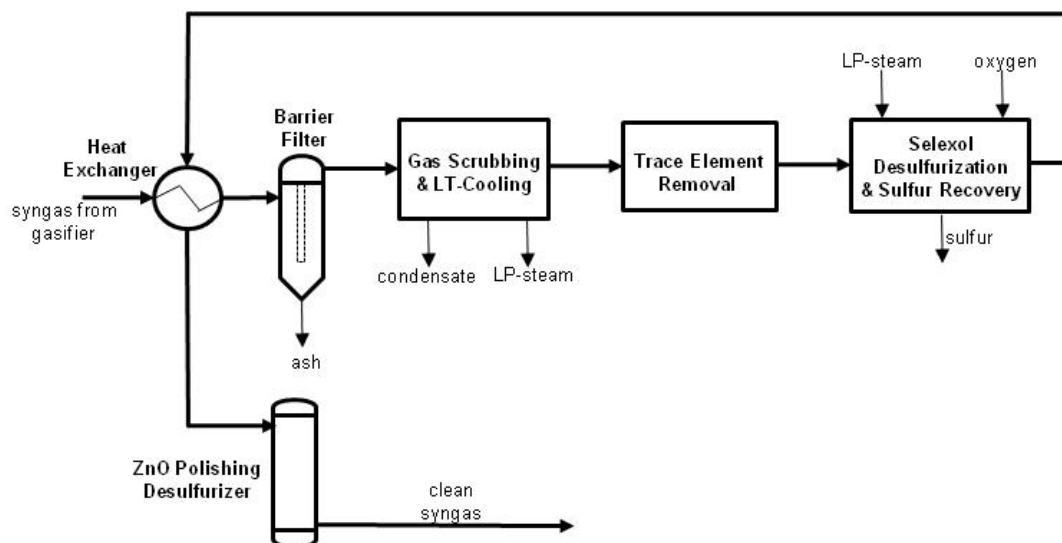


Exhibit 3-6 summarizes the major syngas cleaning section assumptions and specifications. The inherent assumption in this evaluation is that the coal syngas subsequent to the listed cleaning steps is acceptable to the SOFC unit for long-term operation. This long-term success has not yet been demonstrated.

Exhibit 3-6. Gas Cleaning Area Assumptions for the Conventional Gasifier Cases

Specification/Assumptions	
Gas cleaning technology	Conventional dry gas cleaning
Number of parallel trains	2
Particulate removal	Barrier filter at 316 °C (600 °F)
HCl removal	Water scrubber
Ammonia removal	Low-temperature gas cooling to 35 °C (95 °F)
Hg, As, Se, Cd, P	Activated-carbon fixed beds at 35 °C (95 °F)
Bulk desulfurization	Selexol at 35 °C (95 °F)
Sulfur recovery	Conventional Claus plant with tail gas recycle
Polishing desulfurization	ZnO fixed beds at 316 °C (600 °F)
Syngas preheating source	HP-steam heating for CoP gasifier

The gas cleaning area is modified slightly for the catalytic gasifier configuration, as indicated in Exhibit 3-7. The particulate removal temperature has been increased to 427°C (800°F), and zinc oxidize syngas polishing temperature has been increased to 371°C (700°F). Clean syngas is reheated to 371°C (700°F) for sulfur polishing by a gas-gas recuperative heat exchange, which cools the gasifier syngas to 232°C (450°F).

Exhibit 3-7. Gas Cleaning Section Assumptions for the Catalytic Gasifier

Specification/Assumptions	
Gas cleaning technology	Conventional dry gas cleaning
Number parallel trains	1
Particulate removal	Barrier filter at 371 °C (700 °F)
HCl removal	Water scrubber
Ammonia removal	Low-temperature gas cooling to 35 °C (95 °F)
Hg, As, Se, Cd, P	Activated-Carbon fixed beds at 35 °C (95 °F)
Bulk desulfurization	Selexol at 35 °C (95 °F)
Sulfur recovery	Conventional Claus plant with tail gas recycle
Polishing desulfurization	ZnO fixed beds at 371 °C (700 °F) to 100 ppbv total sulfur
Syngas preheating source	Syngas recuperation

3.1.4.1 Raw Gas Cooling/Particulate Removal

The raw syngas from the gasifier is cooled to its desired temperature in the syngas cooler unit, which consists of a fire-tube boiler and convective superheating and economizing sections. 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, high-pressure synthesis gas in relatively small tubes.

The cooled gas from the syngas cooler 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. In the cases using the conventional gasifier, the recovered fines are pneumatically returned to the first stage of the gasifier. The recycled char and recycled particulate results in high overall carbon conversion.

The syngas, after particulate removal, is cooled to 232°C (450°F) through heat exchange with the steam cycle before entering the syngas scrubber in conventional gasifier cases.

3.1.4.2 Syngas Scrubber and Low-temperature Cooling Section

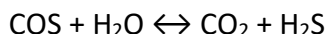
The cooled syngas passes to a syngas scrubber, where a water wash is primarily used to remove chlorides and any particulate that might have penetrated the barrier filter. The syngas exits the scrubber saturated at about 169°C (337°F). This is followed by low-temperature cooling to 35°C (95°F), primarily to remove NH₃ and generate condensate streams.

The sour water stripper removes NH₃, H₂S, and other impurities from the scrubber and other condensate streams. The stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from synthesis gas coolers. 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 sulfur recovery unit. Remaining water is sent to wastewater treatment.

3.1.4.3 COS Hydrolysis

The COS hydrolysis pretreatment provides a means to reduce the COS concentration in the feed to the acid gas removal process. Several catalyst manufacturers, including Haldor Topsoe and Porocel, offer a catalyst that promotes the COS hydrolysis reaction. Syngas exiting the scrubber is reheated to about 186°C (367°F) by using HP steam from the HRSG prior to entering a COS hydrolysis reactor. About 99.5 percent of the COS is converted to CO₂ and H₂O.

The equimolar COS hydrolysis reaction is represented as follows.



Although the slightly exothermic reaction favors higher conversion at lower temperatures, the reaction kinetics are slower at lower temperatures. 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 hydrogen sulfide (H₂S) and carbon monoxide (CO) to form COS and H₂.

Although the reaction itself is exothermic, the heat of the reaction is dissipated among the large amount of non-reacting components, and 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.

3.1.4.4 Trace Removal

The gas exiting the COS reactor passes through a series of heat exchangers and knockout drums to lower the syngas temperature to 35°C (95°F) and to separate entrained water. The cooled syngas then passes through a carbon bed to remove 95 percent of the Hg and other trace metals.

A conceptual design for an activated, sulfur-impregnated, carbon bed adsorption system was developed for mercury control in the IGCC plants being studied. 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. [26] IGFC-specific design considerations are discussed below.

The packed carbon bed vessels are located upstream of the Selexol acid gas removal unit, where syngas enters at a temperature near 38°C (100°F). Eastman Chemical also operates their beds ahead of their acid gas removal unit at a temperature of 30°C (86°F). [27]

An empty-vessel-basis gas residence time of approximately 20 seconds was used based on Eastman Chemical's experience. 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 and was consequently selected for this application.

The bed density was assumed to be 30 lb/ft³ based on the Calgon Carbon Corporation HGR-P sulfur-impregnated pelletized activated carbon. [26] These parameters determined the size of the vessels and the amount of carbon required. The gasifier train has one mercury removal step.

Eastman Chemicals replaces its bed every 18 to 24 months. However, bed replacement is not because of mercury loading, but for other reasons, including buildup of pressure drop, water, and other contaminants in the bed.

For this study, 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 percent). Mercury capacity of sulfur-impregnated carbon can be as high as 20 wt percent. [28] The mercury-laden carbon is considered to be a hazardous waste, and the disposal cost estimate reflects this categorization.

It is assumed that other trace species, such as arsenic, selenium, cadmium, and phosphorus will also be effectively removed by this unit.

3.1.4.5 Acid gas Removal Process

A key function of syngas cleaning is acid gas removal with sulfur recovery. The total sulfur content of the syngas is reduced to less than 30 ppmv, including all sulfur species, in particular the total of COS and H₂S. The Selexol process was chosen for AGR in all the pathways based on

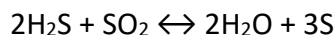
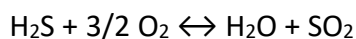
the gasifier operation at high pressure, which favors the physical solvent dimethyl ether of polyethylene glycol, used in the Selexol process. [29]

Cool, particulate-free syngas enters the Selexol absorber unit at approximately 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 H₂. 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 removed 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 flash pressure reduction in a series of flash vessels. The stripper acid gas stream, consisting of H₂S and CO₂, with some N₂, is then sent to the Claus unit.

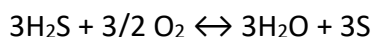
3.1.4.6 Sulfur Recovery/Tail Gas Cleanup Process

The conventional three-stage Claus plant, with indirect reheat and feeds with a high H₂S content, exceeds 98 percent sulfur recovery efficiency. [29]

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.

One-third of the H₂S is burned in the furnace with oxygen to give sufficient SO₂ to react with the remaining H₂S. Since these reactions are highly exothermic, a waste heat boiler that recovers this heat is needed to generate high-pressure steam following the furnace. Sulfur is condensed in a condenser that follows the high-pressure steam recovery section. Low-pressure 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 tail gas treating unit (TGTU).

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 from 1093 to 1427°C (2000 to 2600°F), and as the temperature decreases, conversion increases dramatically. Claus plant sulfur recovery efficiency depends on many factors, such as H₂S concentration of the feed gas, the number of catalytic stages, and the gas reheat method. In many refinery and other conventional Claus applications, tail gas treating involves the removal of the remaining sulfur compounds from gases exiting the

sulfur recovery unit. Tail gas from a typical Claus process contains small-but-varying quantities of COS, CS₂, H₂S, SO₂, and elemental sulfur vapors. In addition, there is some H₂, CO, and CO₂ in the tail gas. To remove the rest of the sulfur compounds from the tail gas, all the sulfur-bearing species must first be converted to H₂S. The resulting H₂S is absorbed into a solvent and the clean gas vented or recycled for further processing. In all the IGFC cases, the Claus plant tail gas is hydrogenated, water is separated, and tail gas is compressed and is then returned to the AGR process for further treatment.

3.1.4.7 Sulfur Polishing

Several commercial sorbents are available for syngas sulfur polishing. Zinc oxide-based sorbents, having one of the highest affinities for hydrogen sulfide removal, are applicable for desulfurization to levels less than 100 ppbv and are offered by several catalyst vendors.

They operate at relatively high temperatures—260–427°C (500–800°F)—and are typically applied in batch-operated, packed bed vessels. These vessels are normally operated with syngas downflow through the packed bed, and the packed bed is supported on a ceramic or metal syngas distribution device that promotes uniform syngas flow through the bed and maintains gas velocities at the distributor low enough to prevent sorbent particle attrition. The sorbents are manufactured with sizes that allow reasonable gas velocities through the beds with acceptable pressure drops. The sorbent particles have pore structures that provide rapid reaction conditions so that a distinct reaction front moves through the bed. When sulfur breakthrough is approached in the bed, or when the bed pressure drop becomes excessive, the vessel is taken out of service, drained, and refilled with fresh sorbent. The bulk desulfurized syngas from the Selexol unit is preheated by gas-to-gas heat exchange with the warm syngas from the barrier filter, or by indirect steam heating with high-pressure steam.

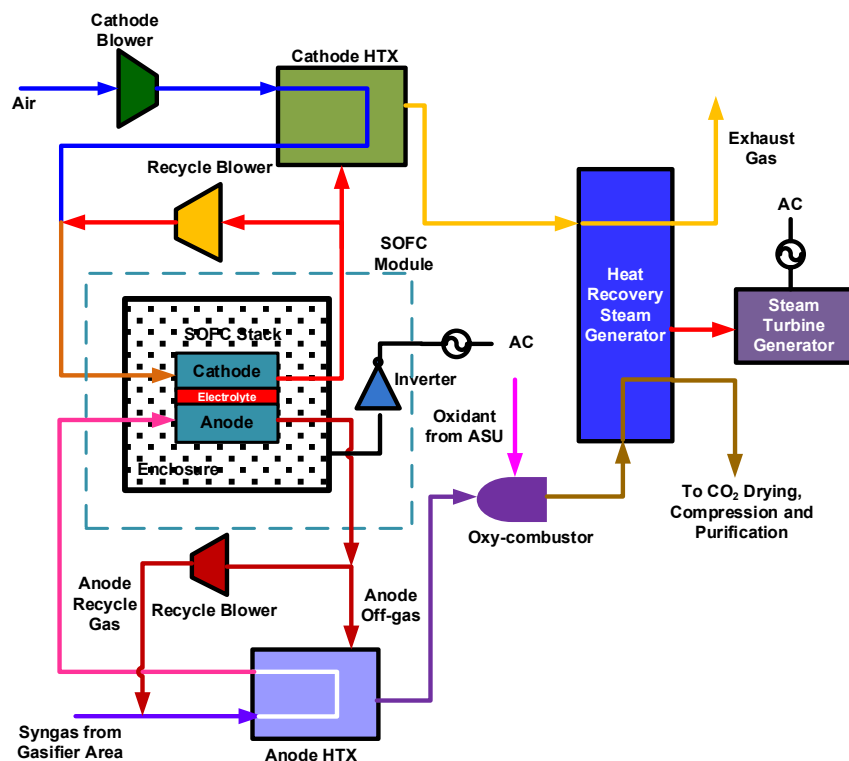
3.1.5 IGFC Power Island

The IGFC power island for Scenario 1 with atmospheric SOFC operation is shown in Exhibit 3-8 and consists of the SOFC unit with DC-AC inverter, a syngas expander, an anode off-gas combustor, an HRSG that captures heat from the combusted anode off-gas, and a steam bottoming cycle. The primary difference between systems with and without CCS is the utilization of the vitiated cathode gas as the oxidant for the combustion of the anode off-gas in the steam bottoming cycle.

The IGFC power island for the pressurized Scenario 2, depicted in Exhibit 3-9, includes, in addition, a cathode gas expander, and, in some cases of a system with CCS, an oxy-combustor expander. An option to inject NG into the combustor to boost the bottoming cycle is also included for the cases without CCS.

Exhibit 3-8. IGFC Power Island – Scenario 1 – Atmospheric SOFC

(a) With CCS



(b) Without CCS

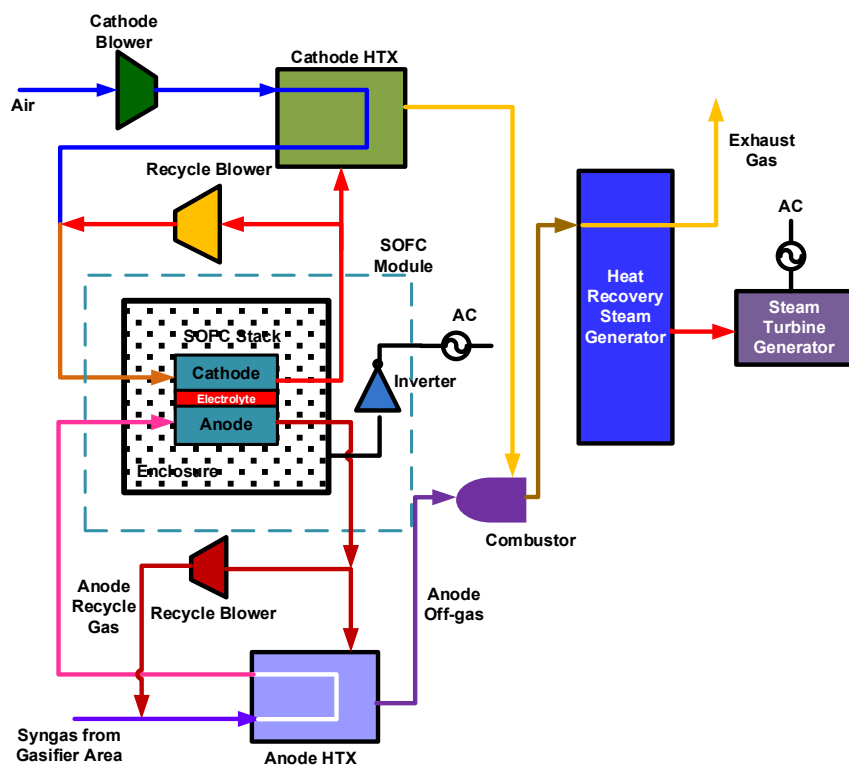
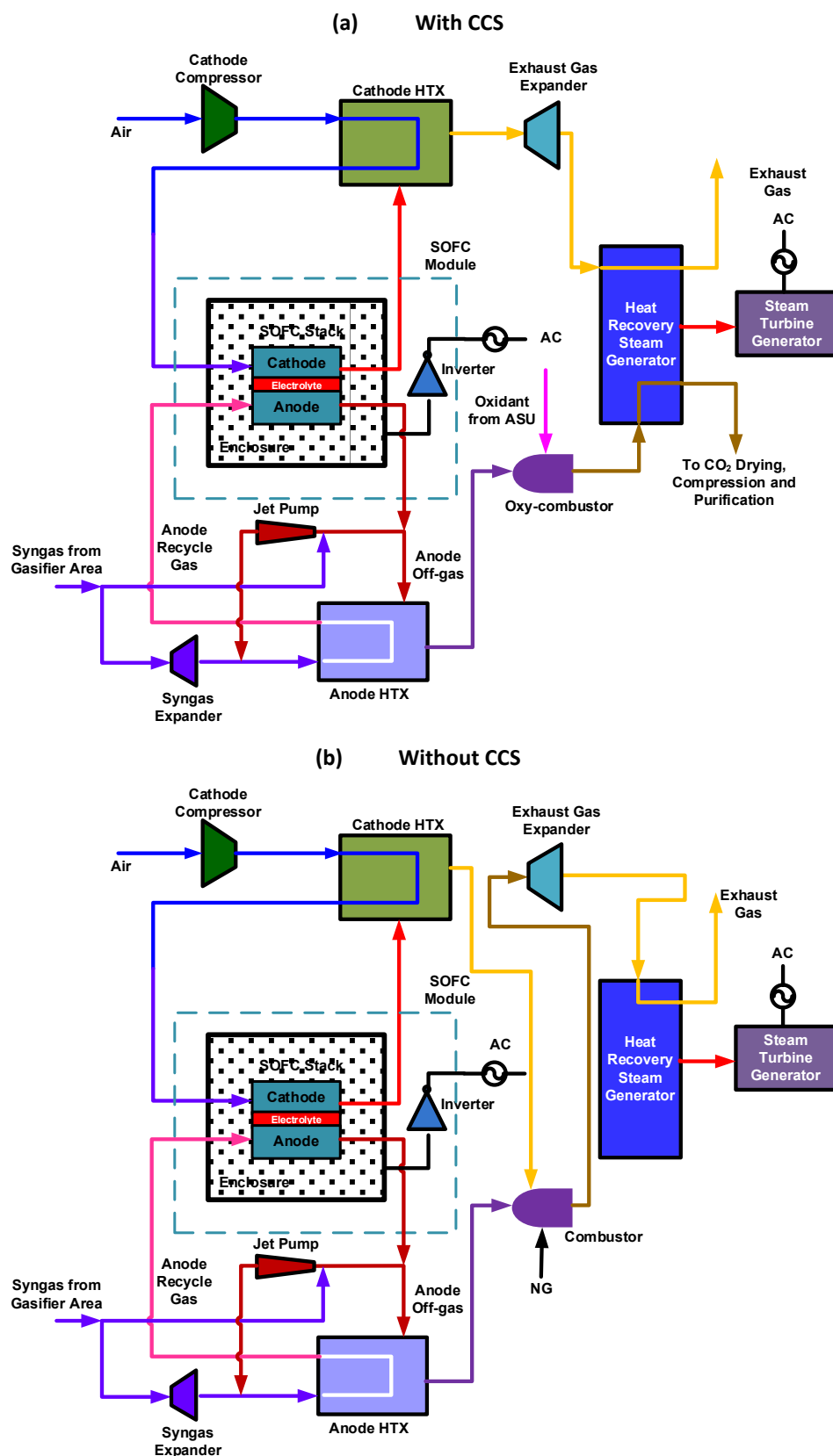


Exhibit 3-9. IGFC Power Island – Scenario 2 – Pressurized SOFC



3.1.5.1 SOFC Unit

The SOFC unit ancillary components consist of cathode air blowers, cathode heat exchangers that recuperatively heat the cathode air up to the fuel cell inlet temperature, cathode hot gas recycle blowers, anode heat exchangers that recuperatively heat the anode gas up to the fuel cell inlet temperature, and anode hot gas recycle blowers. Hot gas blowers capable of operation at the required conditions of the anode and cathode recycle gas streams have been under development for the last decade. [30]

The HRSG produces low-pressure and high-pressure process steam, as well as high-pressure power steam for the subcritical steam bottoming cycle. The cooling water system uses a mechanical draft, wet cooling tower arrangement.

For the operation of SOFC at elevated pressures assumed in Scenario 2, the cathode air is compressed to the pressurized fuel gas inlet pressure, and no cathode gas recycle is used. The cathode off-gas is expanded to atmospheric pressure to generate power to drive the cathode gas compressor. Anode gas recycle is accomplished using a syngas-driven jet pump in this pressurized case.

The major assumptions for the SOFC power island for the pathway cases are listed in Exhibit 3-10. The anode recycle rate was adjusted to maintain a total O₂-to-carbon atomic ratio of at least 2.1 in the anode inlet gas in all the study cases to avoid carbon deposition in the fuel cell.

The anode off-gas is combusted using O₂ in an advanced oxy-combustor with excess O₂ limited to 1 mole percent. It is assumed that an anode off-gas oxy-combustor can be developed that can operate stably with 1 mole percent excess O₂.

The anode off-gas is combusted using oxygen in an advanced oxy-combustor with excess oxygen limited to 1 mole percent. It is assumed that an anode off-gas oxy-combustor can be developed that can operate stably with 1 mole percent excess oxygen.

The combusted anode gas which consists of CO₂, one mole percent excess oxygen, water vapor, and minor traces of syngas contaminants (e.g., sulfur species, HCl, NO_x, trace elements) is directed towards the CO₂ drying, purification, and compression unit discussed next.

Exhibit 3-10. IGFC Power Island Reference Case Assumptions

Parameters	Scenario 1	Scenario 2
Natural Gas/Syngas Expander		
Outlet pressure, MPa (psia)	0.14 (20)	0.34 (50)
Efficiency, adiabatic %	90	
Generator efficiency (%)	98	
Fuel Cell System		
Current density, mA/cm ²	400	
Reference case stack fuel utilization, %	70	
Cell stack inlet pressure, MPa (psia)	0.12 (15.1)	2.2 (285)
Reference case overall fuel utilization, %	80	
Oxygen-to-carbon ratio	2.1	
Reference case stack fuel utilization, %	80	
Cell stack inlet temperature, °C (°F)	Determined from PNNL model	
Cell stack outlet temperature, °C (°F)	Computed in Aspen®	
Stack anode-side pressure drop, MPa (psi)	0.00062 (0.11)	
Fuel Cell System Ancillary Components		
Anode gas recycle method	Hot gas fan	Fuel gas jet pump
Anode recycle gas fan efficiency, adiabatic %	80	N/A
Anode heat exchanger pressure drop, MPa (psi)	0.0014 (0.2)	
Cathode gas recycle method	Hot gas fan	None
Cathode recycle gas rate, %	50	0
Cathode recycle gas fan eff., adiabatic %	80	N/A
Cathode heat exchanger pressure drop, MPa (psi)	0.0055 (0.8)	
Cathode blower/compressor eff., adiabatic %	82	90
Cathode gas expander efficiency, adiabatic %	N/A	95
Reference rectifier DC-to-AC efficiency, %	97	
Recycle blower motor drives eff., %	87.6	N/A
Other electric motor drives efficiency, %	95	95
Transformer efficiency, %	99.65	99.65

3.1.5.2 Natural Gas Injection

In the pressurized cases without CCS, natural gas is injected (at the combustor pressure after throttling down from the supply pressure of 500 psia) into the combustor to ensure high enough turbine inlet temperature to the cathode expander to boost the bottoming Brayton cycle, which is desirable from an IGFC performance perspective. A natural gas price of \$4.42/MMBtu was used in estimating the costs.

3.1.6 Heat Recovery Steam Generator and Steam Bottoming Cycle

The BB report [2] provides a detailed description of the HRSG and steam power cycle process configuration and equipment in typical IGCC and NGCC power plants. The HRSG and steam power cycle for the IGFC cases evaluated in this report are expected to be similar in configuration and operating conditions to the comparable IGCC and NGCC systems. Only simplified simulation of the steam system was conducted in this evaluation, as described in Section 2.9.3.

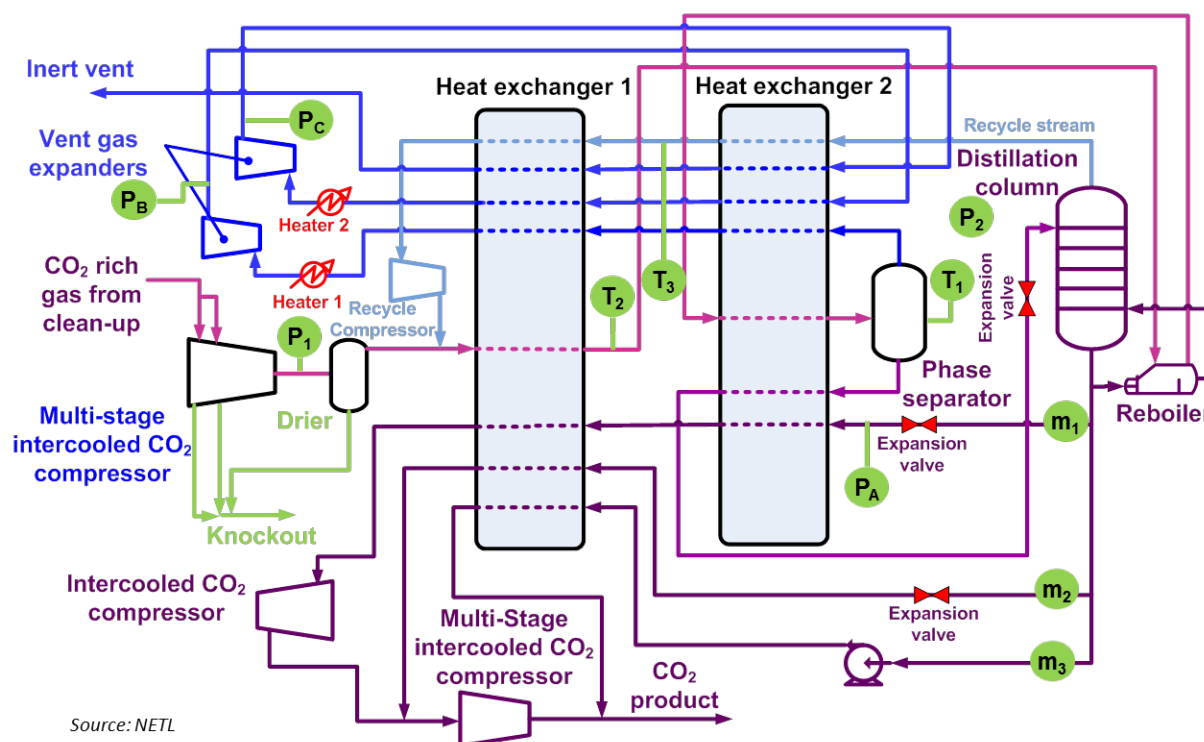
3.1.7 CO₂ Compression, Drying, and Purification

After completion of heat recovery, the oxy-combustion off-gas stream is sent to the CPU for compression, drying, and purification to pipeline and EOR specifications.^{g, h} An auto-refrigerated CPU process based on the process developed for the advanced oxy-combustion studies [31] is utilized in the present study.

Exhibit 3-11 depicts the block diagram of the auto-refrigerated CPU process, which was developed based on the patents issued to Air Products and Chemicals, Inc. [32] and to Air Liquide. [33] Wet CO₂-rich gas from the flue gas desulfurizer is compressed to an initial pressure, P_1 , (≈ 30 bar/435 psia) and is dried to remove moisture. The dried CO₂-rich gas is purified to the specified levels by liquefaction, followed by flash phase separation and distillation as in the external refrigeration process. However, cooling of the dried CO₂ gas down to a temperature, T_1 ($\approx 60^\circ\text{F}$) is accomplished entirely through heat exchange with the cold non-condensables and the pure CO₂ product, part of which is expanded to satisfy the refrigeration needs, in heat exchangers HX1 and HX2 along with the reboiler of the distillation column. The reboiler boilup ratio of the distiller can be tuned to achieve the required CO₂ purity.

^g The QGESS reference [39] recommends a purity of the CO₂ at least 95 percent, as a conceptual design basis for EOR purposes; however, distillation methods used to meet the more stringent O₂ concentration requirement of 100 ppmv or less generally result in 99.9 percent + CO₂ purity.

^h An O₂ concentration of 10 ppmv in the CO₂ product stream was selected as the basis for conceptual design since it represents the lower limit of the range of values recommended in literature [39] for EOR applications. The number of distillation stages can be reduced slightly to design to the upper limit of 100 ppmv for O₂ concentration. [39] However, the impact of the associated small decrement in distillation cost on the overall cost is expected to be insignificant.

Exhibit 3-11. Auto-Refrigerated CPU Process Developed for the Present Study

3.1.8 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.

3.1.9 Instrumentation and Control

An integrated plant-wide DCS is provided. The DCS is a redundant microprocessor-based, functionally distributed control system. The control room houses an array of multiple video monitors and keyboard units. The monitors/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 and is designed to be operational and accessible 99.5 percent of the time 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. The 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, which is highly dependent on stoichiometry, temperature, and pressure.

4 SCENARIO 1 – IGFC WITH ATMOSPHERIC-PRESSURE SOFC

Pressurized clean syngas from the gasifier is expanded to near-atmospheric pressures to fuel an atmospheric SOFC system in Scenario 1. A total of seven cases, which featured systems with and without carbon capture were analyzed as per Exhibit 1-4 and Exhibit 1-5 for this scenario. These included three carbon capture cases that took advantage of the VGR concept that enables high fuel utilization. Detailed results for only salient cases (Reference case with CCS, Case 3 without CCC and Case 6 with VGR) that span the scenario are included in this section, along with a summary of results for all the cases.

4.1 REFERENCE ATMOSPHERIC IGFC

The reference case uses the conventional E-Gas™ gasifier combined with an SOFC unit operating at atmospheric pressure. The Coal Gasification Area contains the coal preparation system, the slag handling system, the coal water-slurry feeding system, the coal gasification system, the air separation system, and the raw syngas cooling system. The Gas Cleaning Area uses conventional dry gas cleaning technology based on single-stage Selexol acid gas removal, together with a ZnO fixed-bed sulfur-polishing unit to supply clean syngas to the SOFC power island. The reference atmospheric pressure SOFC technology parameters are listed in Exhibit 4-1.

Exhibit 4-1. IGFC Parameters Used for the Atmospheric Reference Case

Parameter	Value
Gasifier Technology	Conventional
Syngas CH ₄ content (mole %)	5.8
SOFC Operating Pressure	Atmospheric
Cell Technology	Advanced Cell
Fuel Utilization, %	80
Current Density, mA/cm ²	400
Degradation, %/1000 h	0.2
Inverter Efficiency (%)	97
Stack Cost (\$/kW)	225
Capacity Factor (%)	80

The relevant process data for the numbered streams in the block flow diagram (BFD) for the reference plant with CCS shown in Exhibit 4-2 are tabulated in Exhibit 4-3. The syngas methane content entering the SOFC section is ≈5.8 mole percent. The BFD for the reference plant without CCS, shown in Exhibit 4-4, differs only downstream of the SOFC island—the combustor is fed by the vitiated air instead of oxygen from ASU.

Atmospheric Reference Plant Performance

The performance summary listed in Exhibit 4-5 for the reference plant shows that the reference plant with CCS results in a net plant HHV efficiency of ≈ 43.0 percent, which is much higher than the values achievable with conventional coal power plant technologies. The carbon capture rate for the process is 98.8 percent. The reference plant without CCS has an efficiency of ≈ 47.5 percent, which implies that the introduction of CCS imposes a performance penalty of ≈ 4.5 percentage points.

For both systems with and without CCS, the SOFC power island generates 70 percent of the gross power, followed by the steam cycle at 25 percent and the syngas expander at 5 percent, as seen in Exhibit 4-6. The addition of the CPU and the increased ASU oxygen demand, which together contribute to nearly 64 percent of the auxiliary load as shown in Exhibit 4-7, nearly double the auxiliary load demand for the reference plant with CCS plant relative to the plant without CCS.

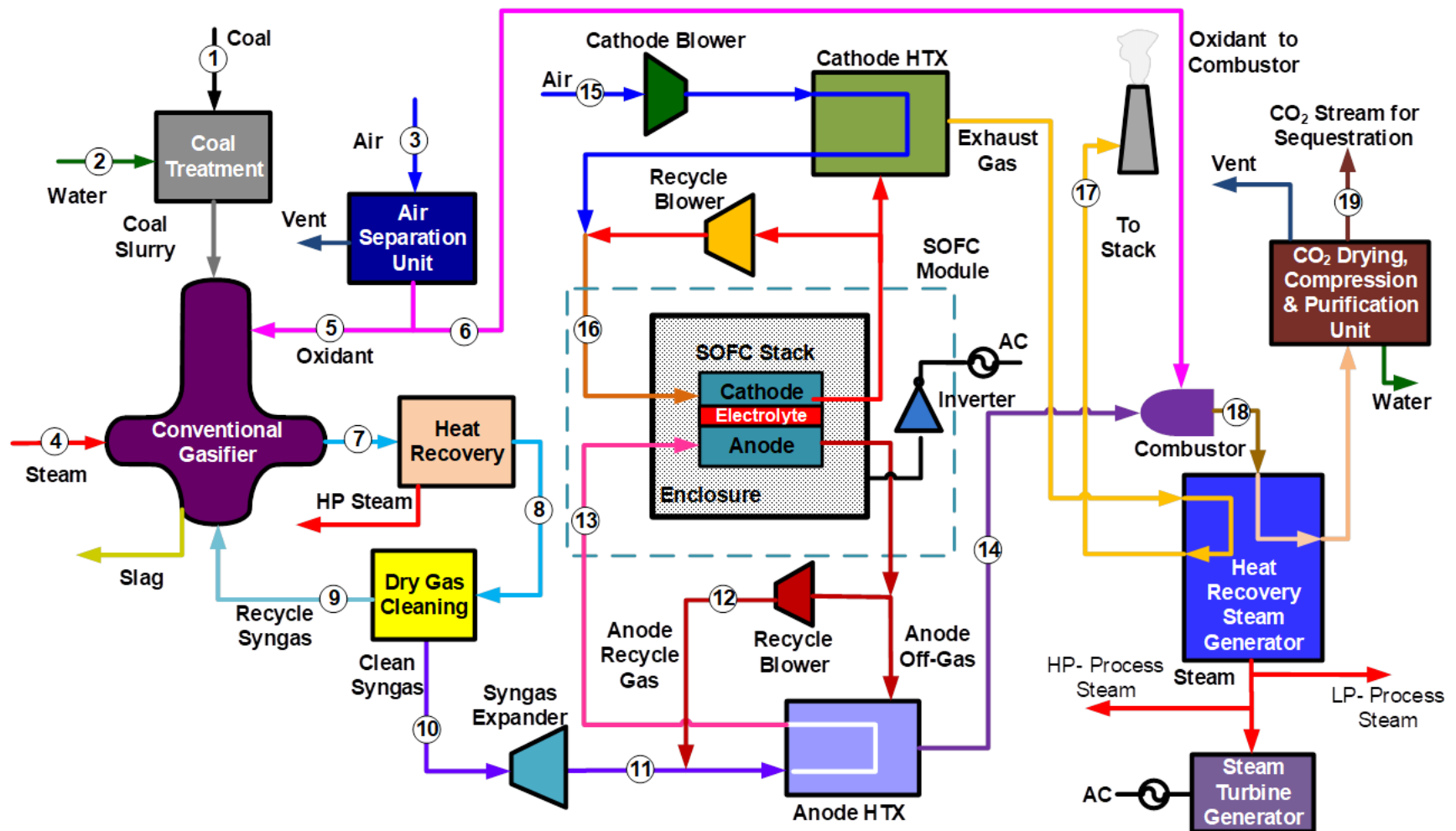
The salient material and energy balance for the reference plant with CCS is listed in Exhibit 4-8, which also shows that the IGFC plant acts as a nearly zero emission power plant, with only a small release of CO_2 shown in the emissions. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination. The nearly complete recovery of water from the oxy-combustion CO_2 product stream results in an IGFC plant water consumption significantly lower than the corresponding value for conventional fossil fuel power plant technologies. The heat and mass balance diagram for the gasifier, ASU, and syngas clean-up is shown in Exhibit 4-9, while the corresponding process flow diagrams (PFD) for the power island and the CPU are shown in Exhibit 4-10 and Exhibit 4-11, respectively.

Atmospheric Reference Plant Costs

The capital and O&M costs for the atmospheric reference plant with CCS are itemized in Exhibit 4-12 and Exhibit 4-13, respectively. The gasifier, syngas cooler and cleanup costs account for over 60 percent of the capital costs, as shown in Exhibit 4-14. The gasifier costs are nearly two times the costs used in the previous pathway and are based on the BB report that considers the recent installations. Any cost reduction in the gasifier systems would greatly benefit the overall cost of the IGFC system. The cathode and anode heat exchangers and the oxy-combustor contribute to 75 percent of the BOP costs, as shown in Exhibit 4-15.

The LCOE for the reference plant is dominated by the capital cost component (≈ 57 percent), as shown in Exhibit 4-16. The LCOE for the reference plant with CCS is $\$97.5/\text{MWh}$, only 10 percent higher than the LCOE for the plant without CCS, indicating a much lower CCS cost penalty relative to conventional systems, which could be as high as 60 percent.

Exhibit 4-2. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Process Diagram



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-3. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
V-L Mole Fraction																			
Ar	0.0000	0.0000	0.0094	0.0000	0.0031	0.0031	0.0006	0.0006	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0094	0.0096	0.0099	0.0010	0.0000
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0443	0.0443	0.0577	0.0583	0.0583	0.0001	0.0150	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2868	0.2868	0.3737	0.3775	0.3775	0.0773	0.1545	0.0773	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.1549	0.1549	0.2022	0.2043	0.2043	0.4959	0.4210	0.4959	0.0003	0.0003	0.0003	0.5673	1.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003	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	0.2675	0.2675	0.3486	0.3519	0.3519	0.0948	0.1609	0.0948	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0000	1.0000	0.0104	1.0000	0.0000	0.0000	0.2315	0.2315	0.0023	0.0013	0.0013	0.3258	0.2424	0.3258	0.0104	0.0107	0.0109	0.4163	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0010	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	0.0000	0.0000	0.0069	0.0069	0.0094	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N ₂	0.0000	0.0000	0.7722	0.0000	0.0019	0.0019	0.0020	0.0020	0.0027	0.0060	0.0060	0.0053	0.0055	0.0053	0.7722	0.7916	0.8121	0.0055	0.0000
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0044	0.0044	0.0017	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	0.2077	0.0000	0.9950	0.9950	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.1878	0.1668	0.0099	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	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	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	0	4,576	30,782	3,671	4,285	1,781	25,558	25,558	2,942	16,506	16,506	47,717	64,223	18,427	129,419	252,484	123,065	18,622	10,433
V-L Flowrate (kg/hr)	0	82,435	888,183	66,137	137,174	57,027	529,270	529,270	63,336	353,761	353,761	1,442,587	1,796,333	557,088	3,734,297	7,265,252	3,530,954	614,115	459,148
Solids Flowrate (kg/hr)	200,669	0	0	0	0	0	187	187	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	148	15	288	130	27	999	232	78	316	49	699	649	575	15	637	167	1,419	37
Pressure (MPa, abs)	0.10	0.69	0.10	3.45	3.45	0.16	3.10	2.96	4.14	2.32	0.14	0.14	0.14	0.1	0.101	0.105	0.101	0.130	15.272
Enthalpy (kJ/kg) ^A	---	566.85	31.06	2,973.90	114.27	23.91	2,286.82	859.91	103.53	484.63	72.66	1,421.79	1,289.90	1,238.9	31.057	693.514	187.613	2,599.637	-207.311
Density (kg/m ³)	---	864.7	1.2	14.8	33.1	2.0	6.0	14.7	30.6	10.1	1.1	0.5	0.5	0.6	1.2	0.4	0.8	0.3	539.3
V-L Molecular Weight	---	18.015	28.854	18.015	32.016	32.016	20.709	20.709	21.526	21.432	21.432	30.232	27.970	30	28.854	28.775	28.692	32.977	44.010
V-L Flowrate (lb _{mol} /hr)	0	10,088	67,862	8,094	9,446	3,927	56,346	56,346	6,487	36,390	36,390	105,198	141,587	40,625	285,321	556,632	271,311	41,055	23,001
V-L Flowrate (lb/hr)	0	181,738	1,958,108	145,808	302,418	125,722	1,166,840	1,166,840	139,632	779,909	779,909	3,180,361	3,960,236	1,228,169	8,232,717	16,017,139	7,784,422	1,353,891	1,012,249
Solids Flowrate (lb/hr)	442,400	0	0	0	0	0	412	412	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	299	59	550	265	80	1,830	450	173	600	120	1,290	1,200	1,066	59	1,179	333	2,586	98
Pressure (psia)	14.7	100.0	14.7	500.0	500.0	23.0	450.0	430.0	600.0	337.0	20.0	20.0	20.0	19.9	14.7	15.3	14.7	18.9	2,215.0
Enthalpy (Btu/lb) ^A	---	243.7	13.4	1,278.5	49.1	10.3	983.2	369.7	44.5	208.4	31.2	611.3	554.6	532.6	13.4	298.2	80.7	1,117.6	-89.1
Density (lb/ft ³)	---	53.983	0.076	0.923	2.067	0.127	0.377	0.918	1.910	0.631	0.069	0.032	0.031	0	0.076	0.025	0.050	0.019	33.666

^A Reference conditions are 32.02 F and 0.089 psia

Exhibit 4-4. Case ACONVIGFC0A – Atmospheric Reference IGFC Plant without CCS – Process Diagram

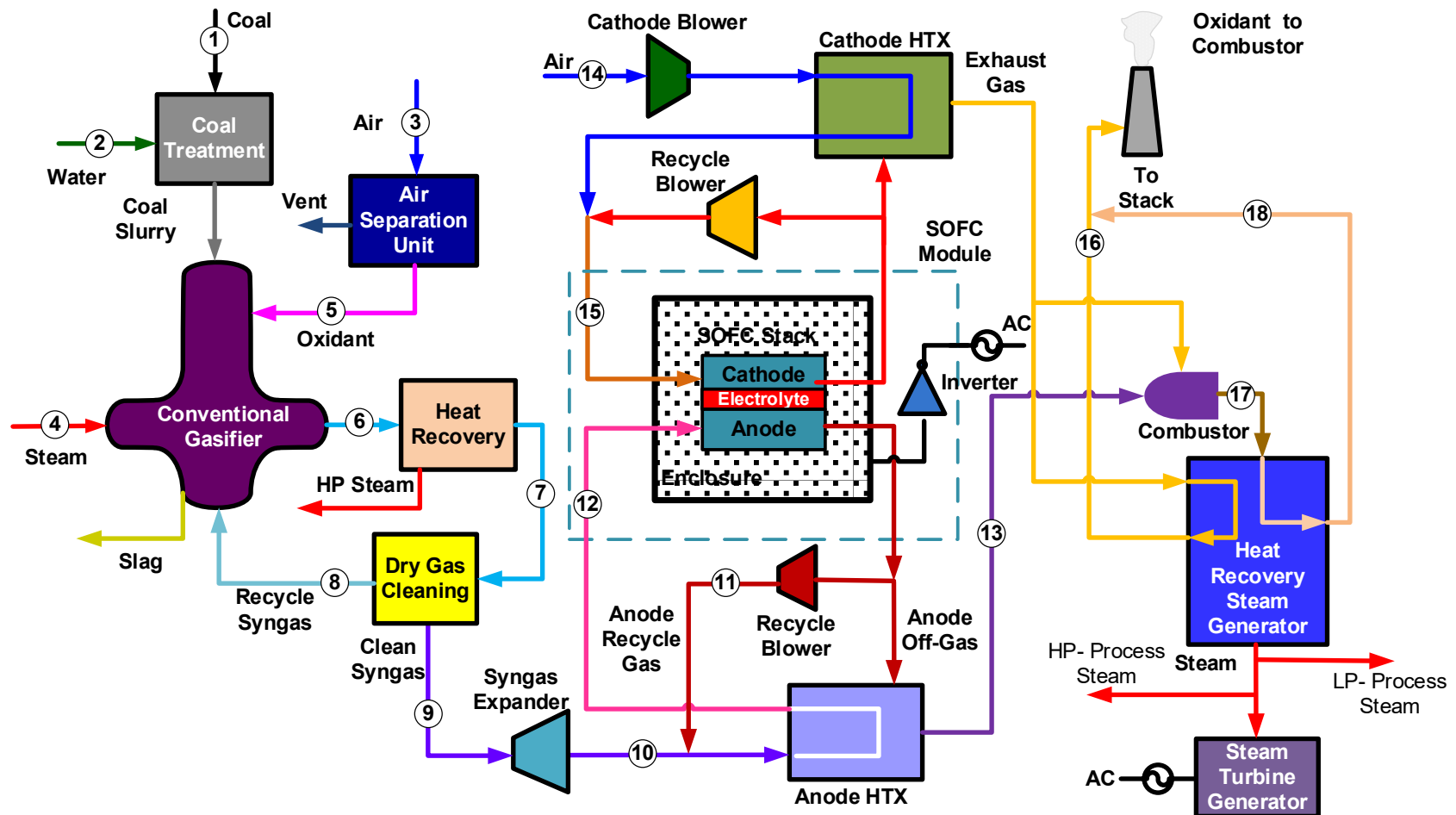


Exhibit 4-5. Case 0 – Atmospheric Reference IGFC Plant with and without CCS – Performance Summary

Carbon Capture	With CCS	No CCS
Power Summary (Gross Power at Generator Terminals, kWe)		
SOFC Power	566,812	513,167
Natural Gas/Syngas Expander Power	39,100	35,400
Steam Turbine Power	198,100	179,900
Total Gross Power, kWe	804,012	728,467
Auxiliary Load Summary, kWe		
Coal Handling	440	430
Coal Milling	2,060	1,870
Sour Water Recycle Slurry Pump	0	0
Ash Handling	860	780
Air Separation Unit Auxiliaries	0	0
Air Separation Unit Main Air Compressor	44,393	28,694
Oxygen Compressor	13,870	12,560
Nitrogen Compressors	0	0
Claus Plant TG Recycle Compressor	1,580	1,430
CO ₂ Compressor	53,540	0
CO ₂ Refrigeration	0	0
Boiler Feedwater Pumps	2,315	2,102
Condensate Pump	200	182
Syngas Recycle Compressor	1,120	1,030
Circulating Water Pump	3,300	2,360
Ground Water Pumps	522	473
Cooling Tower Fans	1,710	1,220
Scrubber Pumps	287	260
Quench Water Pump	879	796
Selexol Auxiliary Power	3,153	2,855
Steam Turbine Auxiliaries	182	166
Claus Plant/TGTU Auxiliaries	229	207
Gas Turbine Auxiliaries	911	828
Cathode Air Blower	10,500	9,550
Cathode Recycle Blower	5,080	4,600
Anode Recycle Blower	650	600
Miscellaneous Balance of Plant ^A	3,523	3,192
Transformer Losses	2,860	2,410
Total Auxiliaries, kWe	154,165	78,594
Net Power, kWe	649,847	649,873
Net Plant Efficiency, % (HHV)	42.96%	47.46%
CO ₂ Capture Rate, %	98.8%	0.0%
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	8,379 (7,942)	7,586 (7,190)
Coal Feed Rate, kg/hr (lb/hr)	200,669 (442,400)	181,675 (400,525)
Thermal Input ^B , kWt	1,512,551	1,369,382
Condenser Cooling Duty, 10 ⁶ kJ/hr (10 ⁶ Btu/hr)	1,118 (1,060)	1,013 (960)
Raw Water Withdrawal, m ³ /min (gpm)	8.7 (2,295)	5.7 (1,498)
Raw Water Consumption, m ³ /min (gpm)	5.8 (1,532)	3.6 (955)

^A Includes plant control systems, lighting, HVAC, and miscellaneous low-voltage loads.

^B HHV of as-received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb).

Exhibit 4-6. Case 0 – Atmospheric Reference IGFC Plant with and without CCS – Gross Power Distribution

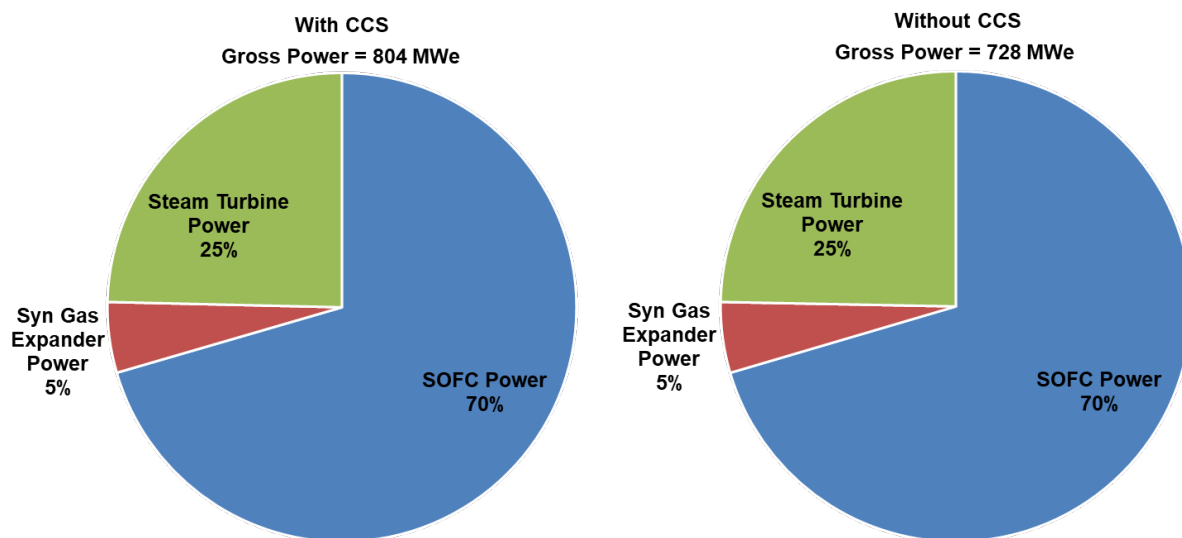


Exhibit 4-7. Case 0 – Atmospheric Reference IGFC Plant with and without CCS – Auxiliary Load Distribution

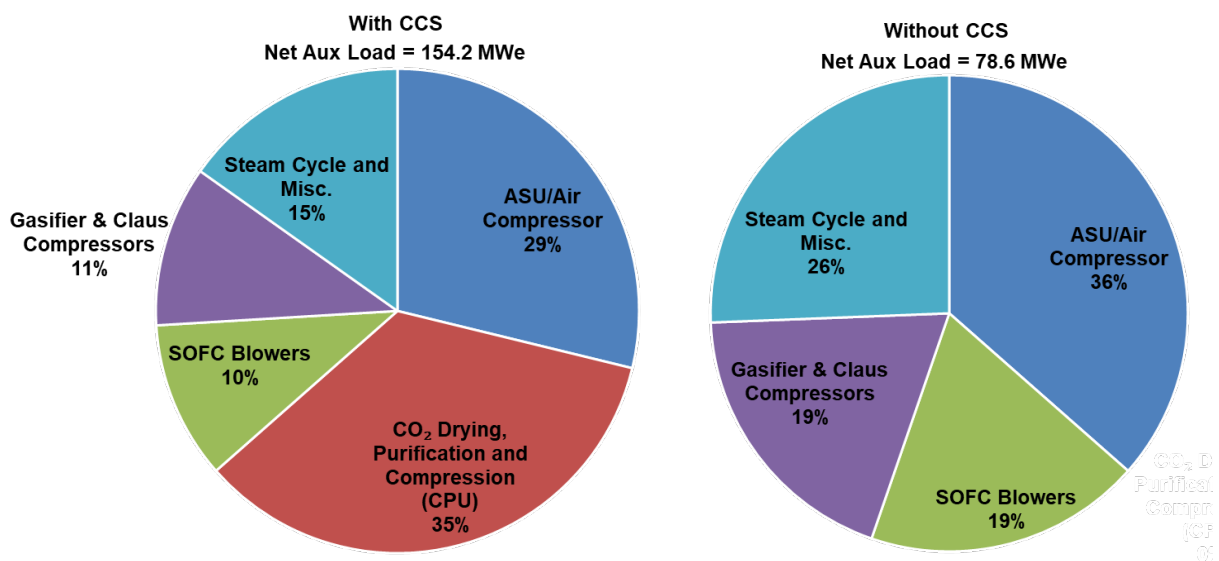


Exhibit 4-8. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Material and Energy Balances

Carbon Balance

Carbon In		Carbon Out	
kg/hr (lb/hr)		kg/hr (lb/hr)	
Coal	127,916 (282,007)	Slag	1,023 (2,256)
Air (CO ₂)	577 (1,273)	Stack Gas	466 (1,028)
		CO ₂ Product	125,309 (276,259)
		N ₂ Product	111 (245)
		Exhaust	1,579 (3,480)
		Convergence Tolerance	3 (8)
Total	128,493 (283,279)	Total	128,493 (283,279)

Sulfur Balance

Sulfur In		Sulfur Out	
kg/hr (lb/hr)		kg/hr (lb/hr)	
Coal	5,030 (11,088)	Elemental Sulfur	5,027 (11,082)
		Polishing Sorbent	3 (6)
		Convergence Tolerance	0 (0)
Total	5,030 (11,088)	Total	5,030 (11,088)

Water Balance

Water Use	Water Demand m ³ /min (gpm)	Internal Recycle m ³ /min (gpm)	Raw Water Withdrawal m ³ /min (gpm)	Process Water Discharge m ³ /min (gpm)	Raw Water Consumption m ³ /min (gpm)
Slag Handling	0.36 (95)	0.36 (95)			
Slurry Water	1.38 (363)	1.38 (363)			
Quench/Wash	0.3 (88)	0.33 (88)			
Condenser Makeup	1.1 (295)	0.0 (0)	1.1 (295)	0.0 (0)	1.1 (295)
Gasifier Steam	1.1 (292)	0.0 (0)	1.1 (292)		
BFW Makeup	0.0 (3)	0.0 (0)	0.0 (3)		
Cooling Tower	12.8 (3,392)	5.27 (1,392)	7.6 (2,000)	2.9 (763)	4.7 (1,237)
CO ₂ Dehydration	0.0 (0)	2.33 (616)	-2.33 (-616)		
SW S Excess Water	0.0 (0)	2.94 (776)	-2.94 (-776)		
Total	16.0 (4,233)	7.34 (1,938)	8.7 (2,295)	2.9 (763)	5.8 (1,532)

Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	5,445 (5,161)	4.6 (4.3)		5,450 (5,165)
ASU Air		27.6 (26.1)		28 (26)
Fuel cell Air		116.0 (109.9)		116 (110)
Raw Water Makeup		32.7 (31.0)		33 (31)
Auxiliary Power			555 (526)	555 (526)
TOTAL	5,445 (5,161)	180.8 (171.3)	555 (526)	6,181 (5,858)
Heat Out GJ/hr (MMBtu/hr)				
Slag	27 (26)	17.1 (16.2)		44 (42)
Sulfur	47 (44)	0.5 (0.5)		47 (45)
CO ₂		-95.2 (-90.2)		-95 (-90)
Cooling Tower Blowdown		21.5 (20.3)		21 (20)
Exhaust		662.5 (627.9)		662 (628)
HRSF Flue Gas		0.0 (0.0)		0 (0)
Condenser		1,120 (1,062)		1,120 (1,062)
Non-Condenser Cooling Tower Loads*		557 (528)		557 (528)
Process Losses*		986 (934)		986 (934)
Power			2,894 (2,743)	2,894 (2,743)
TOTAL	74 (70)	3,268 (3,045)	2,894 (2,743)	6,237 (5,858)

Emissions

	kg/GJ (lb/10 ⁶ Btu)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO ₂	0 (0)	0 (0)	0 (0)
NO _x	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO ₂	1 (3)	52,556 (57,933)	9 (21)

Exhibit 4-9. Case ACONVIGFC0B – Reference Plant with CCS – Heat and Material Balance Diagram – ASU, Gasification, and Clean-up

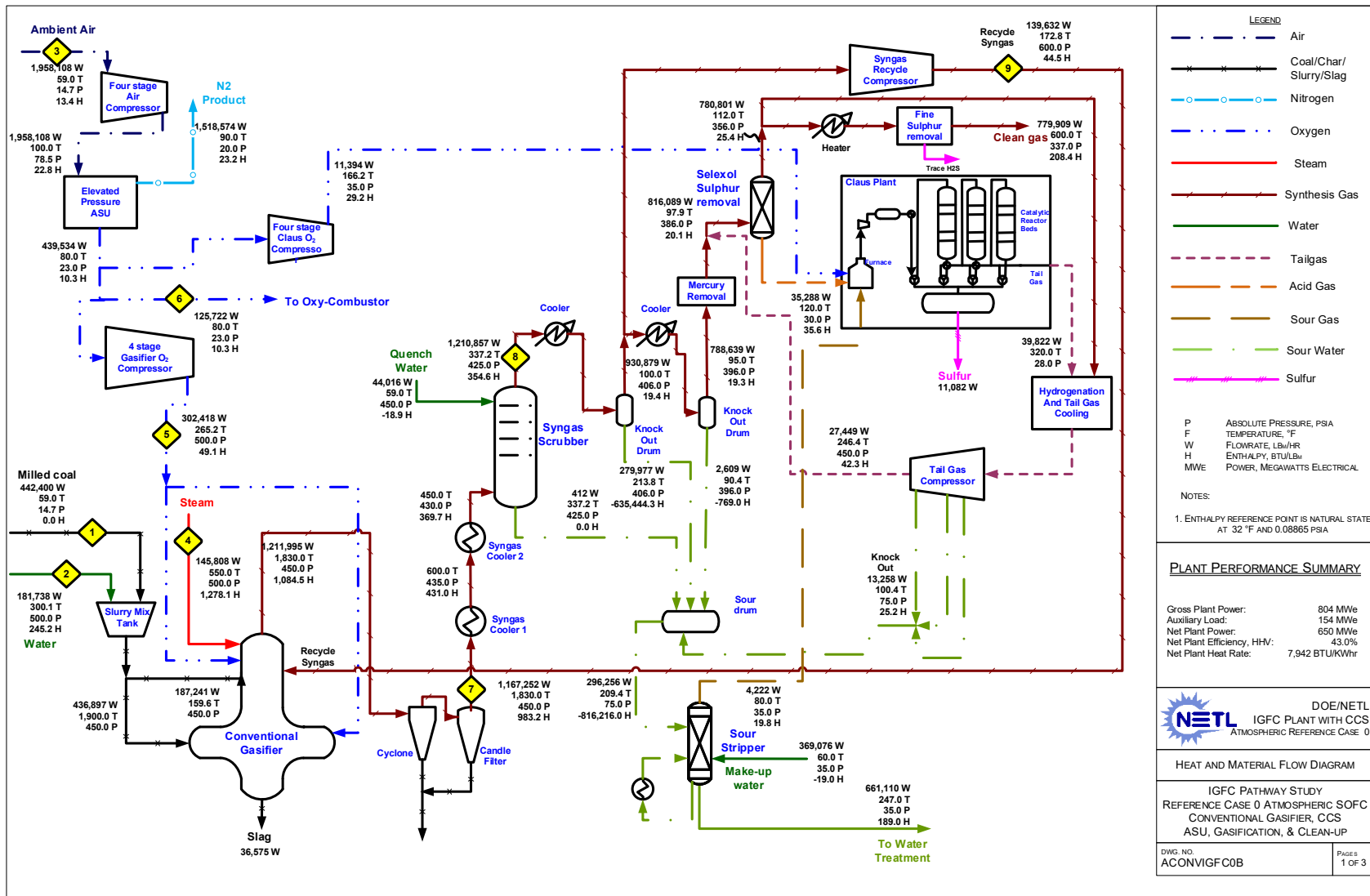


Exhibit 4-10. Case ACONVIGFC0B – Reference Plant with CCS – Heat and Material Balance Diagram – Power Block

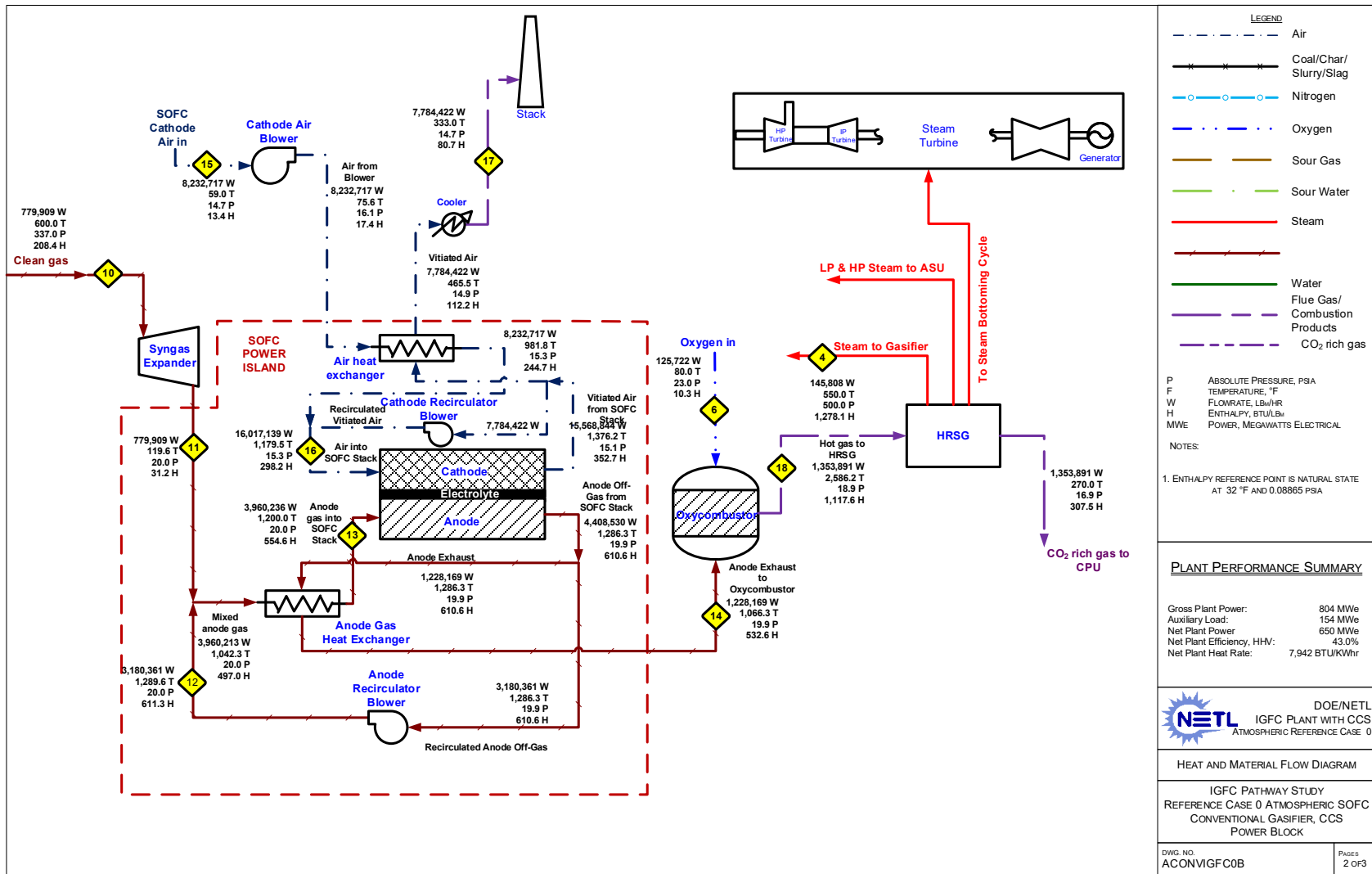
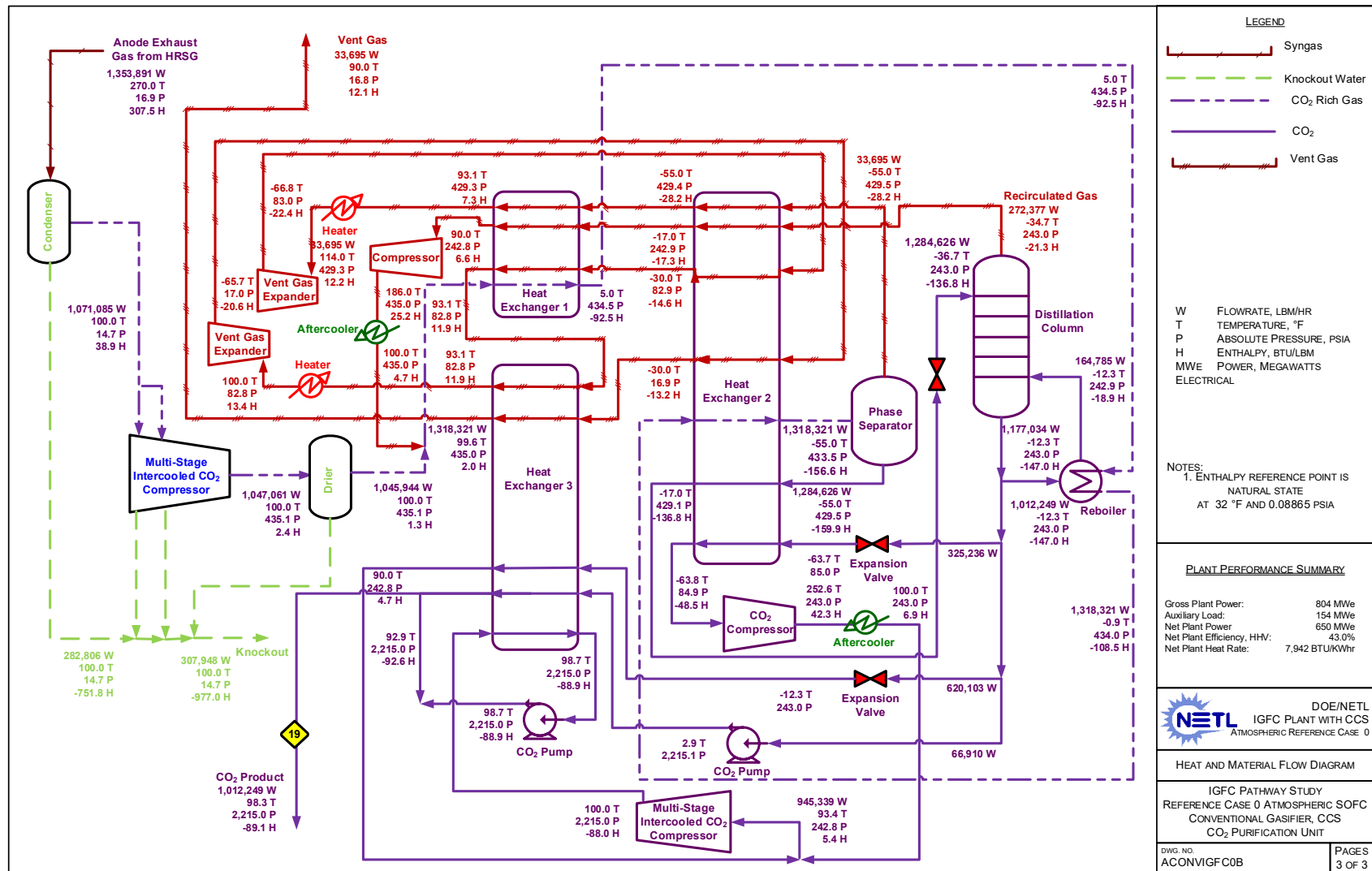


Exhibit 4-11. Case ACONVIGFC0B – Reference Plant with CCS – Heat and Material Balance Diagram – CO₂ Dehydration, Purification, and Compression


TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-12. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Capital Costs

Cost Component	Cost (\$1000)	Specific Cost (\$/kWe AC)
	2018\$	2018\$
SOFC Module		
SOFC Stack	133,735	206
Container	5,277	8
Insulation	2,378	4
Module Assembly	5,944	9
Air Distribution	5,944	9
Fuel Distribution	5,944	9
Module Current Collectors	2,972	5
Module I&C	2,972	5
Inverter	40,418	62
Total SOFC Module with 10 % Extra Installed Area	205,582	316
Syngas Expander	11,676	18
SOFC BOP		
Cathode Air Blower	3,990	6
Cathode Gas Recycle Blower	9,356	14
Cathode Heat Exchanger	27,734	43
Anode Recycle Blower	1,543	2
Anode Heat Exchanger	27,246	42
Oxy-Combustor	28,740	44
Total SOFC BOP	98,609	152
TOTAL SOFC POWER ISLAND	315,868	486
COAL & SORBENT HANDLING	65,931	101
COAL & SORBENT PREP & FEED	38,659	59
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	1,082,016	1665
ASU & oxidant compressor	95,402	147
Other gasification equip & foundations	93,820	144
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	1,271,239	1956
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	0	0
Single-Stage Selexol/MDEA	14,141	22
Claus Plant	156,967	242
Trace removal	5,386	8
COS Hydrolysis	38,219	59
Blowback, Piping, Foundations	9,278	14
Sulfur polishing/NG desulfurization	12,070	19
TOTAL GAS CLEANUP & PIPING	236,062	363
STEAM CYCLE		
HRSF, Ducting, and Stack	69,891	108
Steam Power System	66,838	103
Feedwater and Miscellaneous BOP systems	86,030	132
TOTAL STEAM CYCLE	222,758	343
CO₂ COMPRESSION & PURIFICATION	150,387	231
COOLING WATER SYSTEM	46,347	71
ASH/SPENT SORBENT HANDLING SYS	10,665	16
ACCESSORY ELECTRIC PLANT	67,109	103
INSTRUMENTATION & CONTROL	27,773	43
IMPROVEMENTS TO SITE	30,787	47
BUILDING & STRUCTURES	17,512	27
TOTAL PLANT COST (TPC)	2,500,963	4038
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	8,624	
1 Month Maintenance Materials	1,652	
1 Month Non-fuel Consumables	1,121	
1 Month Waste Disposal	254	
25% of 1 Months Fuel Cost at 100% CF	2,098	
2% of TPC	50,019	
Total Preproduction Costs	63,768	
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	18,819	
0.5% of TPC (spare parts)	12,505	
Total Inventory Capital	31,324	
Initial Cost for Catalyst and Chemicals	55,372	
Land	900	
Other Owner's Costs	375,144	
Financing Costs	67,526	
TOTAL OWNER'S COSTS	594,034	
TOTAL OVERNIGHT COST (TOC)	3,094,997	4763
TASC Multiplier	1.154	
TOTAL AS-SPENT COST (TASC)	3,571,627	5496

Exhibit 4-13. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – O&M Costs

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
OPERATING & MAINTENANCE LABOR		
Operating Labor Rate (base)	39.70	
Stack replacement O&M, \$/hr per stack kW	18.33	
Annual Operating Labor Cost	5,261,256	
Maintenance Labor Cost	8,537,924	
Administrative & Support Labor	3,449,795	
Property Taxes and Insurance	50,019,258	
TOTAL FIXED OPERATING COSTS	67,268,232	14.77
VARIABLE OPERATING COSTS		
Maintenance Material Cost	15,856,144	
Stack replacement		
SOFC stack replacement cost, \$/kW AC	228	
SOFC stack replacement O&M, \$/y per kW	23	
Stack Replacement Cost	14,727,566	3.23
CONSUMABLES		
Water		
Price (\$/1000 gal)	1.900	
Annual (\$)	978,259	
Chemicals		
MU & WT Chem.		
Price (\$/lb)	0.28	
Annual (\$)	790,382	
Carbon (Trace Removal)		
Price (\$/lb)	1.6	
Initial Cost (\$)	865,365	
Annual (\$)	346,146	
COS Catalyst		
Price (\$/m ³)	45,909	
Initial Cost (\$)	32,045,329	
Annual (\$)	6,409,066	
Selexol Solution		
Price (\$/gal)	38.00	
Initial Cost (\$)	19,205,805	
Annual (\$)	556,412	
Claus / DSRP Catalyst		
Price (\$/ft ³)	48	
Annual (\$)	25,698	
ZnO polishing sorbent		
Price (\$/lb)	1.5	
Initial Cost (\$)	793,501	
Annual (\$)	889,518	
COS Catalyst		
Price (\$/ft ³)	772	
Initial Cost (\$)	2,461,746	
Annual (\$)	769,295	
Sub Total Chemicals		
Initial Fill (\$)	55,371,746	
Annual (\$)	9,786,518	
Waste Disposal		

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
Spent Trace Catalyst		
Price (\$/lb)	0.42	
Annual (\$)	96,326	
Ash + HCl Sorbent		
Price (\$/ton)	16.31	
Annual (\$)	2,090,335	
Spent sorbents		
Price (\$/lb)	0.42	
Annual (\$)	249,065	
Subtotal-Waste Disposal (\$)	2,435,727	
TOTAL VARIABLE OPERATING COSTS		
Initial Fill (\$)	55,371,746	
Annual (\$)	43,784,214	9.61
Fuel Coal (ton)		
Price (\$/ton)	51.96	
Annual (\$)	80,546,812	17.69

Exhibit 4-14. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – Capital Cost Components

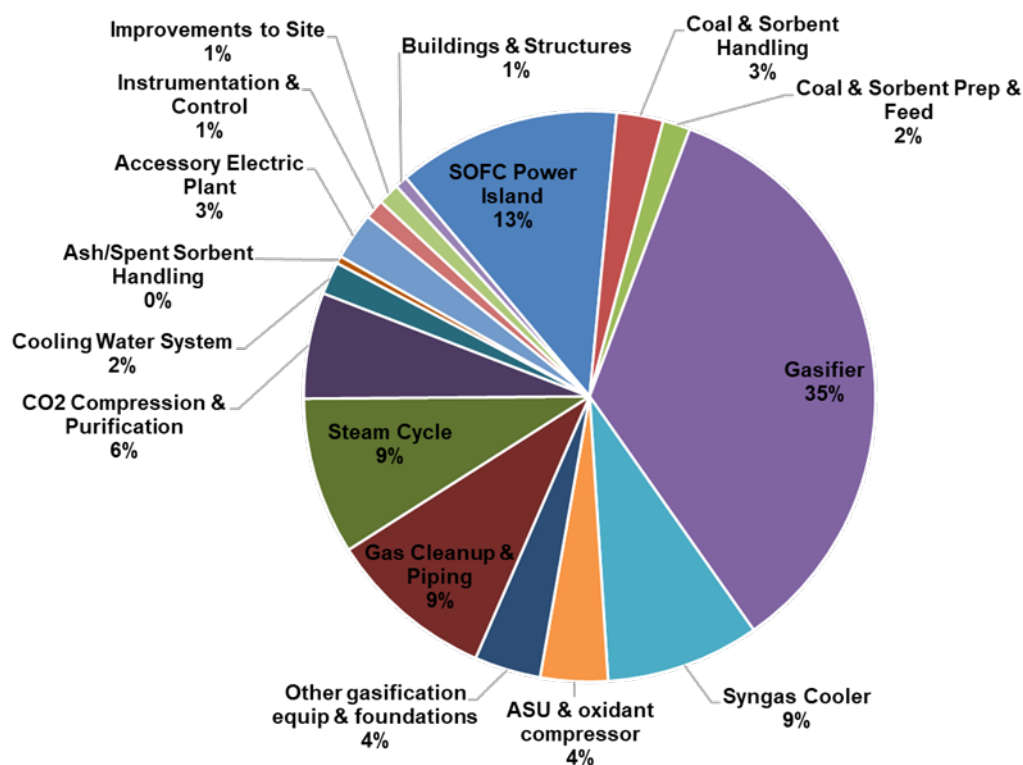


Exhibit 4-15. Case ACONVIGFC0B – Atmospheric Reference IGFC Plant with CCS – BOP Cost Distribution

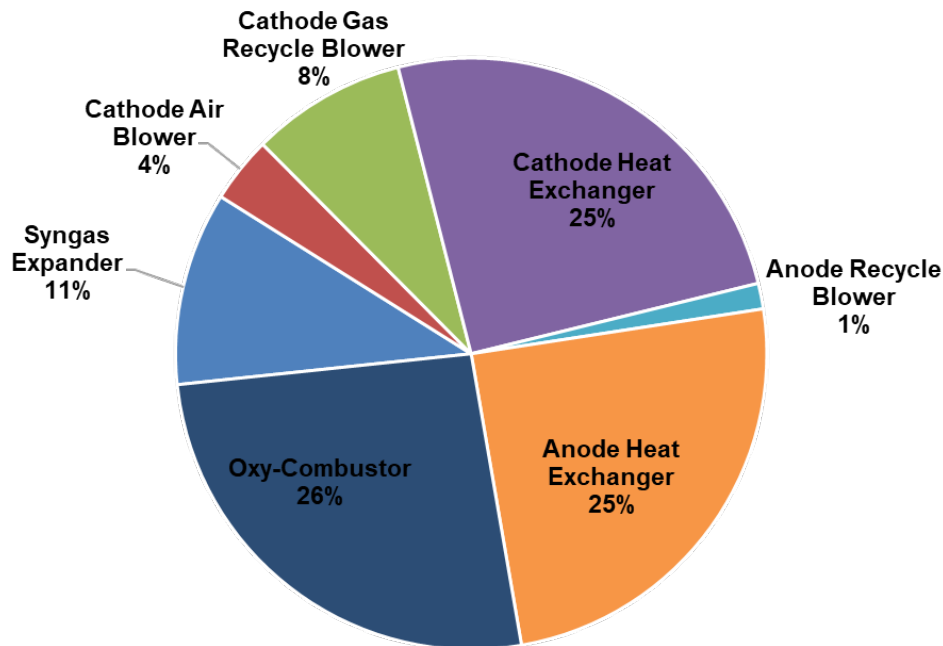


Exhibit 4-16. Case 0 – Atmospheric reference IGFC plant with and without CCS – levelized cost of electricity

Levelized Cost of Electricity (2018\$/MWh)		
Carbon Capture	With CCS	No CCS
Component	\$/MWh	
Variable Costs	27.3	25.5
Fuel Costs	17.7	16.0
Variable O&M Costs	9.6	9.5
Fixed O&M Costs	14.8	13.2
Capital Costs	55.4	49.3
Total LCOE (excluding T&S)	97.5	88.1
T&S	7.1	0.0
Total LCOE (including T&S)	104.6	88.1

4.2 CASE 3: ENHANCED GASIFIER ATMOSPHERIC IGFC

Case 3 uses an enhanced gasifier, the conventional E-Gas™ gasifier, that results in a CH₄ content of 10.8 percent in the syngas entering the SOFC section. The layout is nearly identical to the

reference plant, but the case assumes enhanced SOFC technology parameters consistent with the pathway as listed in Exhibit 4-17.

Exhibit 4-17. IGFC Parameters Used for the Enhanced Gasifier Case 3

Parameter	Value
Gasifier Technology	Enhanced
Syngas CH ₄ content (mole %)	10.8
SOFC Operating Pressure	Atmospheric
Cell Technology	Advanced Cell
Fuel Utilization, %	85
Current Density, mA/cm ²	400
Degradation, %/1000 h	0.2
Inverter Efficiency (%)	97
Stack Cost (\$/kW)	225
Capacity Factor (%)	85

The relevant process data for the numbered streams in the BFD for the Case 3 plant without CCS shown in Exhibit 4-18 are tabulated in Exhibit 4-19. The BFD for the Case 3 plant with CCS is identical to the reference plant BFD shown in Exhibit 4-2. The higher syngas content reduces the SOFC process air flow requirement by nearly 50 percent.

Case 3 Plant Performance

As shown in Exhibit 4-20, the Case 3 plants results in efficiencies that are ≈3 percent higher than the reference plant as a combined result of higher syngas CH₄ content and higher fuel utilization relative to the reference plant. The performance summary for the Case 3 IGFC plant shows that the plant without CCS results in a net plant HHV efficiency of ≈49.7 percent, which is nearly 4.2 percentage points higher than the corresponding plant with CCS. The CCS plant efficiency is nearly 10 percent higher than the values achievable with conventional coal power plant technologies and is comparable to an NGCC system with CCS.

For both systems with and without CCS, the SOFC power island, due to the higher fuel utilization, generates a larger share (76 percent) of the gross power relative to the reference plant, as seen in Exhibit 4-21. As in the reference plant, the addition of the CPU and the increased ASU oxygen demand, which together contribute to nearly 62 percent of the auxiliary load as shown in Exhibit 4-22, nearly double the auxiliary load demand for the plant with CCS plant relative to the plant without CCS.

The salient material and energy balances for the plant without CCS are listed in Exhibit 4-23, which shows that the emissions of the IGFC plant even without CCS is much lower than those of other conventional coal-based technologies. The IGFC plant water consumption is also significantly lower than the corresponding value for conventional fossil fuel power plant

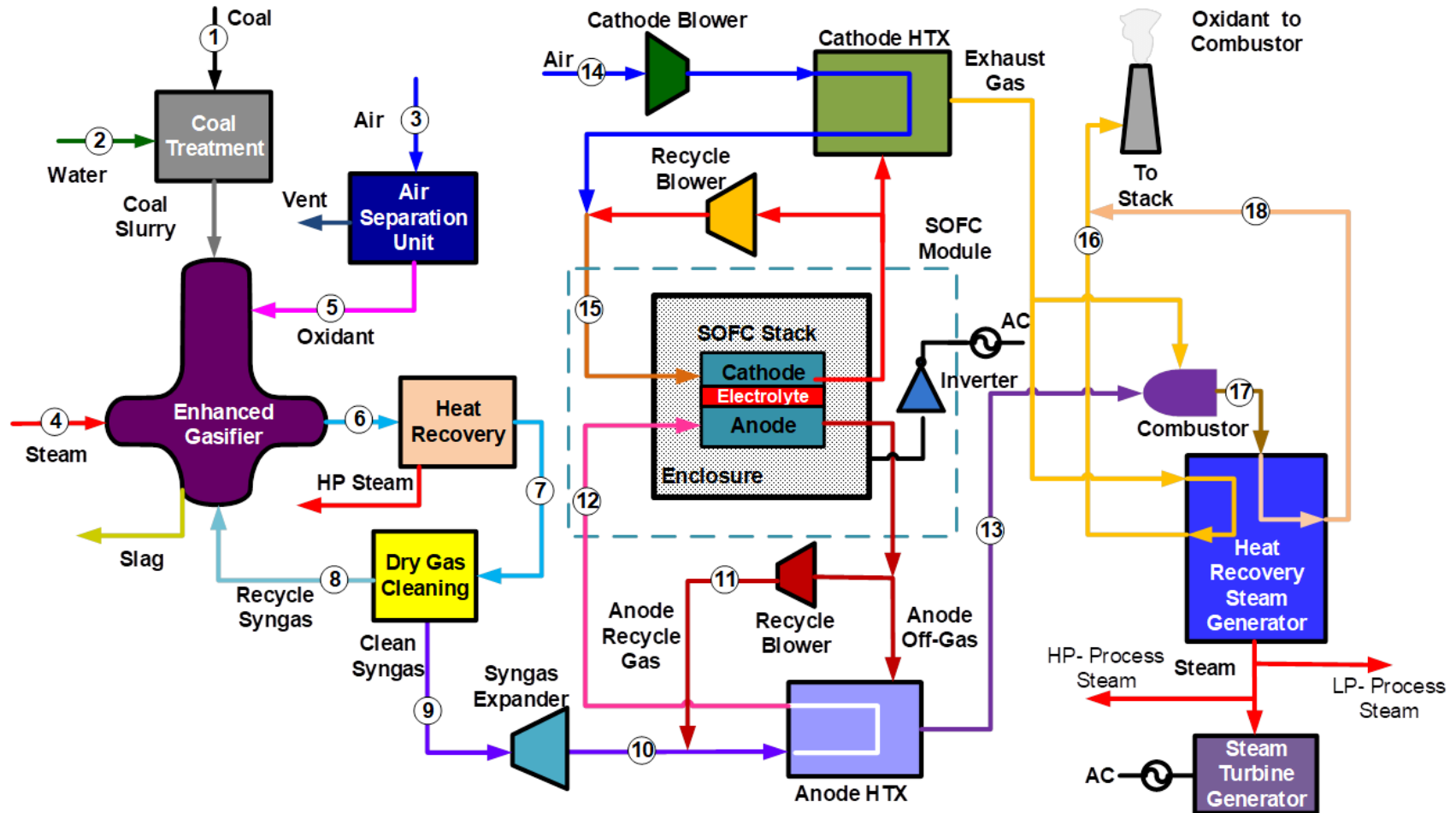
technologies. The heat and mass balance diagrams for the gasifier, ASU, syngas clean-up, and the power island are shown in Exhibit 4-24 and Exhibit 4-25, respectively.

Case 3 Plant Costs

The capital and O&M costs for the Case 3 IGFC plant without CCS are itemized in Exhibit 4-26 and Exhibit 4-27, respectively. The gasifier and cleanup costs account for nearly 60 percent of the capital costs, as shown in Exhibit 4-28 and accordingly, any cost reduction in the gasifier systems would greatly benefit the overall cost of the IGFC system. The cathode and anode heat exchangers and the oxy-combustor contribute to 75 percent of the BOP costs, as shown in Exhibit 4-29.

As in the reference plant, the LCOE for the Case 3 plant is dominated by the capital cost component (≈ 56 percent), as shown in Exhibit 4-30. The LCOE for the plant with CCS is \$89.6/MWh, 10 percent higher than the LCOE for the plant without CCS, indicating a much lower CCS cost penalty relative to conventional systems, which could be as high as 60 percent.

Exhibit 4-18. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Process Diagram



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-19. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
V-L Mole Fraction																		
Ar	0.0000	0.0000	0.0094	0.0000	0.0031	0.0006	0.0006	0.0008	0.0008	0.0008	0.0007	0.0007	0.0007	0.0094	0.0097	0.0099	0.0038	0.0052
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0841	0.0841	0.1096	0.1107	0.1107	0.0000	0.0309	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.2609	0.2609	0.3400	0.3437	0.3437	0.0576	0.1374	0.0576	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0000	0.0000	0.0003	0.0000	0.0000	0.1736	0.1736	0.2266	0.2292	0.2292	0.5020	0.4259	0.5020	0.0003	0.0003	0.0003	0.3991	0.5494
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0004	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	0.2341	0.2341	0.3050	0.3081	0.3081	0.0767	0.1413	0.0767	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0000	1.0000	0.0104	1.0000	0.0000	0.2315	0.2315	0.0023	0.0012	0.0012	0.3578	0.2583	0.3578	0.0104	0.0107	0.0110	0.3135	0.0550
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0011	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	0.0000	0.0073	0.0073	0.0100	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N ₂	0.0000	0.0000	0.7722	0.0000	0.0019	0.0022	0.0022	0.0028	0.0063	0.0063	0.0052	0.0055	0.0052	0.7722	0.7929	0.8147	0.2766	0.3807
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0046	0.0046	0.0018	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	0.2077	0.0000	0.9950	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.1865	0.1641	0.0070	0.0097
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	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	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	0	3,957	16,789	3,244	3,257	20,713	20,713	2,384	13,365	13,365	34,530	47,895	16,324	119,253	232,287	105,364	22,898	16,635
V-L Flowrate (kg/hr)	0	71,292	484,427	58,434	104,286	443,319	443,319	53,476	298,581	298,581	1,052,480	1,351,056	497,564	3,440,947	6,682,905	3,021,959	717,563	604,723
Solids Flowrate (kg/hr)	173,544	0	0	0	0	162	162	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	148	15	288	130	935	232	72	316	55	698	649	559	15	637	167	1,046	38
Pressure (MPa, abs)	0.10	0.69	0.10	3.45	3.45	4.83	4.69	4.14	2.32	0.14	0.14	0.14	0.1	0.101	0.105	0.101	0.130	0.101
Enthalpy (kJ/kg) ^A	---	566.85	31.06	2,973.90	114.27	2,154.16	831.67	89.87	478.63	79.27	1,464.35	1,316.74	1,259.4	31.057	693.615	187.729	1,821.954	102.401
Density (kg/m ³)	---	864.7	1.2	14.8	33.1	10.2	24.2	32.6	10.5	1.1	0.5	0.5	0.6	1.2	0.4	0.8	0.4	1.4
V-L Molecular Weight	---	18.015	28.854	18.015	32.016	21.403	21.403	22.431	22.340	22.340	30.480	28.208	30	28.854	28.770	28.681	31.337	36.353
V-L Flowrate (lb _{mol} /hr)	0	8,724	37,013	7,151	7,181	45,665	45,665	5,256	29,465	29,465	76,127	105,591	35,989	262,907	512,105	232,287	50,482	36,674
V-L Flowrate (lb/hr)	0	157,172	1,067,978	128,824	229,912	977,350	977,350	117,894	658,258	658,258	2,320,322	2,978,568	1,096,941	7,585,989	14,733,284	6,662,279	1,581,957	1,333,186
Solids Flowrate (lb/hr)	382,600	0	0	0	0	356	356	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	299	59	550	265	1,715	450	162	600	130	1,289	1,200	1,039	59	1,179	333	1,915	100
Pressure (psia)	14.7	100.0	14.7	500.0	500.0	700.0	680.0	600.0	337.0	20.0	20.0	20.0	19.9	14.7	15.3	14.7	18.9	14.7
Enthalpy (Btu/lb) ^A	---	243.7	13.4	1,278.5	49.1	926.1	357.6	38.6	205.8	34.1	629.6	566.1	541.4	13.4	298.2	80.7	783.3	44.0
Density (lb/ft ³)	---	53.983	0.076	0.923	2.067	0.636	1.508	2.038	0.658	0.071	0.032	0.032	0	0.076	0.025	0.050	0.023	0.089

^A Reference conditions are 32.02 F and 0.089 psia

Exhibit 4-20. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Performance Summary

Carbon Capture	With CCS	No CCS
Power Summary (Gross Power at Generator Terminals, kW_e)		
SOFC Power	595,945	545,923
Natural Gas/Syngas Expander Power	34,900	32,000
Steam Turbine Power	158,900	141,800
Total Gross Power, kW_e	789,745	719,723
Auxiliary Load Summary, kW_e		
Coal Handling	430	420
Coal Milling	1,950	1,780
Sour Water Recycle Slurry Pump	0	0
Ash Handling	810	740
Air Separation Unit Auxiliaries	0	0
Air Separation Unit Main Air Compressor	36,342	24,213
Oxygen Compressor	12,950	10,550
Nitrogen Compressors	0	0
Claus Plant TG Recycle Compressor	1,490	1,370
CO ₂ Compressor	50,570	0
CO ₂ Refrigeration	0	0
Boiler Feedwater Pumps	1,857	1,657
Condensate Pump	161	143
Syngas Recycle Compressor	870	800
Circulating Water Pump	2,770	1,890
Ground Water Pumps	493	452
Cooling Tower Fans	1,430	990
Scrubber Pumps	262	240
Quench Water Pump	831	761
Selexol Auxiliary Power	2,908	2,662
Steam Turbine Auxiliaries	146	130
Claus Plant/TGTU Auxiliaries	216	198
Gas Turbine Auxiliaries	731	652
Cathode Air Blower	10,560	9,720
Cathode Recycle Blower	5,080	4,660
Anode Recycle Blower	520	470
Miscellaneous Balance of Plant ^A	3,460	3,153
Transformer Losses	2,770	2,350
Total Auxiliaries, kW_e	139,606	70,002
Net Power, kW_e	650,138	649,721
Net Plant Efficiency, % (HHV)	45.51%	49.67%
CO₂ Capture Rate, %	98.7%	0.0%
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	7,910 (7,497)	7,248 (6,870)
Coal Feed Rate, kg/hr (lb/hr)	189,511 (417,800)	173,544 (382,600)
Thermal Input^B, kWt	1,428,444	1,308,097
Condenser Cooling Duty, 10⁶ kJ/hr (10⁶ Btu/hr)	897 (850)	802 (760)
Raw Water Withdrawal, m³/min (gpm)	6.7 (1,773)	3.9 (1,024)
Raw Water Consumption, m³/min (gpm)	4.3 (1,134)	2.2 (588)

^A Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads.

^B HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb).

Exhibit 4-21. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Gross Power Distribution

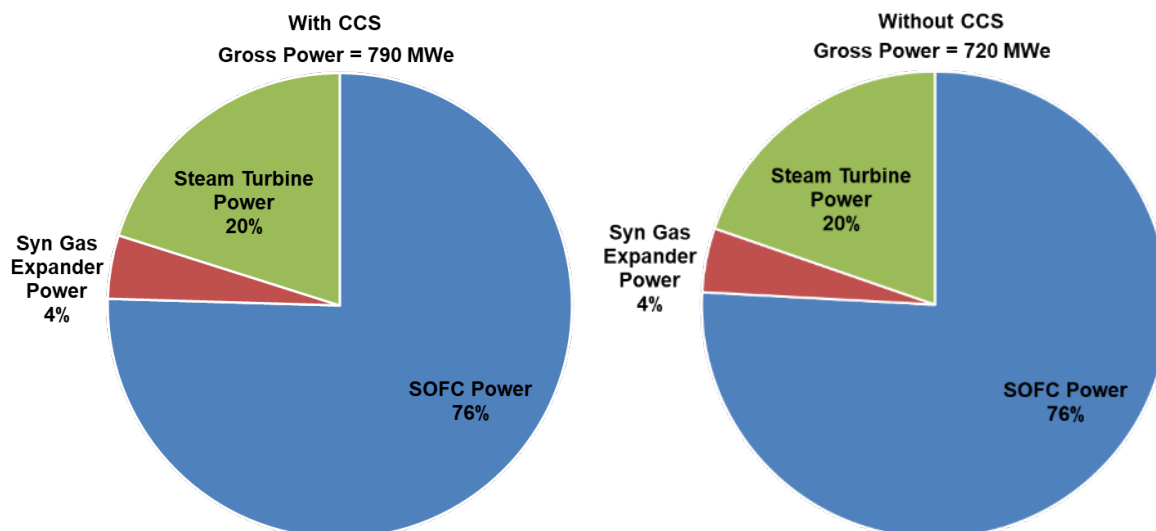


Exhibit 4-22. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Auxiliary Load Distribution

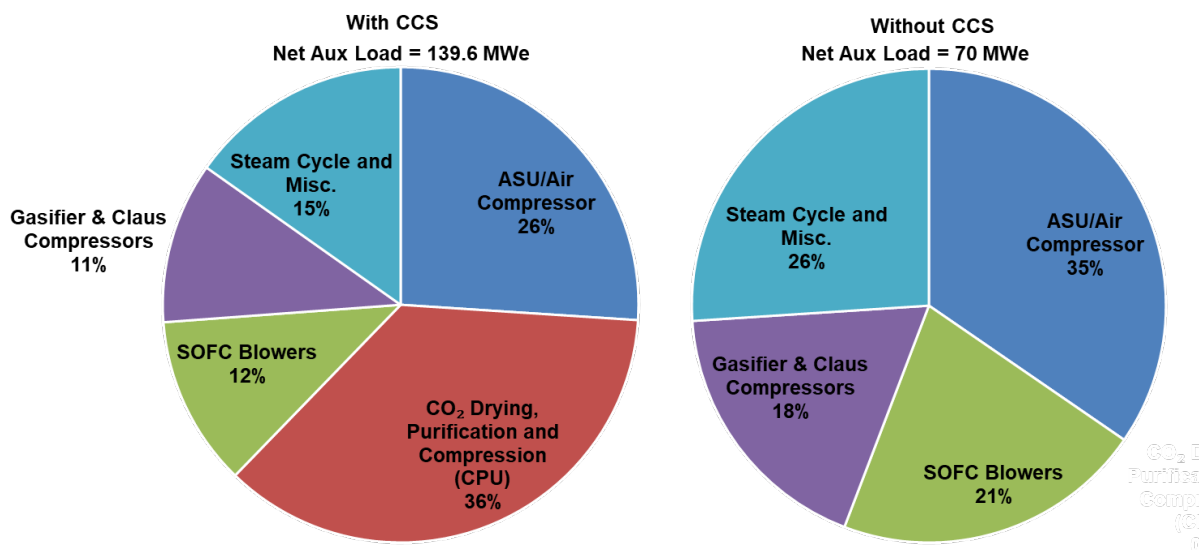


Exhibit 4-23. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Material and Energy Balances
Carbon Balance

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	110,625 (243,887)	Slag	885 (1,951)
Air (CO ₂)	490 (1,081)	Stack Gas	401 (883)
		CO ₂ Product	109,768 (241,997)
		N ₂ Product	60 (133)
			0 (0)
		Convergence Tolerance	2 (4)
Total	111,116 (244,968)	Total	111,116 (244,968)

Sulfur Balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	4,350 (9,590)	Elemental Sulfur	4,347 (9,584)
		Polishing Sorbent	2 (5)
		Convergence Tolerance	0 (0)
Total	4,350 (9,590)	Total	4,350 (9,590)

Water Balance

Water Use	Water Demand m ³ /min (gpm)	Internal Recycle m ³ /min (gpm)	Raw Water Withdrawal m ³ /min (gpm)	Process Water Discharge m ³ /min (gpm)	Raw Water Consumption m ³ /min (gpm)
Slag Handling	0.31 (82)	0.31 (82)			
Slurry Water	1.19 (314)	1.19 (314)			
Quench/Wash	0.3 (70)	0.26 (70)			
Condenser Makeup	1.0 (260)	0.0 (0)	1.0 (260)	0.0 (0)	1.0 (260)
Gasifier Steam	1.0 (258)	0.0 (0)	1.0 (258)		
BFW Makeup	0.0 (3)	0.0 (0)	0.0 (3)		
Cooling Tower	7.3 (1,939)	4.45 (1175)	2.9 (764)	1.7 (436)	1.2 (328)
CO ₂ Dehydration	0.0 (0)	1.88 (498)	-1.88 (-498)		
SW/S Excess Water	0.0 (0)	2.56 (677)	-2.56 (-677)		
Total	10.1 (2,665)	6.21 (1641)	3.9 (1,024)	1.7 (436)	2.2 (588)

Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	4,709 (4,463)	3.9 (3.7)		4,713 (4,467)
ASU Air		15.0 (14.3)		15 (14)
Fuel cell Air		106.9 (101.3)		107 (101)
Raw Water Makeup		14.6 (13.8)		15 (14)
Auxiliary Power			252 (239)	252 (239)
TOTAL	4,709 (4,463)	140.4 (133.1)	252 (239)	5,102 (4,835)
Heat Out GJ/hr (MMBtu/hr)				
Slag	24 (22)	13.6 (12.9)		37 (35)
Sulfur	40 (38)	0.4 (0.4)		41 (39)
CO ₂		61.9 (58.7)		62 (59)
Cooling Tower Blowdown		12.3 (11.6)		12 (12)
Exhaust		567.3 (537.7)		567 (538)
Condenser		802 (760)		802 (760)
Non-Condenser Cooling Tower Loads*		157 (148)		157 (148)
Process Losses		902 (854)		902 (854)
Difference**		-69 (-65)		-69 (-65)
Power			2,591 (2,456)	2,591 (2,456)
TOTAL	64 (60)	2,447 (2,319)	2,591 (2,456)	5,102 (4,835)

Emissions

	kg/GJ (lb/10 ⁶ Btu)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO ₂	0 (0)	0 (0)	0 (0)
NO _x	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO ₂	86 (199)	005,736 (3,313.25)	561 (1,237)

Exhibit 4-24. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC without CCS – Heat and Material Balance Diagram – ASU, Gasification, and Clean-up

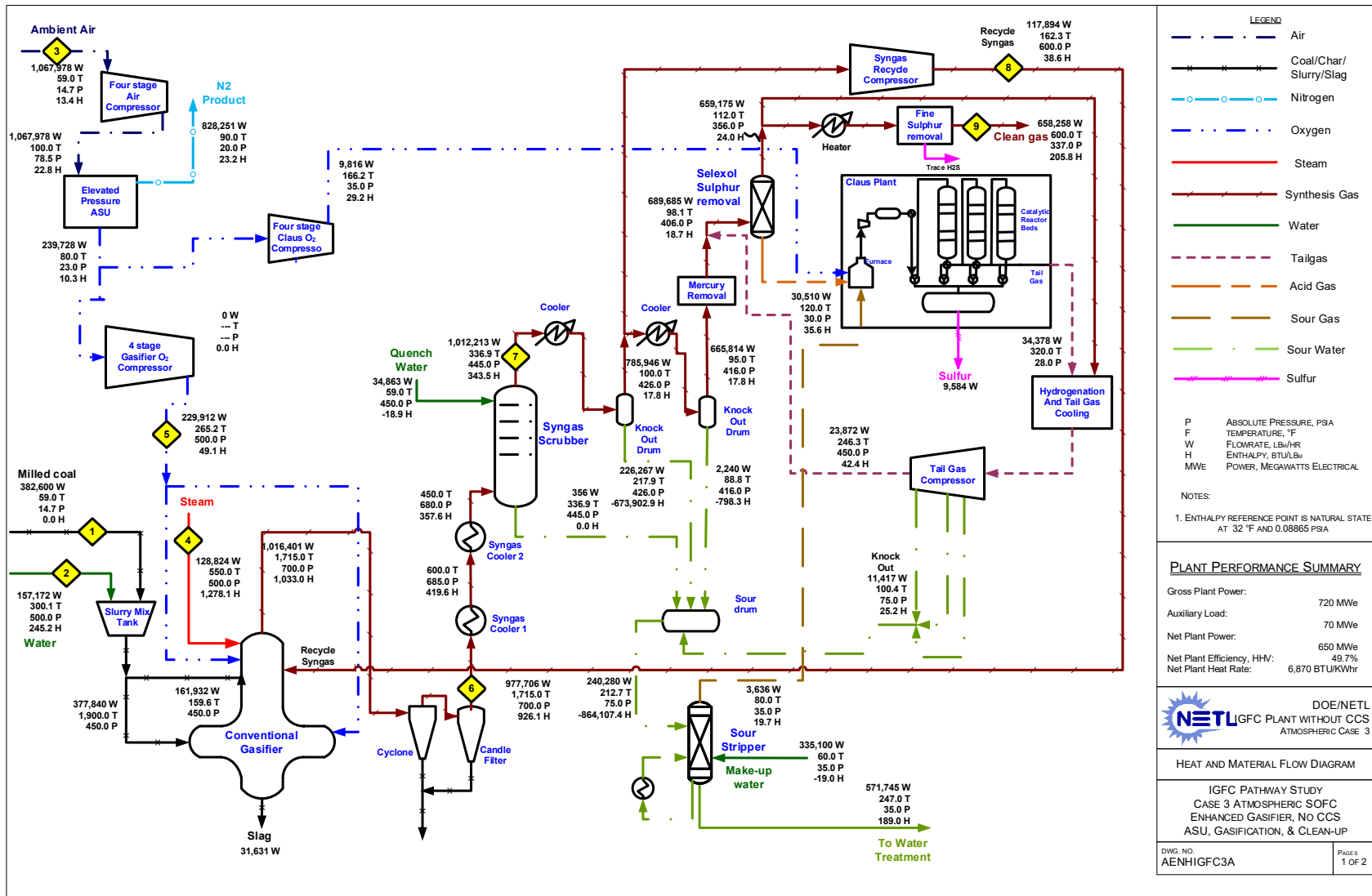
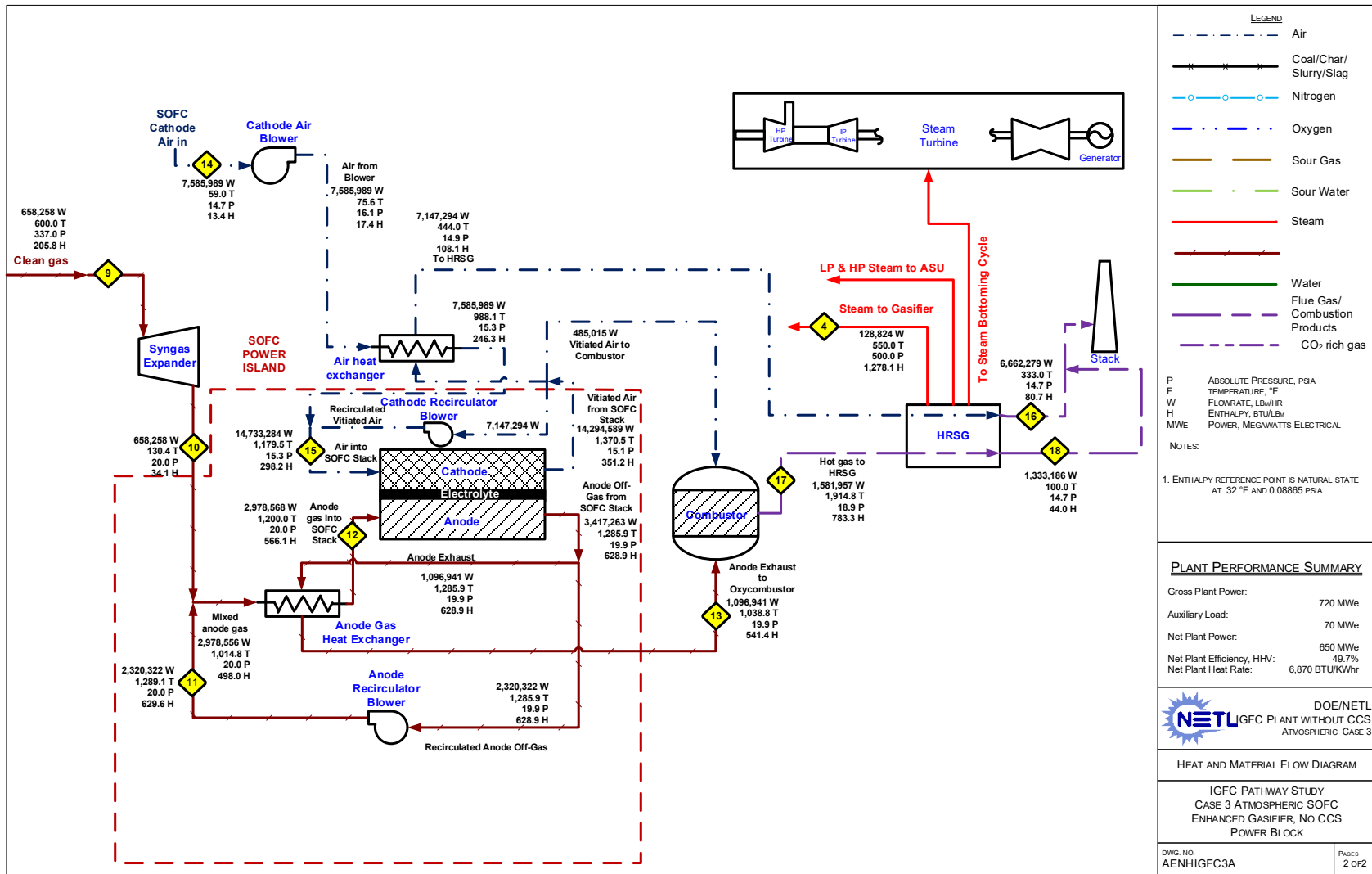


Exhibit 4-25. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC without CCS – Heat and Material Balance Diagram – Power Block



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-26. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC plant without CCS – Capital Costs

Cost Component	Cost (\$1000)	Specific Cost (\$/kWe AC)
	2018\$	2018\$
SOFC Module		
SOFC Stack	129,802	200
Container	5,122	8
Insulation	2,308	4
Module Assembly	5,769	9
Air Distribution	5,769	9
Fuel Distribution	5,769	9
Module Current Collectors	2,884	4
Module I&C	2,884	4
Inverter	39,229	60
Total SOFC Module with 10 % Extra Installed Area	199,536	307
Syngas Expander	10,060	15
SOFC BOP		
Cathode Air Blower	3,707	6
Cathode Gas Recycle Blower	8,674	13
Cathode Heat Exchanger	26,713	41
Anode Recycle Blower	1,264	2
Anode Heat Exchanger	24,348	37
Oxy-Combustor	21,168	33
Total SOFC BOP	85,873	132
TOTAL SOFC POWER ISLAND	295,469	455
COAL & SORBENT HANDLING	60,254	93
COAL & SORBENT PREP & FEED	35,125	54
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	1,041,051	1602
ASU & oxidant compressor	62,412	96
Other gasification equip & foundations	92,954	143
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	1,196,417	1841
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	0	0
Single-Stage Selexol/MDEA	11,742	18
Claus Plant	142,414	219
Trace removal	4,245	7
COS Hydrolysis	45,092	69
Blowback, Piping, Foundations	8,032	12
Sulfur polishing/NG desulfurization	10,165	16
TOTAL GAS CLEANUP & PIPING	221,689	341
STEAM CYCLE		
HRSO, Ducting, and Stack	58,436	90
Steam Power System	53,177	82
Feedwater and Misc BOP systems	71,090	109
TOTAL STEAM CYCLE	182,703	281
CO₂ COMPRESSION & PURIFICATION	0	0
COOLING WATER SYSTEM	31,976	49
ASH/SPENT SORBENT HANDLING SYS	9,776	15
ACCESSORY ELECTRIC PLANT	50,874	78
INSTRUMENTATION & CONTROL	25,064	39
IMPROVEMENTS TO SITE	30,628	47
BUILDING & STRUCTURES	16,844	26
TOTAL PLANT COST (TPC)	2,156,627	3319
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	7,890	
1 Month Maintenance Materials	1,340	
1 Month Non-fuel Consumables	1,150	
1 Month Waste Disposal	219	
25% of 1 Months Fuel Cost at 100% CF	1,814	
2% of TPC	43,133	
Total Preproduction Costs	55,547	
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	16,730	
0.5% of TPC (spare parts)	10,783	
Total Inventory Capital	27,513	
Initial Cost for Catalyst and Chemicals	60,431	
Land	900	
Other Owner's Costs	323,494	
Financing Costs	58,229	
TOTAL OWNER'S COSTS	526,113	
TOTAL OVERNIGHT COST (TOC)	2,682,740	4129
TASC Multiplier	1.154	
TOTAL AS-SPENT COST (TASC)	3,095,882	4765

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-27. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – O&M Costs

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
OPERATING & MAINTENANCE LABOR		
Operating Labor Rate (base)	39.70	
Stack replacement O&M, \$/hr per stack kW	18.33	
Annual Operating Labor Cost	5,261,256	
Maintenance Labor Cost	7,362,411	
Administrative & Support Labor	3,155,917	
Property Taxes and Insurance	43,132,540	
TOTAL FIXED OPERATING COSTS	58,912,124	12.18
VARIABLE OPERATING COSTS		
Maintenance Material Cost	13,673,049	
Stack replacement		
SOFC stack replacement cost, \$/kW AC	231	
SOFC Stack replacement O&M, \$/yr per kW	23	
Stack Replacement Cost	14,960,768	3.09
CONSUMABLES		
Water		
Price (\$/1000 gal)	1.900	
Annual (\$)	423,989	
Chemicals		
MU & WT Chem.		
Price (\$/lb)	0.28	
Annual (\$)	374,802	
Carbon (Trace Removal)		
Price (\$/lb)	1.6	
Initial Cost (\$)	748,392	
Annual (\$)	318,067	
COS Catalyst		
Price (\$/m ³)	45,909	
Initial Cost (\$)	39,403,491	
Annual (\$)	8,373,242	
Selexol Solution		
Price (\$/gal)	38.00	
Initial Cost (\$)	16,609,722	
Annual (\$)	511,276	
Claus / DSRP Catalyst		
Price (\$/ft ³)	48	
Annual (\$)	23,613	
ZnO polishing sorbent		
Price (\$/lb)	1.5	
Initial Cost (\$)	642,157	
Annual (\$)	817,361	
COS Catalyst		
Price (\$/ft ³)	772	
Initial Cost (\$)	3,027,005	
Annual (\$)	890,296	
Sub Total Chemicals		
Initial Fill (\$)	60,430,767	
Annual (\$)	11,308,656	
Waste Disposal		

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
Spent Trace Catalyst		
Price (\$/lb)	0.42	
Annual (\$)	88,512	
Ash + HCl Sorbent		
Price (\$/ton)	16.31	
Annual (\$)	1,920,767	
Spent sorbents		
Price (\$/lb)	0.42	
Annual (\$)	228,861	
Subtotal-Waste Disposal (\$)	2,238,141	
TOTAL VARIABLE OPERATING COSTS		
Initial Fill (\$)	60,430,767	
Annual (\$)	42,604,603	8.81
Fuel Coal		
Price (\$/ton)	51.96	
Annual (\$)	74,012,853	15.30

Exhibit 4-28. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – Capital Cost Components

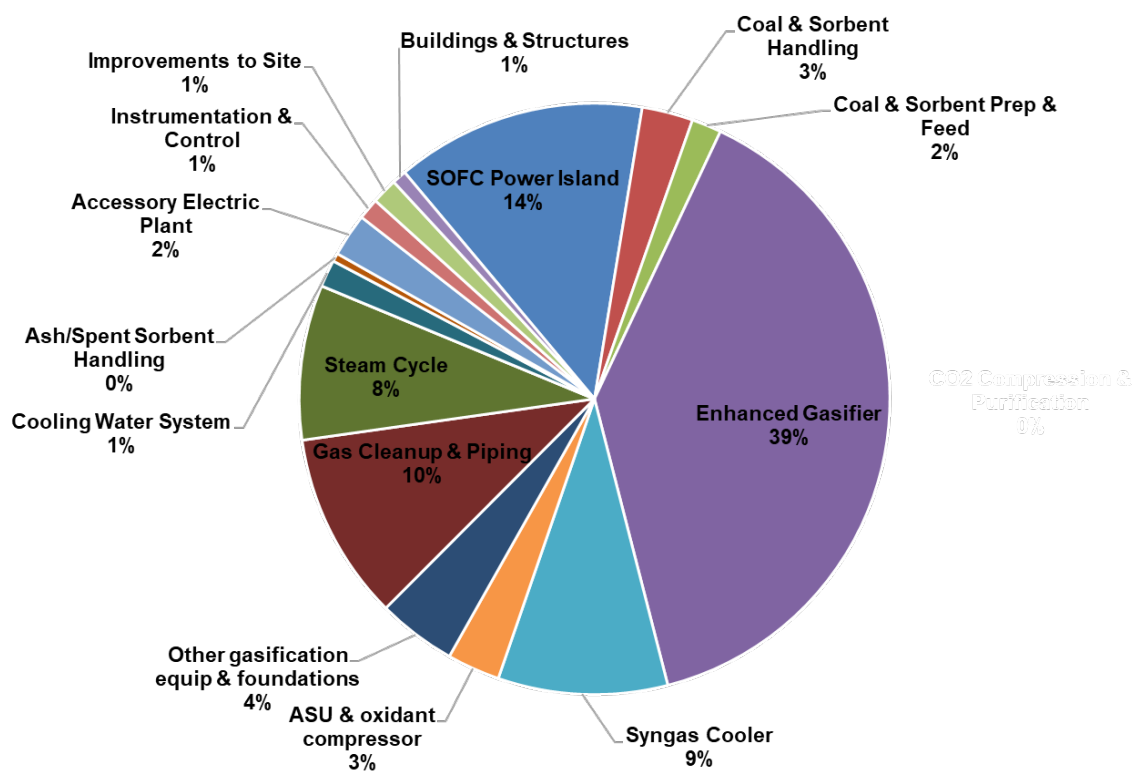
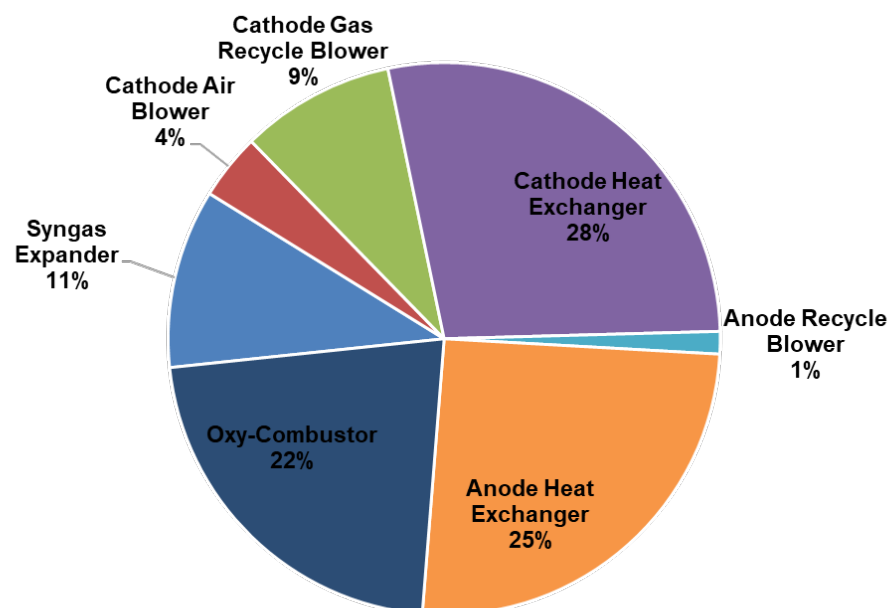


Exhibit 4-29. Case AENHIGFC3A – Enhanced Gasifier Atmospheric IGFC Plant without CCS – BOP Cost Distribution**Exhibit 4-30. Case 3 – Enhanced Gasifier Atmospheric IGFC Plant with and without CCS – Levelized Cost of Electricity**

Levelized Cost of Electricity (2018\$/MWh)		
Carbon Capture	With CCS	No CCS
Component	\$/MWh	
Variable Costs	25.6	24.1
<i>Fuel Costs</i>	16.7	15.3
<i>Variable O&M Costs</i>	9.0	8.8
Fixed O&M Costs	13.5	12.2
Capital Costs	50.5	45.2
Total LCOE (excluding T&S)	89.6	81.5
T&S	6.7	0.0
Total LCOE (including T&S)	96.3	81.5

4.3 CASE 6BV: ADVANCED GASIFIER ATMOSPHERIC IGFC WITH VGR

Case 6BV represents the culmination of the atmospheric pathway for IGFC plants with CCS, with the utilization of an advanced catalytic gasifier that results in a CH₄ content of ≈24 percent in the syngas entering the SOFC section. The VGR concept (only effective with CCS) that enables

97.5 percent fuel utilization through recirculation of the CPU vent gas is also implemented. The SOFC technology parameters are listed in Exhibit 4-31 and include BOP enhancements as well as a reduction in stack cost.

Exhibit 4-31. IGFC Parameters Used for the Advanced Gasifier Case 6BV

Parameter	Value
Gasifier Technology	Catalytic
Syngas CH ₄ content (mole %)	31.9%
SOFC Operating Pressure	Atmospheric
Cell Technology	Advanced Cell
Fuel Utilization, %	97.5
Current Density, mA/cm ²	400
Degradation, %/1000 h	0.2
Inverter Efficiency (%)	98
Stack Cost (\$/kW)	200
Capacity Factor (%)	85

The relevant process data for the numbered streams in the block flow diagram (BFD) for the Case 6BV plant without CCS shown in Exhibit 4-32 are tabulated in Exhibit 4-33. Preliminary runs indicated that a VGR recirculation rate of 90 percent is optimal. The higher syngas CH₄ content reduces the SOFC process air flow requirement by over 50 percent relative to the enhanced gasifier case.

Case 6BV Plant Performance

As shown in Exhibit 4-34, the Case 6BV plants results in the highest efficiency, 61.1 percent, along the pathway, ≈ 19 percentage points higher than that of the reference plant and 14–29 percentage points higher than conventional technologies, mainly due to the combined result of higher syngas CH₄ content and higher fuel utilization enabled by the VGR. In contrast to other cases, the VGR eliminates the CCS penalty - efficiency of Case 6BV plant with CCS is even slightly greater than the efficiency of the comparable advanced gasifier plant without CCS (Case 6 with 61 percent efficiency). The carbon capture is somewhat reduced in this case to 94.3 percent due to the circulating vent gas flows.

The SOFC power island, due to the higher fuel utilization, generates a desirably larger share (91 percent) of the gross power, relative to the reference plant, as seen in Exhibit 4-35. The ASU auxiliary load is reduced due to the elimination of the oxycombustor in this case and the auxiliary load is dominated by the CPU and the steam cycle as shown in Exhibit 4-36. The VGR slightly increases the CPU auxiliary load relative to the reference case due to the larger recirculation flows.

The salient material and energy balances for the plant with CCS are listed in Exhibit 4-37, which, shows that the emission of the IGFC plant even without CCS is much lower than those of other

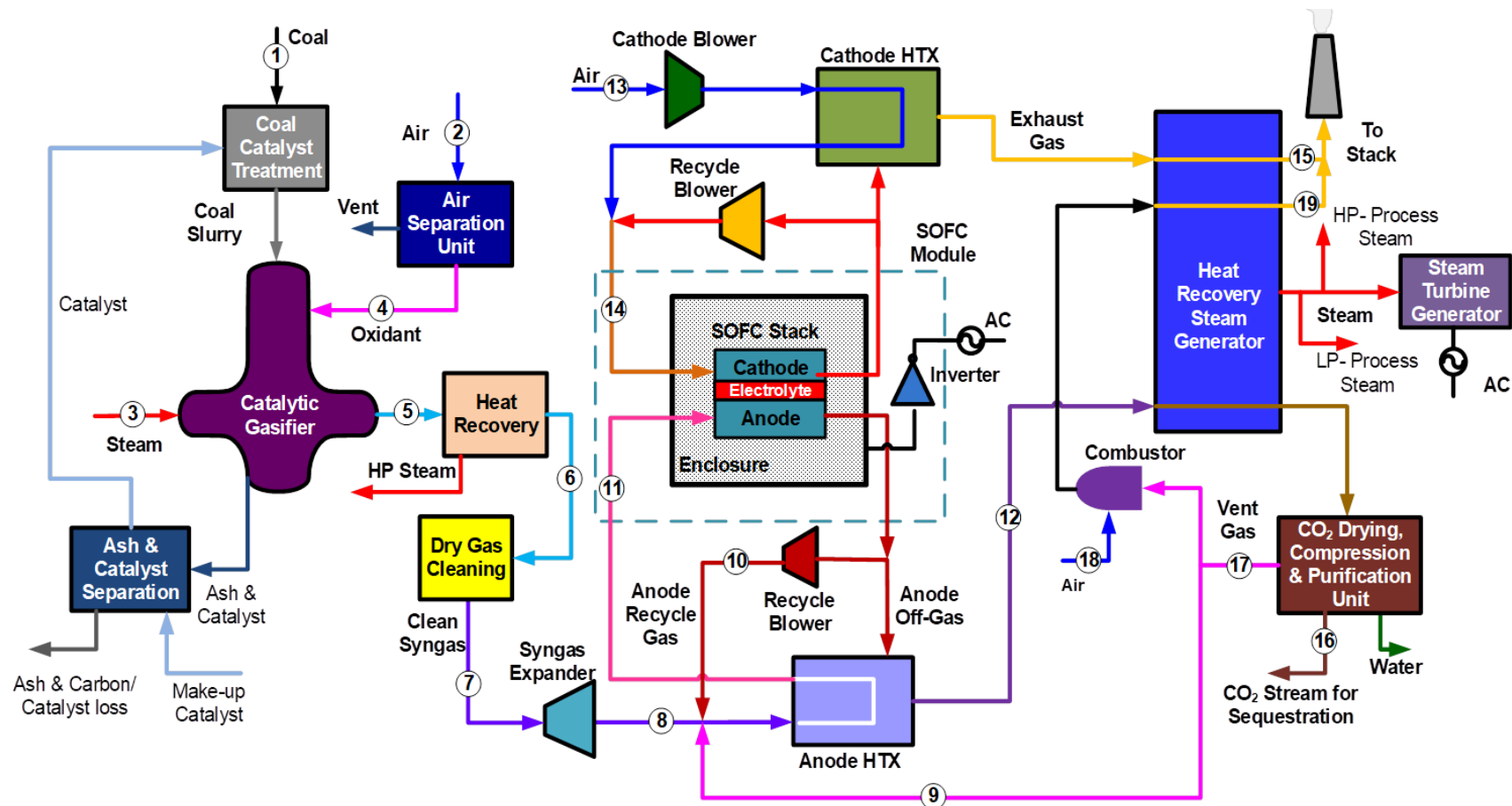
conventional coal-based technologies. The IGFC plant water consumption is also significantly lower than the corresponding value for conventional fossil fuel power plant technologies. The heat and mass balance diagrams for the gasifier, ASU, and syngas clean-up, the power island, and CPU are shown in Exhibit 4-38, Exhibit 4-39, and Exhibit 4-40, respectively.

Case 6BV Plant Costs

The capital and O&M costs for the Case 6BV IGFC plant are itemized in Exhibit 4-41 and Exhibit 4-42 respectively. The lower cost of the catalytic gasifier along with the highly efficient system reduce the gasifier and clean-up cost contributions to 50 percent of the capital costs as shown in Exhibit 4-43. The BOP cost distribution is shown in Exhibit 4-44.

The LCOE for the Case 6 plant is \$58.8/MWh, as shown in Exhibit 4-45, which presents the least cost of electricity amongst all the IGFC plants with CCS considered in this report.

Exhibit 4-32. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Process Diagram



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-33. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
V-L Mole Fraction																			
Ar	0.0000	0.0094	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0005	0.0001	0.0002	0.0001	0.0094	0.0100	0.0107	0.0000	0.0005	0.0094	0.0066
CH ₄	0.0000	0.0000	0.0000	0.0000	0.1974	0.1974	0.3190	0.3190	0.0000	0.0000	0.0747	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0565	0.0565	0.0914	0.0914	0.0489	0.0802	0.0785	0.0802	0.0000	0.0000	0.0000	0.0000	0.0489	0.0000	0.0000
CO ₂	0.0000	0.0003	0.0000	0.0000	0.2120	0.2120	0.3441	0.3441	0.2894	0.3479	0.3389	0.3479	0.0003	0.0003	0.0003	1.0000	0.2894	0.0003	0.1559
COS	0.0000	0.0000	0.0000	0.0000	0.0001	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
H ₂	0.0000	0.0000	0.0000	0.0000	0.1475	0.1475	0.2384	0.2384	0.5595	0.0995	0.1960	0.0995	0.0000	0.0000	0.0000	0.0000	0.5595	0.0000	0.0000
H ₂ O	0.0000	0.0104	1.0000	0.0000	0.3742	0.3742	0.0004	0.0004	0.0000	0.4422	0.2772	0.4422	0.0104	0.0111	0.0119	0.0000	0.0000	0.0104	0.2645
HCl	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	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	0.0000	0.0000	0.0000	0.0000	0.0071	0.0071	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
N ₂	0.0000	0.7722	0.0000	0.0000	0.0038	0.0038	0.0067	0.0067	0.1017	0.0301	0.0346	0.0301	0.7722	0.8234	0.8820	0.0000	0.1017	0.7722	0.5718
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0007	0.0007	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	0.2077	0.0000	0.9999	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.1551	0.0951	0.0000	0.0000	0.2077	0.0012
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	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	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	0	4,241	10,605	735	15,286	15,286	9,441	9,441	5,602	25,248	40,291	21,066	59,389	111,388	51,998	6,912	6,224	920	1,353
V-L Flowrate (kg/hr)	0	122,373	191,047	23,510	328,836	328,836	221,854	221,854	101,416	670,875	994,141	559,760	1,713,628	3,190,761	1,477,133	304,216	112,684	26,534	37,803
Solids Flowrate (kg/hr)	141,283	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	15	291	157	704	204	371	53	149	721	649	509	15	595	132	20	32	15	149
Pressure (MPa, abs)	0.10	0.10	7.58	7.58	6.72	6.62	5.61	0.14	0.11	0.14	0.14	0.13	0.10	0.1	0.099	15.272	0.116	0.101	0.101
Enthalpy (kJ/kg) ^A	---	31.06	2,860.25	135.67	2,108.19	908.36	608.96	76.51	263.38	1,808.15	1,529.58	1,470.31	31.06	648.3	154.657	-251.493	54.928	31.057	596.392
Density (kg/m ³)	---	1.2	38.4	67.9	17.7	42.0	24.3	1.2	0.6	0.4	0.4	0.6	1.2	0.4	0.8	783.8	0.8	1.2	0.8
V-L Molecular Weight	---	28.854	18.015	31.999	21.512	21.512	23.498	23.498	18.104	26.571	24.674	26.571	28.854	29	28.407	44.010	18.104	28.854	27.946
V-L Flowrate (lb _{mol} /hr)	0	9,350	23,379	1,620	33,701	33,701	20,815	20,815	12,350	55,662	88,826	46,443	130,931	245,567	114,637	15,239	13,722	2,027	2,982
V-L Flowrate (lb/hr)	0	269,785	421,186	51,831	724,958	724,958	489,104	489,104	223,583	1,479,027	2,191,706	1,234,060	3,777,903	7,034,424	3,256,522	670,682	248,426	58,498	83,341
Solids Flowrate (lb/hr)	311,475	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	59	555	314	1,300	400	700	127	300	1,329	1,200	949	59	1,103	270	67	90	59	300
Pressure (psia)	14.7	14.7	1,100.0	1,100.0	975.0	960.0	814.0	19.7	16.2	19.7	19.7	19.6	14.7	14.9	14.3	2,215.0	16.8	14.7	14.7
Enthalpy (Btu/lb) ^A	---	13.4	1,229.7	58.3	906.4	390.5	261.8	32.9	113.2	777.4	657.6	632.1	13.4	278.7	66.5	-108.1	23.6	13.4	256.4
Density (lb/ft ³)	---	0.076	2.398	4.240	1.102	2.623	1.518	0.074	0.036	0.027	0.027	0.034	0.076	0	0.052	48.934	0.052	0.076	0.050

^A Reference conditions are 32.02 F and 0.089 psia

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-34. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Performance Summary

Carbon Capture	With CCS
Power Summary (Gross Power at Generator Terminals, kW_e)	
SOFC Power	647,900
Natural Gas/Syngas Expander Power	31,800
Steam Turbine Power	35,600
Total Gross Power, kW_e	715,300
Auxiliary Load Summary, kW_e	
Coal Handling	390
Coal Milling	454
Sour Water Recycle Slurry Pump	1,765
Ash Handling	972
Air Separation Unit Auxiliaries	129
Air Separation Unit Main Air Compressor	1,040
Oxygen Compressor	0
Nitrogen Compressors	4,893
Claus Plant TG Recycle Compressor	3,100
CO ₂ Compressor	0
CO ₂ Refrigeration	1,190
Boiler Feedwater Pumps	34,770
Condensate Pump	0
Syngas Recycle Compressor	618
Circulating Water Pump	46
Ground Water Pumps	1,030
Cooling Tower Fans	334
Scrubber Pumps	530
Quench Water Pump	214
Selexol Auxiliary Power	1,978
Steam Turbine Auxiliaries	15
Claus Plant/TGTU Auxiliaries	161
Gas Turbine Auxiliaries	7
Cathode air blower	3,180
Cathode recycle blower	2,470
Anode recycle blower	400
Miscellaneous Balance of Plant ^A	3,049
Transformer Losses	2,330
Total Auxiliaries, kW_e	65,066
Net Power, kW _e	650,234
Net Plant Efficiency, % (HHV)	61.1%
CO ₂ Capture Rate, %	94.3%
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	5,896 (5,588)
Coal Feed Rate, kg/hr (lb/hr)	141,283 (311,475)
Thermal Input ^B , kW _t	1,064,923
Condenser Cooling Duty, 10 ⁶ kJ/hr (10 ⁶ Btu/hr)	211 (200)
Raw Water Withdrawal, m ³ /min (gpm)	3.5 (922)
Raw Water Consumption, m ³ /min (gpm)	2.6 (690)

^A Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads.

^B HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb).

Exhibit 4-35. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Gross Power Distribution

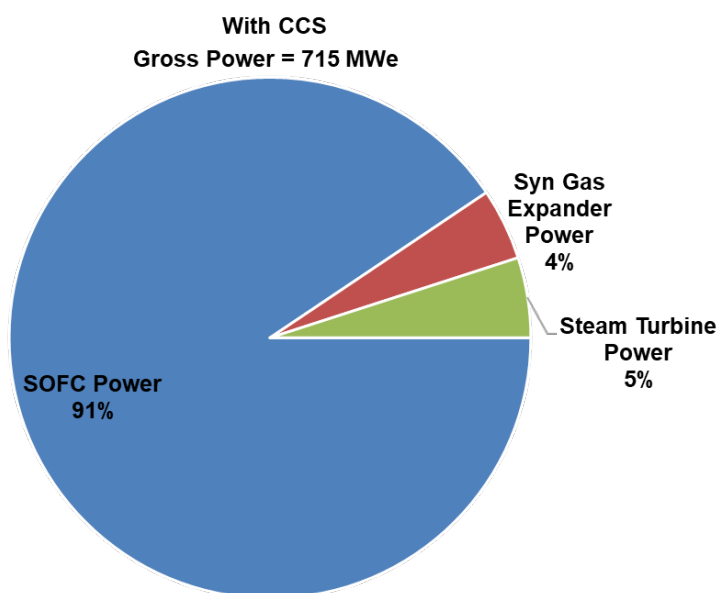


Exhibit 4-36. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Auxiliary Load Distribution

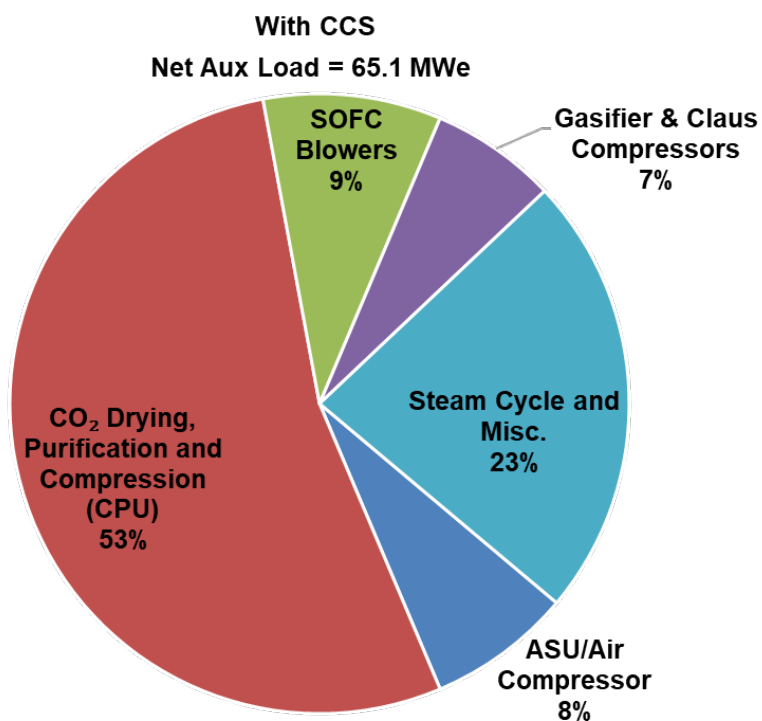


Exhibit 4-37. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Material and Energy Balances

Carbon Balance

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	90,060 (198,549)	Slag	4,665 (10,285)
Air (CO ₂)	233 (513)	Stack Gas	214 (472)
Cat Carbonator	161 (354)	CO ₂ Product	83,026 (183,040)
		N ₂ Product	15 (34)
		Carbonator Vent	3 (7)
		Vent Gas	2,533 (5,584)
		Convergence Tolerance	-3 (-6)
Total	90,453 (199,416)	Total	90,453 (199,416)

Sulfur Balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	3,541 (7,807)	Elemental Sulfur	3,539 (7,803)
		Polishing Sorbent	2 (4)
		Convergence Tolerance	0 (0)
Total	3,541 (7,807)	Total	3,541 (7,807)

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 (115)	0.43 (115)			
Slurry Water	0.05 (13)	0.05 (13)			
Quench/Wash	2.7 (724)	2.74 (724)			
Condenser Makeup	3.2 (857)	0.0 (0)	3.2 (857)	0.0 (0)	3.2 (857)
Gasifier Steam	3.2 (842)	0.0 (0)	3.2 (842)		
BFW Makeup	0.1 (15)	0.0 (0)	0.1 (15)		
Cooling Tower	3.9 (1,034)	3.67 (969)	0.2 (65)	0.9 (233)	-0.6 (-167)
CO ₂ Dehydration	0.0 (0)	2.38 (630)	-2.38 (-630)		
SWS Excess Water	0.0 (0)	1.28 (339)	-1.28 (-339)		
Total	10.4 (2,744)	6.89 (1,821)	3.5 (922)	0.9 (233)	2.6 (690)

Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	3,834 (3,634)	3.2 (3.0)		3,837 (3,637)
ASU Air		3.8 (3.6)		4 (4)
Fuel cell Air		54.0 (51.2)		54 (51)
Raw Water Makeup		13.1 (12.4)		13 (12)
Auxiliary Power			234 (222)	234 (222)
TOTAL	3,834 (3,634)	74.2 (70.3)	234 (222)	4,142 (3,926)
Heat Out GJ/hr (MMBtu/hr)				
Slag	153 (145)	14.9 (14.1)		168 (159)
Sulfur	33 (31)	0.4 (0.3)		33 (31)
CO ₂		-76.5 (-72.5)		-77 (-73)
Cooling Tower Blowdown		6.5 (6.2)		7 (6)
Exhaust		251.0 (237.9)		251 (238)
Condenser		208 (197)		208 (197)
Non-Condenser Cooling Tower Loads*		301 (285)		301 (285)
Process Losses*		674 (639)		674 (639)
Difference**		3 (2)		3 (2)
Power			2,575 (2,441)	2,575 (2,441)
TOTAL	186 (176)	1,381 (1,309)	2,575 (2,441)	4,142 (3,926)

Emissions

	kg/GJ (lb/10 ⁶ Btu)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO ₂	0 (0)	0 (0)	0 (0)
NO _x	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO ₂	21 (49)	596,282 (657,288)	112 (247)

Exhibit 4-38. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Heat and Material Balance Diagram – ASU, Gasifier, and Clean-up

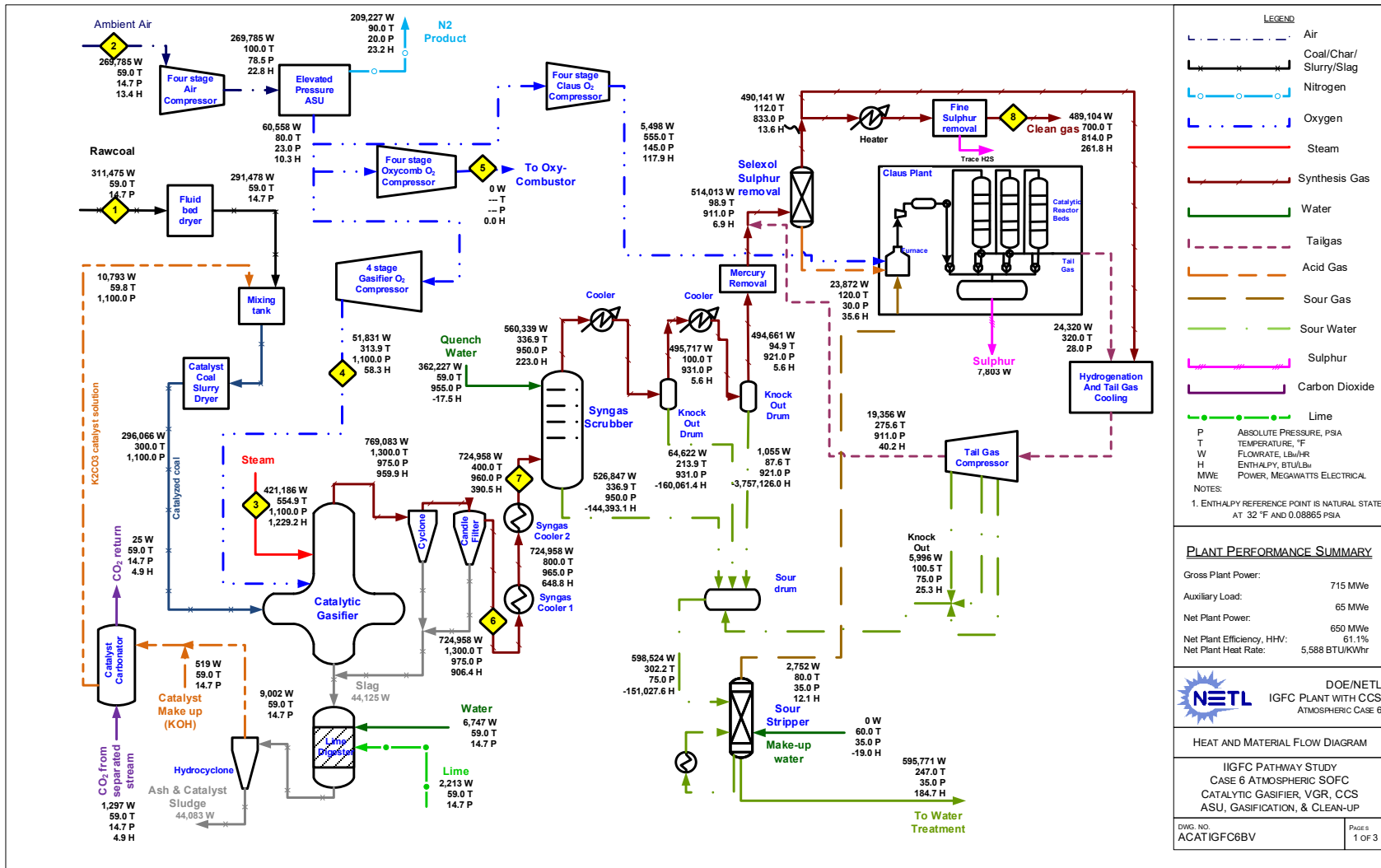


Exhibit 4-39. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Heat and Material Balance Diagram – Power Block

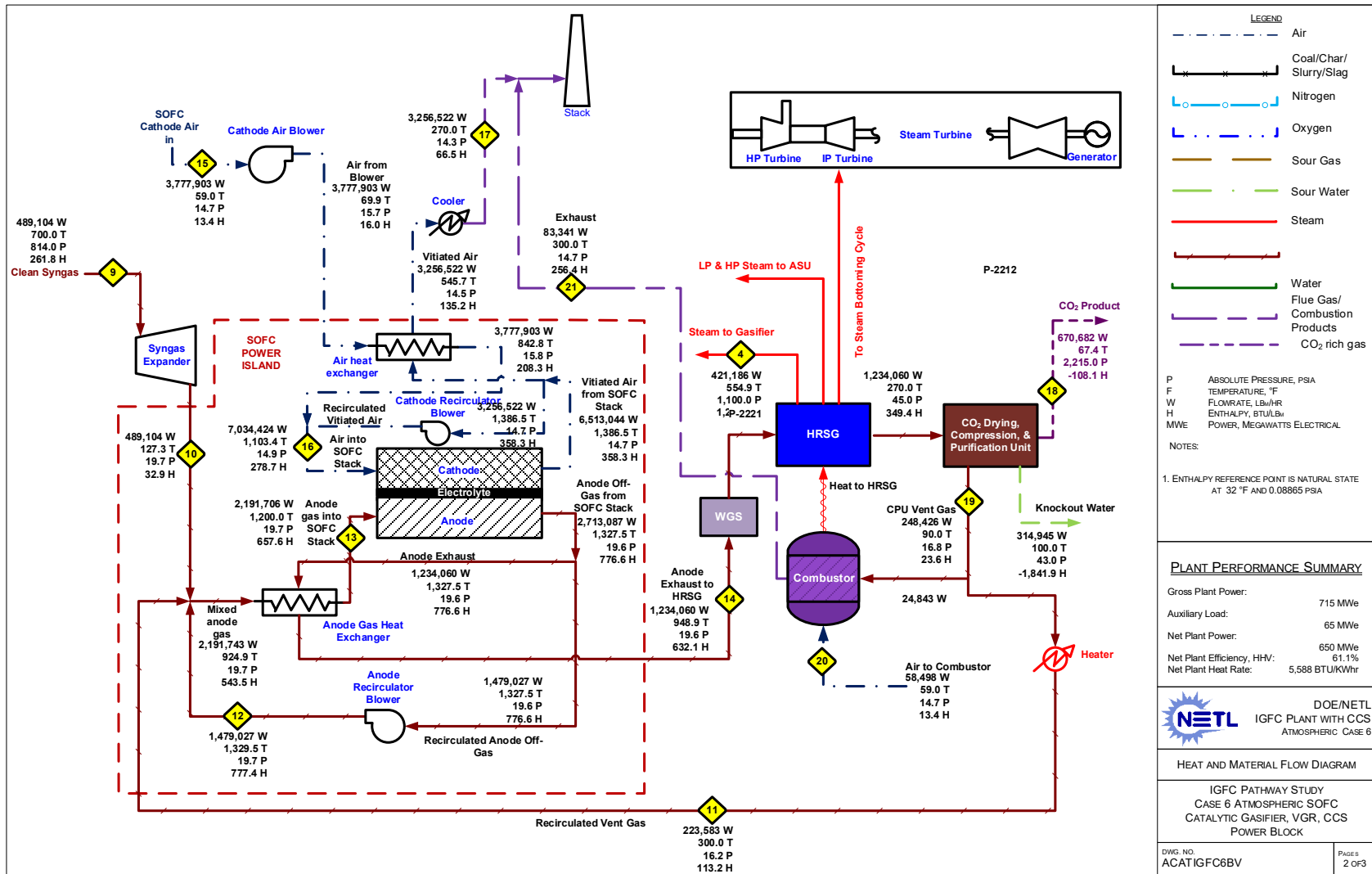
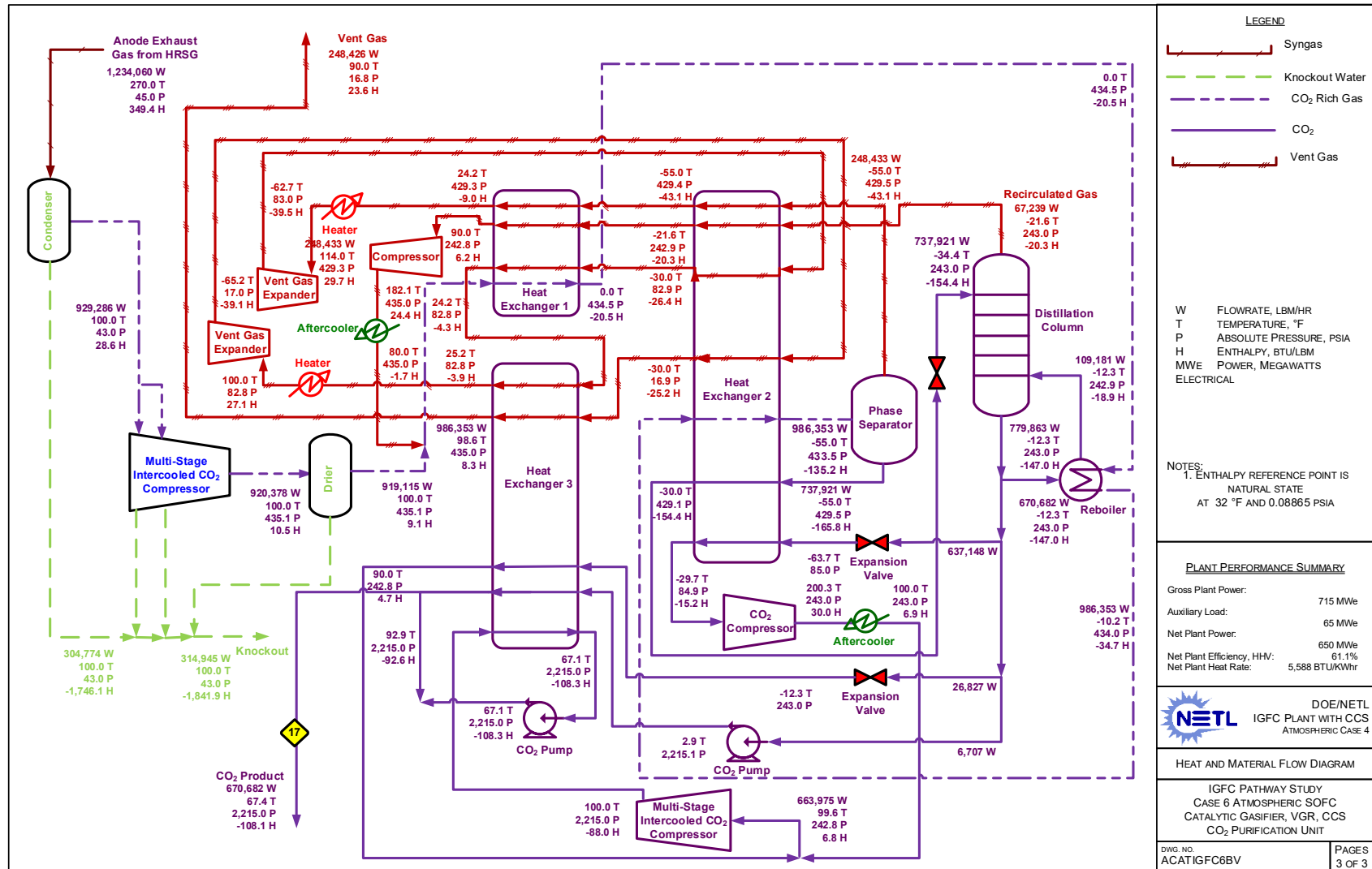


Exhibit 4-40. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Heat and Material Balance Diagram – CO₂ Dehydration, Purification, and Compression



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-41. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC plant with VGR – Capital Costs

Cost Component	Cost (\$1000)	Specific Cost (\$/kWe AC)
	2018\$	2018\$
SOFC Module		
SOFC Stack	136,357	210
Container	6,053	9
Insulation	2,727	4
Module Assembly	6,818	10
Air Distribution	6,818	10
Fuel Distribution	6,818	10
Module Current Collectors	3,409	5
Module I&C	3,409	5
Inverter	46,362	71
Total SOFC Module with 10 % Extra Installed Area	218,770	336
Syngas Expander	7,745	12
SOFC BOP		
Cathode Air Blower	2,273	3
Cathode Gas Recycle Blower	5,204	8
Cathode Heat Exchanger	22,119	34
Anode Recycle Blower	1,180	2
Anode Heat Exchanger	23,501	36
Oxy-Combustor	0	0
Total SOFC BOP	54,277	83
TOTAL SOFC POWER ISLAND	280,793	432
COAL & SORBENT HANDLING	53,041	82
COAL & SORBENT PREP & FEED	44,736	69
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	522,810	804
ASU & oxidant compressor	22,447	35
Other gasification equip & foundations	91,829	141
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	637,086	980
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	0	0
Single-Stage Selexol/MDEA	5,003	8
Claus Plant	124,081	191
Trace removal	3,029	5
COS Hydrolysis	14,582	22
Blowback, Piping, Foundations	6,441	10
Sulfur polishing/NG desulfurization	4,591	7
TOTAL GAS CLEANUP & PIPING	157,728	243
STEAM CYCLE		
HRSF, Ducting, and Stack	10,373	16
Steam Power System	27,986	43
Feedwater and Misc BOP systems	59,583	92
TOTAL STEAM CYLCE	97,942	151
CO₂ COMPRESSION & PURIFICATION	56,321	87
COOLING WATER SYSTEM	20,661	32
ASH/SPENT SORBENT HANDLING SYS	30,500	47
ACCESSORY ELECTRIC PLANT	47,417	73
INSTRUMENTATION & CONTROL	24,826	38
IMPROVEMENTS TO SITE	30,350	47
BUILDING & STRUCTURES	15,014	23
TOTAL PLANT COST (TPC)	1,496,126	2301
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	6,481	
1 Month Maintenance Materials	930	
1 Month Non-fuel Consumables	724	
1 Month Waste Disposal	42	
25% of 1 Months Fuel Cost at 100% CF	1,477	
2% of TPC	29,923	
Total Preproduction Costs	39,575	
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	13,165	
0.5% of TPC (spare parts)	7,481	
Total Inventory Capital	20,646	
Initial Cost for Catalyst and Chemicals	25,954	
Land	900	
Other Owner's Costs	224,419	
Financing Costs	40,395	
TOTAL OWNER'S COSTS	351,890	
TOTAL OVERNIGHT COST (TOC)	1,848,016	2842
TASC Multiplier	1.154	
TOTAL AS-SPENT COST (TASC)	2,132,610	3280

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-42. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – O&M Costs

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
OPERATING & MAINTENANCE LABOR		
Operating Labor Rate (base)	39.70	
Stack replacement O&M, \$/hr per stack kW	18.33	
Annual Operating Labor Cost	5,261,256	
Maintenance Labor Cost	5,107,557	
Administrative & Support Labor	2,592,203	
Property Taxes and Insurance	29,922,525	
TOTAL FIXED OPERATING COSTS	42,883,542	8.86
VARIABLE OPERATING COSTS		
Maintenance Material Cost	9,485,464	
Stack replacement		
SOFC stack replacement cost, \$/kW AC	210	
SOFC Stack replacement O&M, \$/yr per kW	21	
Stack Replacement Cost	13,597,041	2.81
CONSUMABLES		
Water		
Price (\$/1000 gal)	1.900	
Annual (\$)	497,142	
Chemicals		
MU & WT Chem.		
Price (\$/lb)	0.28	
Annual (\$)	337,492	
Carbon (Trace Removal)		
Price (\$/lb)	1.6	
Initial Cost (\$)	609,267	
Annual (\$)	258,938	
COS Catalyst		
Price (\$/m ³)	45,909	
Initial Cost (\$)	9,609,401	
Annual (\$)	2,041,998	
Selexol Solution		
Price (\$/gal)	38.00	
Initial Cost (\$)	13,521,989	
Annual (\$)	416,230	
Claus / DSRP Catalyst		
Price (\$/ft ³)	48	
Annual (\$)	19,224	
ZnO polishing sorbent		
Initial Fill (lb)		
Price (\$/lb)	1.5	
Initial Cost (\$)	211,987	
Annual (\$)	665,414	
KOH coal catalyst makeup		
Initial Fill (ft ³)	747,141	
Use (ft ³ /day)	12,452	
Price (\$/ft ³)	0.19	
Initial Cost (\$)	142,660	
Annual (\$)	737,670	
Lime for catalyst recovery		
Initial Fill (lb)	3,186,613	
Use (lb/day)	53,110	
Price (\$/lb)	0.12	
Initial Cost (\$)	382,394	
Annual (\$)	1,977,293	

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
Sub Total Chemicals		
Initial Fill (\$)	25,954,099	
Annual (\$)	6,888,495	
Waste Disposal		
Spent Trace Catalyst		
Price (\$/lb)	0.42	
Annual (\$)	72,058	
Ash + HCl Sorbent		
Price (\$/ton)	16.31	
Annual (\$)	165,883	
Spent sorbents		
Price (\$/lb)	0.42	
Annual (\$)	186,316	
Subtotal-Waste Disposal (\$)	424,257	
TOTAL VARIABLE OPERATING COSTS		
Initial Fill (\$)	25,954,099	
Annual (\$)	30,892,399	6.38
Fuel Coal		
Price (\$/ton)	51.96	
Annual (\$)	60,253,929	12.44

Exhibit 4-43. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – Capital Cost Components

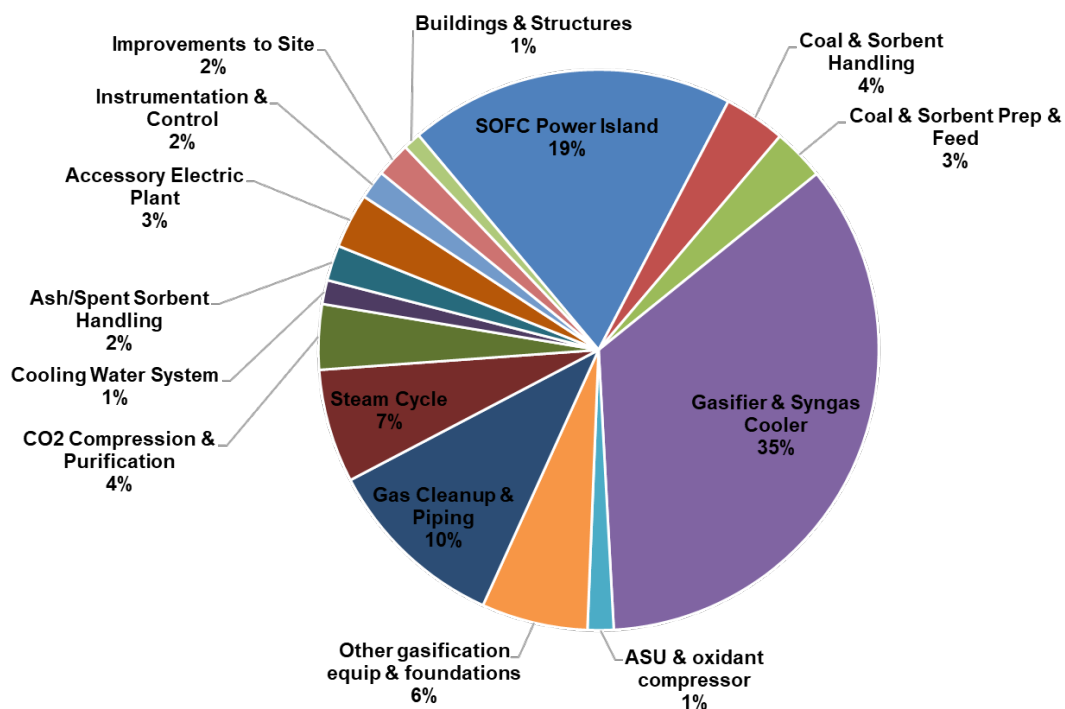


Exhibit 4-44. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC Plant with VGR – BOP Cost Distribution

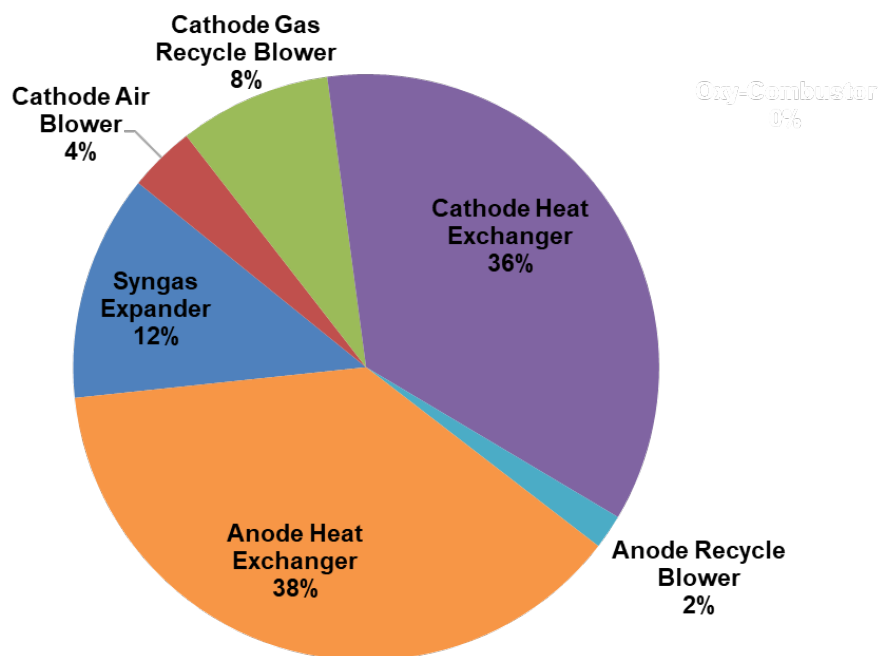


Exhibit 4-45. Case ACATIGFC6BV – Advanced Gasifier Atmospheric IGFC plant with VGR – Levelized Cost of Electricity

Levelized Cost of Electricity (2018\$/MWh)	
Carbon Capture	With CCS
Component	\$/MWh
Variable Costs	18.8
<i>Fuel Costs</i>	12.4
<i>Variable O&M Costs</i>	6.4
Fixed O&M Costs	8.9
Capital Costs	31.1
Total LCOE (excluding T&S)	58.8
T&S	4.7
Total LCOE (including T&S)	63.5

4.4 PATHWAY 1 RESULTS

The performance and costs of the atmospheric pathway IGFC plants are summarized in Exhibit 4-46, Exhibit 4-47, Exhibit 4-48, and Exhibit 4-49 for plants with and without CCS. The efficiency varies from 43 percent in the reference case to 61.1 percent for the atmospheric plants with CCS, while the LCOE declines from a value of \$97.5/MWh to a value of \$58.8/MWh, which represents a 40 percent reduction in costs through technological advancements along the pathway. For plants without CCS, the pathway results in a 38 percent reduction in LCOE along with an improvement of 23 percentage points in plant efficiency. These results will be discussed further in comparison with the pressurized pathway and the conventional technologies in Section 5.

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-46. Pathway 1 – Atmospheric IGFC Plants with CCS Performance Summary

Case		Case 0	Case 1	Case 2	Case 2V	Case 3	Case 3V	Case 4	Case 5	Case 6	Case 6V
Gasifier Technology		Conventional				Enhanced		Catalytic			
SOFC Degradation Rate (%/1000 h)		0.2									
Fuel Utilization (%)		80	85		97.5	85	97.5	85			97.5
Capacity Factor (%)		80		85				80	85		
Inverter Efficiency (%)		97								98	
SOFC Stack Cost (\$/kW)		225								200	
Performance											
SOFC Parameters	Current Density (mA/cm²)	400	400	400	400	400	400	400	400	400	400
	Cell Potential (V)	0.858	0.849	0.849	0.849	0.844	0.849	0.852	0.852	0.852	0.834
	Power Density (mW/cm²)	343	340	340	340	338	340	341	341	341	334
Gross Power (MWe)		804	799	799	795	790	779	713	713	711	715
Auxiliary Loads (MWe)		154	149	149	145	140	129	62	62	60	65
Air Separation Unit (kWe)		44,393	40,694	40,694	28,903	36,342	23,996	6,420	6,420	5,081	4,893
CO ₂ Drying, Purification and Compression (CPU) (kWe)		53,540	52,400	52,400	64,810	50,570	61,220	29,510	29,510	29,180	34,770
SOFC Blowers (kWe)		16,230	17,400	17,400	16,040	16,160	10,680	5,090	5,090	5,020	6,050
Compressors (kWe)		16,570	16,220	16,220	15,130	15,310	14,160	7,875	7,875	7,778	6,818
Steam Cycle and Miscellaneous (kWe)		23,432	22,414	22,414	20,495	21,225	19,273	43,704	43,704	43,263	48,046
Net Power (MWe)		650	650	650	650	650	650	650	650	650	650
Coal Flowrate (lb/hr)		442,400	432,950	432,950	403,450	417,800	378,800	327,800	327,800	324,075	311,475
Net Electric Efficiency, HHV (%)		43.0	43.9	43.9	47.1	45.5	50.2	58.0	58.0	58.7	61.1
Net Plant Heat Rate, HHV (Btu/kWh)		7,942	7,774	7,774	7,242	7,497	6,800	5,881	5,881	5,815	5,588
CO ₂ Capture rate (%)		98.8	98.7	98.7	96.1	98.7	95.9	98.6	98.6	98.6	94.3
CO ₂ Captured (tonnes per year)		3,546,910	3,471,017	3,687,955	3,345,812	3,558,637	3,134,435	2,513,636	2,670,739	2,640,389	2,496,945
CO ₂ Emissions (lb/MWhgross)		20.6	20.9	20.9	50.0	20.3	49.0	16.1	16.1	15.9	246.8
CO ₂ Emissions (lb/MWhnet)		25.4	25.6	25.6	61.2	24.6	58.7	17.6	17.6	17.4	271.5
Raw Water Consumption (gpm/MWhnet)		2.36	2.06	2.06	1.73	1.74	1.51	1.50	1.50	1.48	1.06

Exhibit 4-47. Pathway 1 – Atmospheric IGFC Plants with CCS Cost Summary

Case	Case 0	Case 1	Case 2	Case 2V	Case 3	Case 3V	Case 4	Case 5	Case 6	Case 6V
Gasifier Technology	Conventional				Enhanced		Catalytic			
SOFC Degradation Rate (%/1000 h)	0.2									
Fuel Utilization (%)	80	85		97.5	85	97.5	85			97.5
Capacity Factor (%)	80		85				80	85		
Inverter Efficiency (%)	97								98	
SOFC Stack Cost (\$/kW)	225								200	
Cost										
Total Plant Cost (TPC) (1000\$)	2,500,963	2,469,563	2,469,563	2,370,701	2,421,613	2,276,869	1,556,032	1,556,032	1,532,178	1,496,126
Total Overnight Cost (TOC) (1000\$)	3,094,997	3,055,481	3,055,364	2,942,472	2,995,692	2,825,595	1,922,481	1,922,408	1,893,174	1,848,016
Total As-Spent Cost (TASC) (1000\$)	3,571,627	3,526,025	3,525,890	3,395,613	3,457,029	3,260,737	2,218,544	2,218,459	2,184,723	2,132,610
Levelized Cost of Electricity (\$/MWh)										
Variable Costs	27.3	26.8	26.4	25.4	25.6	24.1	20.3	20.0	19.4	18.8
Fuel Costs	17.7	17.3	17.3	16.1	16.7	15.1	13.1	13.1	13.0	12.4
Variable O&M Costs	9.6	9.5	9.1	9.2	9.0	9.0	7.2	6.9	6.5	6.4
Fixed O&M Costs	14.8	14.6	13.7	13.2	13.5	12.8	9.7	9.2	9.0	8.9
Capital Costs	55.4	54.8	51.5	49.6	50.5	47.6	34.4	32.4	31.9	31.1
Total LCOE (excluding T&S)	97.5	96.2	91.7	88.2	89.6	84.5	64.4	61.5	60.4	58.8
T&S	7.1	6.9	6.9	6.3	6.7	5.9	5.0	5.0	4.9	4.7
Total LCOE (including T&S)	104.6	103.1	98.6	94.5	96.3	90.4	69.4	66.5	65.3	63.5

Exhibit 4-48. Pathway 1 – Atmospheric IGFC Plants without CCS Performance Summary

Case		Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology		Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)		0.2						
Fuel Utilization (%)		80	85					
Capacity Factor (%)		80		85		80	85	
Inverter Efficiency (%)		97						98
SOFC Stack Cost (\$/kW)		225						200
Performance								
SOFC Parameters	Current Density (mA/cm²)	400	400	400	400	400	400	400
	Cell Potential (V)	0.858	0.849	0.849	0.844	0.852	0.852	0.852
	Power Density (mW/cm²)	343	340	340	338	341	341	341
Gross Power (MWe)		728	728	728	720	681	681	680
Auxiliary Loads (MWe)		79	78	78	70	31	31	29
Air Separation Unit (kWe)		28,694	28,249	28,249	24,213	6,182	6182	4896
CO ₂ Drying, Purification and Compression (CPU) (kWe)		0	0	0	0	0	0	0
SOFC Blowers (kWe)		14,750	15,860	15,860	14,850	4,900	4900	4840
Compressors (kWe)		15,020	14,780	14,780	12,720	7,581	7581	7505
Steam Cycle and Miscellaneous (kWe)		20,130	19,356	19,356	18,219	13,065	13,065	12,975
Net Power (MWe)		650	650	650	650	650	650	650
Coal Flowrate (lb/hr)		400,525	394,300	394,300	382,600	315,012	315,012	311,700
Net Electric Efficiency, HHV (%)		47.5	48.2	48.2	49.7	60.4	60.4	61.0
Net Plant Heat Rate, HHV (Btu/kWh)		7,190	7,082	7,082	6,870	5,654	5,654	5,593
CO ₂ Capture rate (%)		0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Captured (tonnes per year)		0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Emissions (lb/MWhgross)		1278.7	1260.4	1260.4	1236.5	1028.5	1028.5	1019.8
CO ₂ Emissions (lb/MWhnet)		1433.3	1412.2	1412.2	1369.7	1077.5	1077.5	1066.0
Raw Water Consumption (gpm/MWnet)		1.47	1.23	1.23	0.91	1.07	1.07	1.05

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 4-49. Pathway 1 – Atmospheric IGFC Plants without CCS Cost Summary

Case	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology	Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)	0.2						
Fuel Utilization (%)	80	85					
Capacity Factor (%)	80		85		80	85	
Inverter Efficiency (%)	97						98
SOFC Stack Cost (\$/kW)	225						200
Cost							
Total Plant Cost (TPC) (1000\$)	2,211,045	2,189,424	2,189,424	2,156,627	1,455,125	1,455,125	1,426,953
Total Overnight Cost (TOC) (1000\$)	2,754,017	2,723,013	2,722,897	2,682,740	1,799,159	1,799,089	1,764,708
Total As-Spent Cost (TASC) (1000\$)	3,178,135	3,142,357	3,142,224	3,095,882	2,076,230	2,076,149	2,036,473
Levelized Cost of Electricity (\$/MWh)							
Variable Costs	25.5	25.0	24.7	24.1	19.5	19.2	18.7
Fuel Costs	16.0	15.8	15.8	15.3	12.6	12.6	12.5
Variable O&M Costs	9.5	9.3	8.9	8.8	6.9	6.6	6.2
Fixed O&M Costs	13.2	13.1	12.3	12.2	9.2	8.7	8.5
Capital Costs	49.3	48.8	45.9	45.2	32.2	30.3	29.7
Total LCOE (excluding T&S)	88.1	87.0	82.9	81.5	60.9	58.2	56.9
T&S	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total LCOE (including T&S)	88.1	87.0	82.9	81.5	60.9	58.2	56.9

5 SCENARIO 2 – IGFC WITH PRESSURIZED SOFC

Scenario 2 utilizes the high-pressure syngas from the gasifier with a pressurized SOFC unit. Pressurized SOFC can be configured in two general, alternative arrangements:

1. The anode off-gas oxy-combustor is followed by hot gas expander power generation (expansion ratio about 18). An HRSG produces steam for power generation, and the remaining low-pressure, wet CO₂ stream is dried and compressed (compression ratio about 149).
2. The anode off-gas oxy-combustor is followed directly by an HRSG for steam bottoming power generation. The remaining high-pressure, wet CO₂ stream is dried and compressed (compression ratio about 8.4).

Configuration 2 is expected to be the least complex and most effective approach and is utilized for this evaluation. All areas of the plant are identical to the Case 1 plant areas except for the power island and the CO₂ dehydration and compression area.

5.1 REFERENCE PRESSURIZED IGFC

The reference pressurized case uses the conventional E-Gas™ gasifier combined with an SOFC unit operating at elevated pressure. The reference pressurized SOFC technology parameters are listed in Exhibit 5-1.

Exhibit 5-1. IGFC Parameters Used for the Pressurized Reference Case

Parameter	Value
Gasifier Technology	Conventional
Syngas CH ₄ content (mole %)	5.8
SOFC Operating Pressure, psia (atm)	44.1 (3)
Cell Technology	Advanced Cell
Fuel Utilization, %	80
Current Density, mA/cm ²	400
Degradation, %/1000 h	0.2
Inverter Efficiency (%)	97
Stack Cost (\$/kW)	225
Capacity Factor (%)	80

The reference plant with CCS's relevant process data for the numbered streams in the block flow diagram (BFD), shown in Exhibit 5-2, are tabulated in Exhibit 5-3. The syngas methane content entering the SOFC section is ≈5.8 mole percent. The BFD for the reference plant without CCS, shown in Exhibit 5-4, differs only downstream of the SOFC island — the combustor is fed by the

vitiated air instead of oxygen from ASU. Also, natural gas is injected into the combustor to boost cathode expander inlet temperatures.

Pressurized Reference Plant Performance

The performance summary listed in Exhibit 5-5 for the reference plant shows that the reference plant with CCS results in a net plant HHV efficiency of ≈ 38.6 percent, which is lower than the atmospheric reference plant efficiency. The absence of cathode gas recirculation in the pressurized case leads to a much higher process flow, leading, in turn, to a higher compression power and lower efficiency. The reference plant without CCS has an efficiency of ≈ 52.6 percent, which implies that the introduction of CCS imposes a performance penalty of ≈ 14 percent in this case. The difference is primarily due to the expander power boost using natural gas in the case without CCS, as can be seen from Exhibit 5-6, which shows that expander power accounts for more than 41 percent of the gross power generated in the case without CCS.

The air compressor load nearly doubles the auxiliary load of the pressurized plant relative to the atmospheric IGFC. The auxiliary load for the CCS plant is further burdened by the addition of the CPU, shown in Exhibit 5-7.

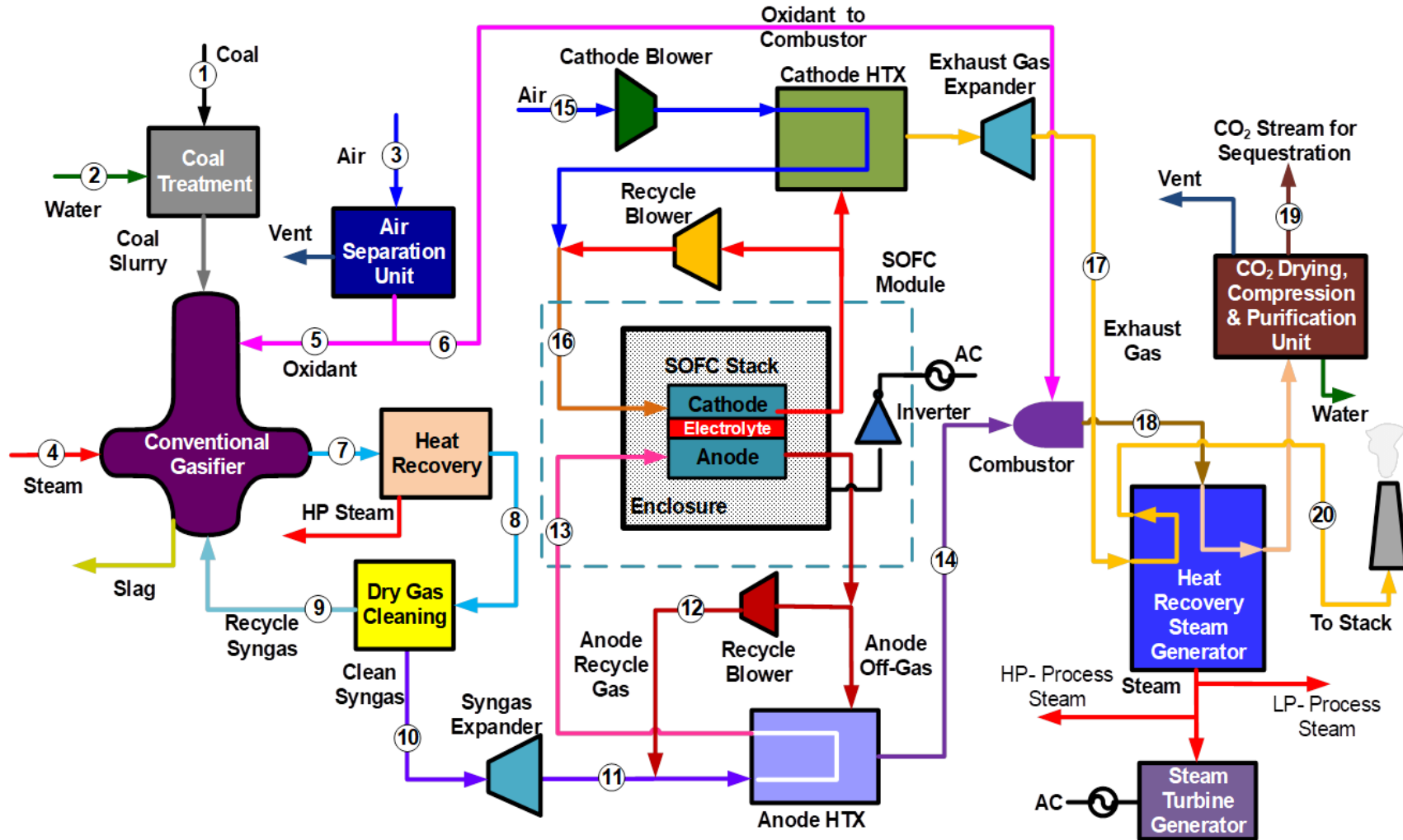
The salient material and energy balances for the reference plant with CCS is listed in Exhibit 5-8, which also shows that the IGFC plant acts as a nearly zero-emission power plant, with only a small release of CO_2 shown in the emissions. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination. The nearly complete recovery of water from the oxy-combustion CO_2 product stream results in an IGFC plant water consumption significantly lower than the corresponding value for conventional fossil fuel power plant technologies. The heat and mass balance diagram for the gasifier, ASU, and syngas clean-up is shown in Exhibit 5-9, while the corresponding PFDs for the power island and the CPU are shown in Exhibit 5-10 and Exhibit 5-11, respectively.

Pressurized Reference Plant Costs

The capital and O&M costs for the atmospheric reference plant with CCS are itemized in Exhibit 5-12 and Exhibit 5-13, respectively. The gasifier and cleanup costs account for nearly 49 percent of the capital costs as shown in Exhibit 5-14. The gasifier costs are nearly two times the costs used in the previous pathway and are based on the BB report, which considers the recent installations. Any cost reduction in the gasifier systems would greatly benefit the overall cost of the IGFC system. The cathode heat exchanger contributes to 50 percent of the BOP costs as shown in Exhibit 5-15.

The LCOE for the reference plant is dominated by the capital cost component (≈ 57 percent) as shown in Exhibit 5-16. The LCOE for the reference plant with CCS is $\$106.7/\text{MWh}$, about 23 percent higher than the LCOE for the plant without CCS, mainly due to large efficiency differences.

Exhibit 5-2. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Process Diagram



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 5-3. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
V-L Mole Fraction																				
Ar	0.0000	0.0000	0.0094	0.0000	0.0031	0.0031	0.0006	0.0006	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0094	0.0094	0.0097	0.0010	0.0000	0.0097
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0443	0.0443	0.0577	0.0583	0.0583	0.0033	0.0152	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2869	0.2869	0.3737	0.3775	0.3775	0.0913	0.1535	0.0766	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.1548	0.1548	0.2021	0.2043	0.2043	0.4824	0.4220	0.4967	0.0003	0.0003	0.0003	0.5673	1.0000	0.0003
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003	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	0.2675	0.2675	0.3486	0.3519	0.3519	0.1068	0.1600	0.0942	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0000	1.0000	0.0104	1.0000	0.0000	0.0000	0.2315	0.2315	0.0023	0.0013	0.0013	0.3101	0.2430	0.3260	0.0104	0.0104	0.0107	0.4163	0.0000	0.0107
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0010	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	0.0000	0.0000	0.0069	0.0069	0.0094	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N ₂	0.0000	0.0000	0.7722	0.0000	0.0019	0.0019	0.0020	0.0020	0.0027	0.0060	0.0060	0.0054	0.0055	0.0053	0.7722	0.7722	0.7952	0.0055	0.0000	0.7952
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0044	0.0044	0.0017	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	0.2077	0.0000	0.9950	0.9950	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.2077	0.1841	0.0099	0.0000	0.1841
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	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	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	0	5,097	34,289	4,088	4,773	1,984	28,469	28,469	3,277	18,386	15,629	56,308	71,936	20,510	244,296	244,296	237,217	20,744	11,621	237,217
V-L Flowrate (kg/hr)	0	91,826	989,376	73,654	152,800	63,527	589,544	589,544	70,549	394,048	334,941	1,679,292	2,014,224	620,550	7,048,969	7,048,969	6,822,459	684,076	511,451	6,822,459
Solids Flowrate (kg/hr)	223,530	0	0	0	0	0	208	208	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	148	15	288	130	27	999	232	78	316	118	682	649	594	15	632	132	1,436	37	156
Pressure (MPa, abs)	0.10	0.69	0.10	3.45	3.45	0.16	3.10	2.96	4.14	2.32	0.34	0.34	0.34	0.3	0.101	0.305	0.100	0.338	15.272	0.101
Enthalpy (kJ/kg) ^A	---	566.85	31.06	2,973.90	114.27	23.91	2,286.80	859.90	103.53	484.62	175.42	1,386.02	1,289.91	1,266.0	31.057	685.580	150.980	2,627.201	-207.331	170.684
Density (kg/m ³)	---	864.7	1.2	14.8	33.1	2.0	6.0	14.7	30.6	10.1	2.2	1.3	1.2	1.4	1.2	1.2	0.9	0.8	539.3	0.8
V-L Molecular Weight	---	18.015	28.854	18.015	32.016	32.016	20.708	20.708	21.525	21.431	21.431	29.823	28.000	30	28.854	28.854	28.760	32.978	44.010	28.760
V-L Flowrate (lb _{mol} /hr)	0	11,237	75,594	9,013	10,522	4,374	62,763	62,763	7,226	40,535	34,455	124,138	158,592	45,217	538,580	538,580	522,974	45,732	25,621	522,974
V-L Flowrate (lb/hr)	0	202,442	2,181,202	162,380	336,865	140,053	1,299,722	1,299,722	155,533	868,728	738,419	3,702,204	4,440,603	1,368,078	15,540,318	15,540,318	15,040,948	1,508,130	1,127,557	15,040,947
Solids Flowrate (lb/hr)	492,800	0	0	0	0	0	459	459	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	299	59	550	265	80	1,830	450	173	600	244	1,260	1,200	1,100	59	1,169	270	2,616	98	313
Pressure (psia)	14.7	100.0	14.7	500.0	500.0	23.0	450.0	430.0	600.0	337.0	49.1	49.1	49.1	49.0	14.7	44.3	14.5	49.0	2,215.0	14.7
Enthalpy (Btu/lb) ^A	---	243.7	13.4	1,278.5	49.1	10.3	983.1	369.7	44.5	208.4	75.4	595.9	554.6	544.3	13.4	294.7	64.9	1,129.5	-89.1	73.4
Density (lb/ft ³)	---	53.983	0.076	0.923	2.067	0.127	0.377	0.918	1.910	0.631	0.139	0.079	0.077	0	0.076	0.073	0.053	0.049	33.670	0.051

^A Reference conditions are 32.02 F and 0.089 psia

Exhibit 5-4. Case PCONVIGFC0A – Reference Pressurized IGFC Plant without CCS – Process Diagram

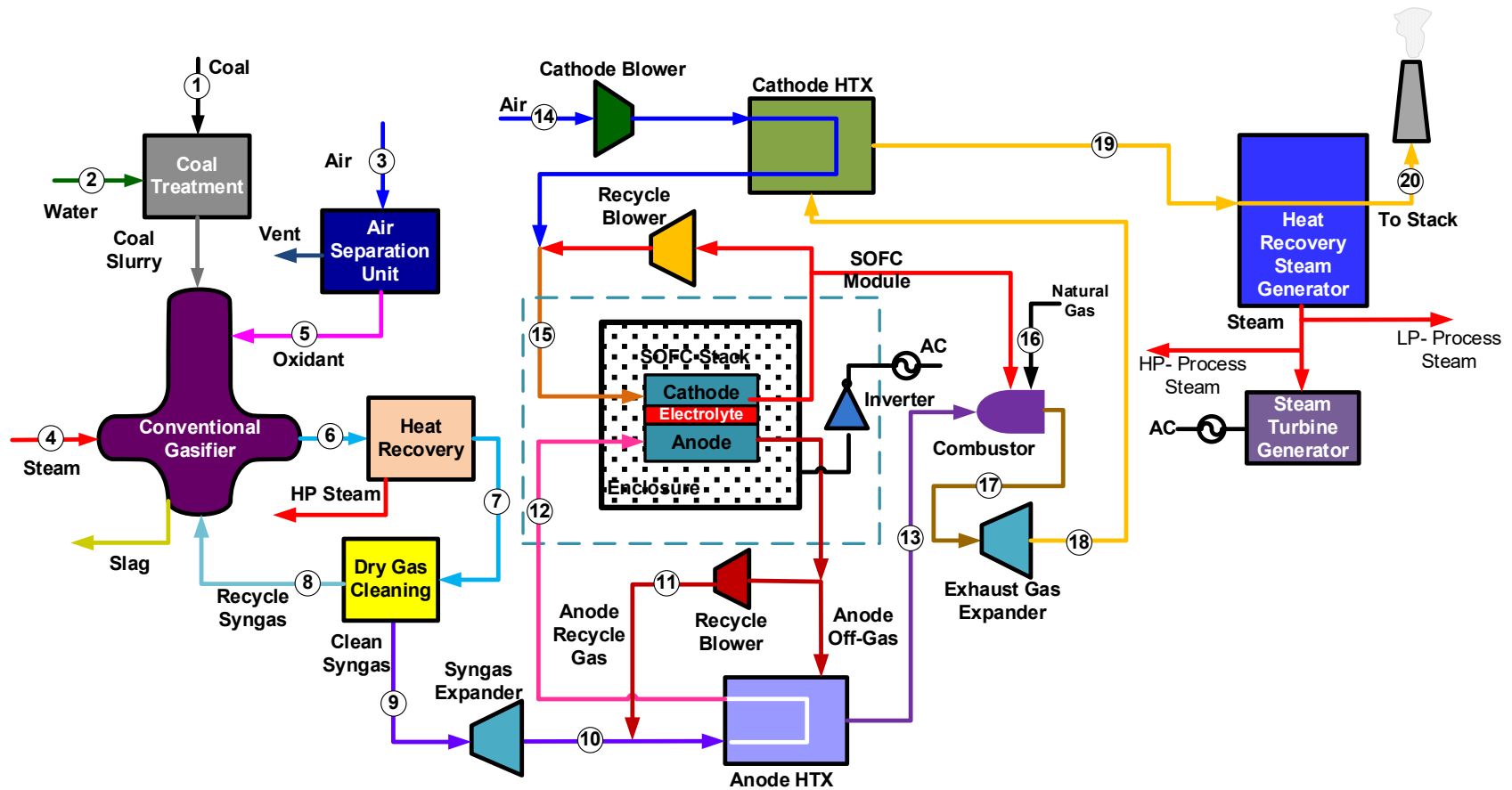


Exhibit 5-5. Reference Pressurized IGFC plant with and without CCS – Performance Summary

Carbon Capture	With CCS	No CCS
Power Summary (Gross Power at Generator Terminals, kW_e)		
SOFC Power	645,997	462,798
Natural Gas/Syngas Expander Power	220,300	372,100
Steam Turbine Power	199,100	67,600
Total Gross Power, kW_e	1,065,397	902,498
Auxiliary Load Summary, kW_e		
Coal Handling	470	400
Coal Milling	2,300	1,650
Sour Water Recycle Slurry Pump	0	0
Ash Handling	960	690
Air Separation Unit Auxiliaries	0	0
Air Separation Unit Main Air Compressor	49,451	25,296
Oxygen Compressor	30,900	22,140
Nitrogen Compressors	0	0
Claus Plant TG Recycle Compressor	1,760	1,260
CO ₂ Compressor	45,890	0
CO ₂ Refrigeration	0	0
Boiler Feedwater Pumps	2,326	790
Condensate Pump	201	68
Syngas Recycle Compressor	1,250	910
Circulating Water Pump	3,290	1,070
Ground Water Pumps	582	417
Cooling Tower Fans	1,700	560
Scrubber Pumps	319	229
Quench Water Pump	980	702
Selexol Auxiliary Power	3,513	2,517
Steam Turbine Auxiliaries	183	62
Claus Plant/TGTU Auxiliaries	255	183
Gas Turbine Auxiliaries	916	311
Cathode air blower	258,770	185,390
Cathode recycle blower	0	0
Anode recycle blower	300	210
Miscellaneous Balance of Plant ^A	4,668	3,954
Transformer Losses	4,360	3,400
Total Auxiliaries, kW_e	415,344	252,208
Net Power, kW_e	650,053	650,290
Net Plant Efficiency, % (HHV)	38.58%	52.99%
CO₂ Capture Rate, %	98.8%	0.0%
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	9,331 (8,844)	6,708 (6,358)
Coal Feed Rate, kg/hr (lb/hr)	223,530 (492,800)	160,145 (353,060)
Thermal Input^B, kWt	1,684,867	1,227,216
Condenser Cooling Duty, 10⁶ kJ/hr (10⁶ Btu/hr)	1,129 (1,070)	380 (360)
Raw Water Withdrawal, m³/min (gpm)	8.2 (2,161)	2.7 (714)
Raw Water Consumption, m³/min (gpm)	5.3 (1,400)	1.8 (467)

^A Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads.

^B HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb).

Exhibit 5-6. Reference Pressurized IGFC Plant with and without CCS – Gross Power Distribution

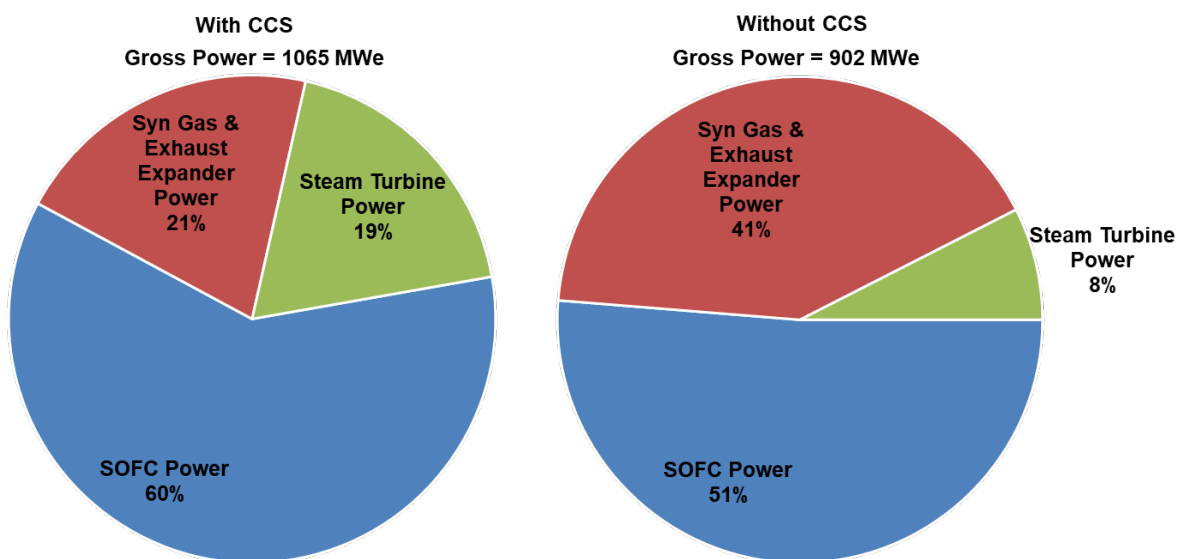


Exhibit 5-7. Reference Pressurized IGFC Plant with and without CCS – Auxiliary Load Distribution

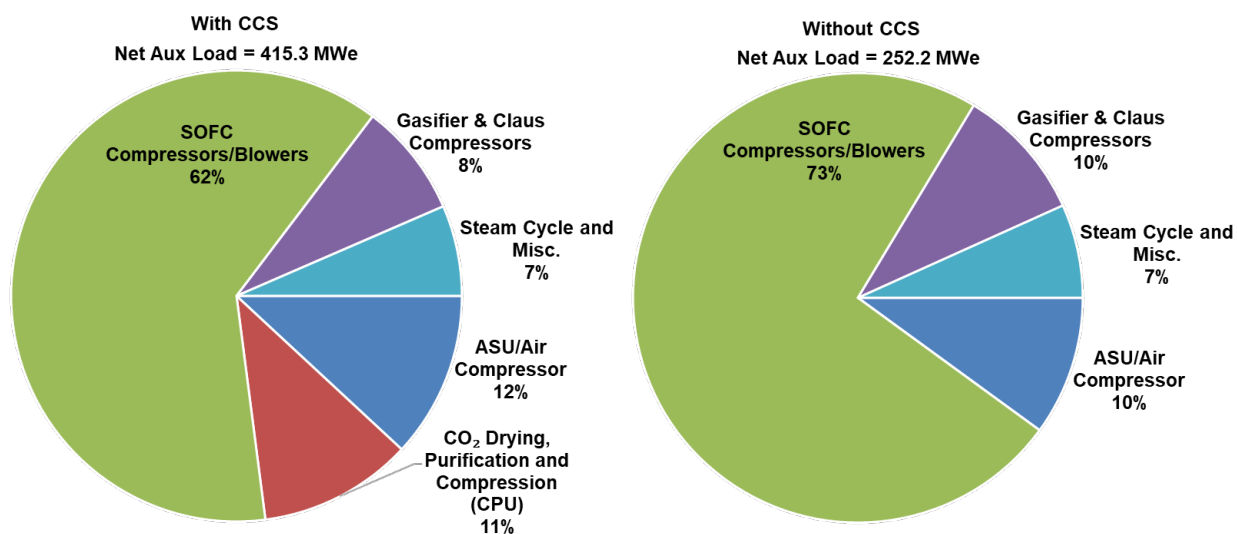


Exhibit 5-8. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Material and Energy Balances
Carbon Balance

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	142,489 (314,134)	Slag	1,140 (2,513)
Air (CO ₂)	1,004 (2,213)	Stack Gas	880 (1,941)
		CO ₂ Product	139,583 (307,728)
		N ₂ Product	124 (272)
		Exhaust	1,759 (3,878)
		Convergence Tolerance	2 (5)
Total	143,493 (316,347)	Total	143,493 (316,347)

Sulfur Balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	5,603 (12,352)	Elemental Sulfur	5,600 (12,345)
		Polishing Sorbent	3 (7)
		Convergence Tolerance	0 (0)
Total	5,603 (12,352)	Total	5,603 (12,352)

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	0.40 (106)	0.40 (106)			
Slurry Water	1.53 (405)	1.53 (405)			
Quench/Wash	0.4 (98)	0.37 (98)			
Condenser Makeup	1.2 (328)	0.0 (0)	1.2 (328)	0.0 (0)	1.2 (328)
Gasifier Steam	1.2 (325)	0.0 (0)	1.2 (325)		
BFW Makeup	0.0 (3)	0.0 (0)	0.0 (3)		
Cooling Tower	12.8 (3,383)	5.87 (1,550)	6.9 (1,833)	2.9 (761)	4.1 (1,072)
CO ₂ Dehydration	0.0 (0)	2.60 (686)	-2.60 (-686)		
SWS Excess Water	0.0 (0)	3.27 (864)	-3.27 (-864)		
Total	16.4 (4,320)	8.17 (2,159)	8.2 (2,161)	2.9 (761)	5.3 (1,400)

Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MM Btu/hr)				
Coal	6,066 (5,749)	5.1 (4.8)		6,071 (5,754)
ASU Air		30.7 (29.1)		31 (29)
Fuel cell Air		218.9 (207.5)		219 (207)
Raw Water Makeup		30.8 (29.2)		31 (29)
Auxiliary Power			1,495 (1,417)	1,495 (1,417)
TOTAL	6,066 (5,749)	285.5 (270.6)	1,495 (1,417)	7,846 (7,437)
Heat Out GJ/hr (MM Btu/hr)				
Slag	30 (29)	19.0 (18.0)		49 (47)
Sulfur	52 (49)	0.6 (0.5)		52 (50)
CO ₂		-106.0 (-100.5)		-106 (-101)
Cooling Tower Blowdown		21.4 (20.3)		21 (20)
Exhaust		1,030.1 (976.3)		1,030 (976)
HRSG Flue Gas		0.0 (0.0)		0 (0)
Condenser		1,126 (1,067)		1,126 (1,067)
Non-Condenser Cooling Tower Loads*		546 (518)		546 (518)
Process Losses*		1,279 (1,212)		1,279 (1,212)
Power			3,835 (3,635)	3,835 (3,635)
TOTAL	82 (78)	3,916 (3,724)	3,835 (3,635)	7,834 (7,437)

Emissions

	kg/GJ (lb/10 ⁶ Btu)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO ₂	0 (0)	0 (0)	0 (0)
NO _x	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO ₂	2 (4)	67,889 (74,835)	9 (20)

Exhibit 5-9. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – ASU, Gasifier, and Clean-up

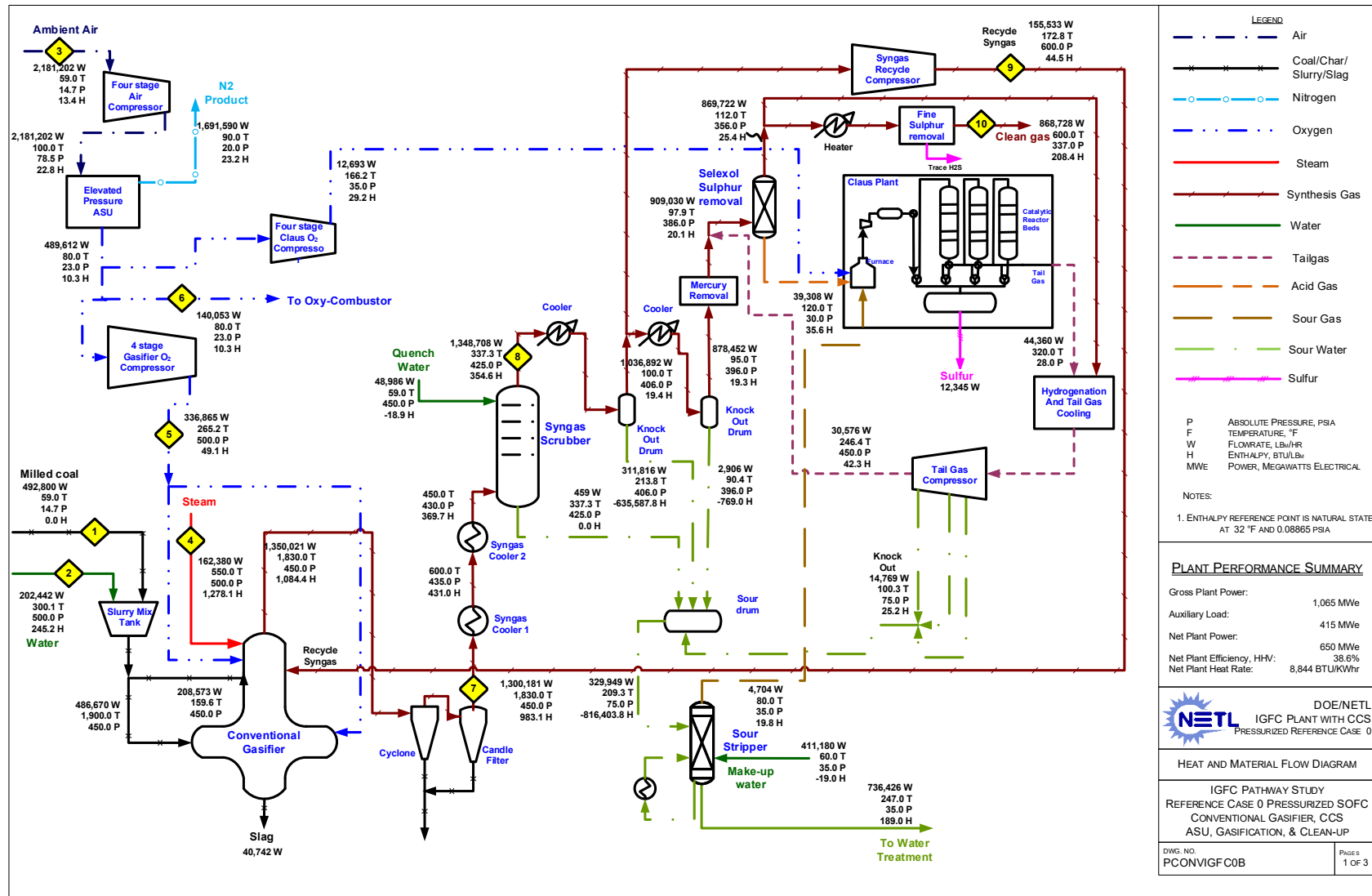


Exhibit 5-10. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – Power Block

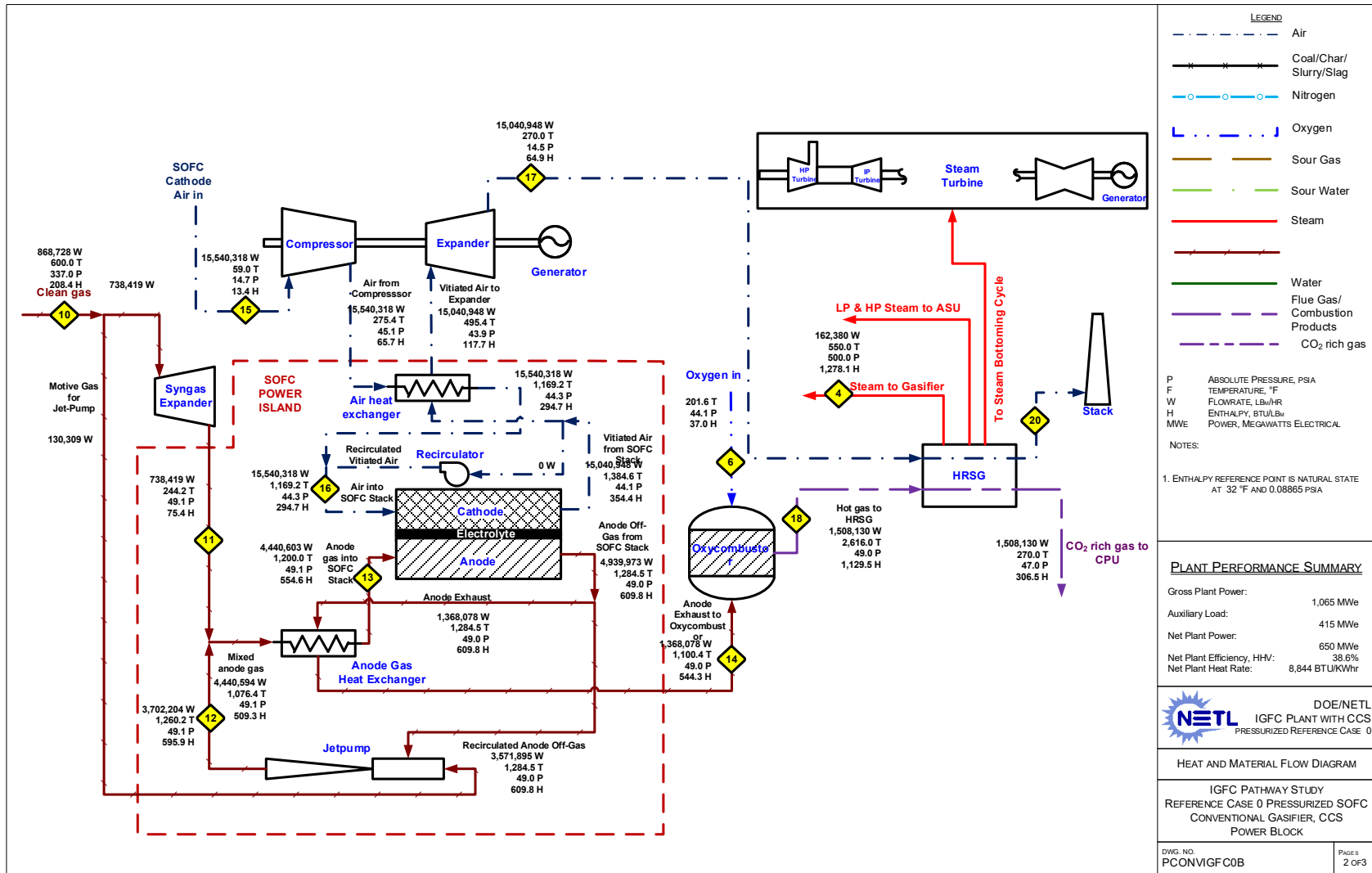


Exhibit 5-11. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – CO₂ Dehydration, Purification, and Compression

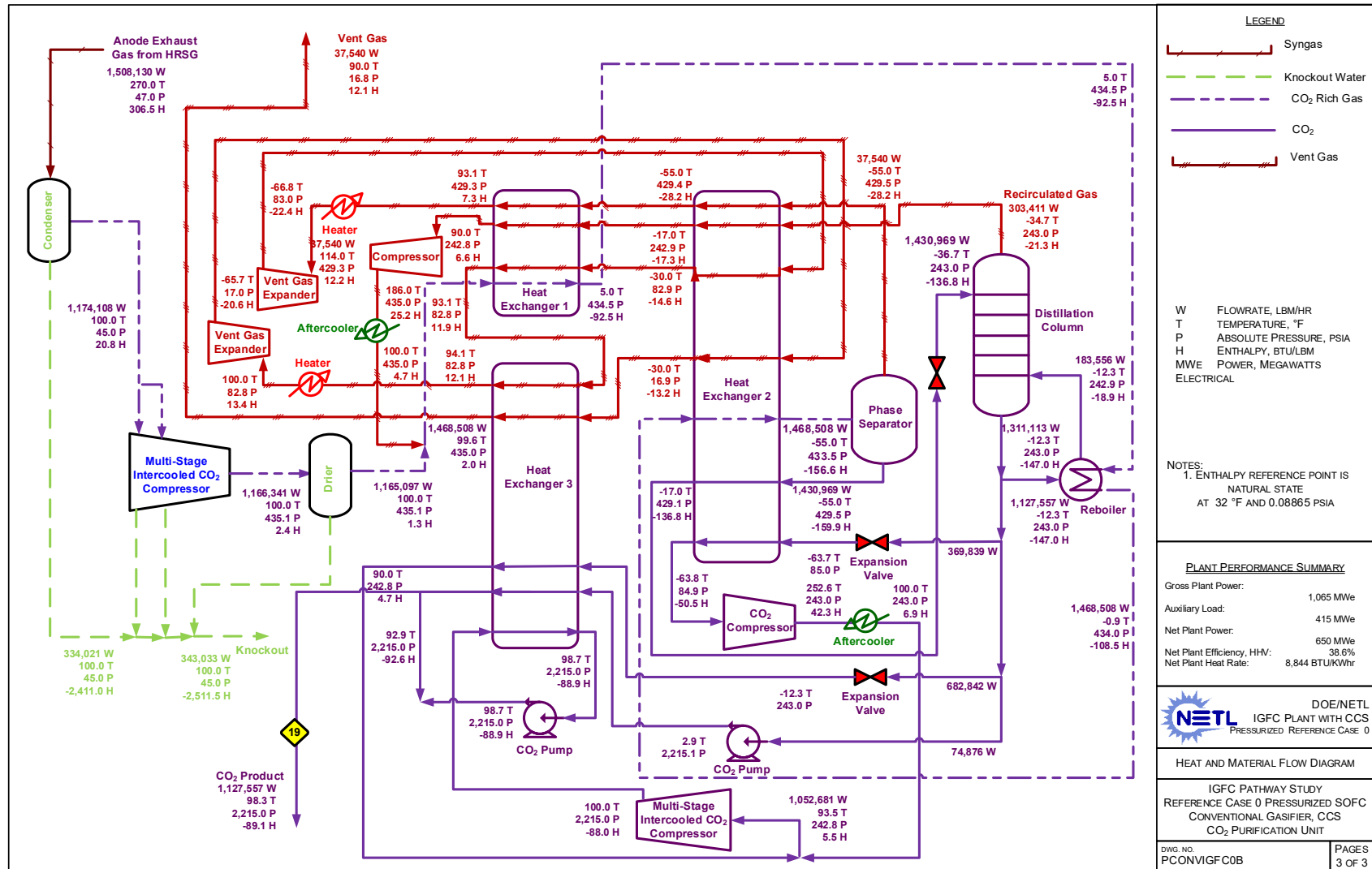


Exhibit 5-12. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – Capital Costs

Cost Component	Cost (\$1000)	Specific Cost (\$/kWe AC)
	2018\$	2018\$
SOFC Module		
SOFC Stack	149,469	230
Container	17,937	28
Insulation	2,657	4
Module Assembly	6,643	10
Air Distribution	6,643	10
Fuel Distribution	6,643	10
Module Current Collectors	3,322	5
Module I&C	3,322	5
Inverter	45,173	69
Total SOFC Module with 10 % Extra Installed Area	241,808	372
Syngas Expander	12,839	20
SOFC BOP		
Cathode Air Blower	25,169	39
Cathode Gas Recycle Blower	0	0
Cathode Heat Exchanger	95,023	146
Anode Recycle Blower	1,105	2
Anode Heat Exchanger	24,328	37
Oxy-Combustor	31,344	48
Total SOFC BOP	176,970	272
TOTAL SOFC POWER ISLAND	431,616	664
COAL & SORBENT HANDLING	70,493	108
COAL & SORBENT PREP & FEED	41,512	64
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	1,107,801	1704
ASU & oxidant compressor	102,887	158
Other gasification equip & foundations	94,506	145
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	1,305,194	2008
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	0	0
Single-Stage Selexol/MDEA	15,250	23
Claus Plant	168,734	260
Trace removal	6,429	10
COS Hydrolysis	41,664	64
Blowback, Piping, Foundations	9,913	15
Sulfur polishing/NG desulfurization	13,209	20
TOTAL GAS CLEANUP & PIPING	255,199	393
STEAM CYCLE		
HRS&G, Ducting, and Stack	84,535	130
Steam Power System	67,289	104
Feedwater and Misc BOP systems	87,203	134
TOTAL STEAM CYCLE	239,027	368
CO₂ COMPRESSION & PURIFICATION	147,982	228
COOLING WATER SYSTEM	46,258	71
ASH/SPENT SORBENT HANDLING SYS	11,379	18
ACCESSORY ELECTRIC PLANT	96,870	149
INSTRUMENTATION & CONTROL	31,592	49
IMPROVEMENTS TO SITE	30,886	48
BUILDING & STRUCTURES	17,504	27
TOTAL PLANT COST (TPC)	2,725,410	4193
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	9,144	
1 Month Maintenance Materials	1,812	
1 Month Non-fuel Consumables	1,215	
1 Month Waste Disposal	283	
25% of 1 Months Fuel Cost at 100% CF	2,337	
2% of TPC	54,886	
Total Preproduction Costs	69,676	
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	20,935	
0.5% of TPC (spare parts)	13,722	
Total Inventory Capital	34,657	
Initial Cost for Catalyst and Chemicals	61,679	
Land	900	
Other Owner's Costs	411,645	
Financing Costs	74,096	
TOTAL OWNER'S COSTS	652,653	
TOTAL OVERNIGHT COST (TOC)	3,378,063	5197
TASC Multiplier	1.154	
TOTAL AS-SPENT COST (TASC)	3,898,284	5997

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 5-13. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – O&M Costs

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
OPERATING & MAINTENANCE LABOR		
Operating Labor Rate (base)	39.70	
Stack replacement O&M, \$/hr per stack kW	18.33	
Annual Operating Labor Cost	5,261,256	
Maintenance Labor Cost	9,368,644	
Administrative & Support Labor	3,657,475	
Property Taxes and Insurance	54,886,016	
TOTAL FIXED OPERATING COSTS	73,173,390	16.06
VARIABLE OPERATING COSTS		
Maintenance Material Cost	17,398,910	
Stack replacement		
SOFC stack replacement cost, \$/kW AC	222	
SOFC Stack replacement O&M, \$/yr per kW	22	
Stack Replacement Cost	14,399,134	3.16
CONSUMABLES		
Water		
Price (\$/1000 gal)	1.900	
Annual (\$)	894,144	
Chemicals		
MU & WT Chem.		
Price (\$/lb)	0.28	
Annual (\$)	744,358	
Carbon (Trace Removal)		
Price (\$/lb)	1.6	
Initial Cost (\$)	963,951	
Annual (\$)	385,581	
COS Catalyst		
Price (\$/m³)	45,909	
Initial Cost (\$)	35,695,510	
Annual (\$)	7,139,102	
Selexol Solution		
Initial Fill (gal)		
Price (\$/gal)	38.00	
Initial Cost (\$)	21,393,808	
Annual (\$)	619,801	
Claus / DSRP Catalyst		
Price (\$/ft³)	48	
Annual (\$)	28,625	
ZnO polishing sorbent		
Price (\$/lb)	1.5	
Initial Cost (\$)	883,886	
Annual (\$)	990,856	
COS Catalyst		
Price (\$/ft³)	772	
Initial Cost (\$)	2,742,155	
Annual (\$)	856,923	
Sub Total Chemicals		

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
Initial Fill (\$)	61,679,311	
Annual (\$)	10,765,246	
Waste Disposal		
Spent Trace Catalyst		
Price (\$/lb)	0.42	
Annual (\$)	107,300	
Ash + HCl Sorbent		
Price (\$/ton)	16.31	
Annual (\$)	2,328,475	
Spent sorbents		
Price (\$/lb)	0.42	
Annual (\$)	277,440	
Subtotal-Waste Disposal (\$)	2,713,215	
TOTAL VARIABLE OPERATING COSTS		
Initial Fill (\$)	61,679,311	
Annual (\$)	46,170,649	10.13
Fuel Coal		
Price (\$/ton)	51.96	
Annual (\$)	89,723,032	19.70

Exhibit 5-14. Case PCONVIGFCOB – Reference Pressurized IGFC Plant with CCS – Capital Cost Components

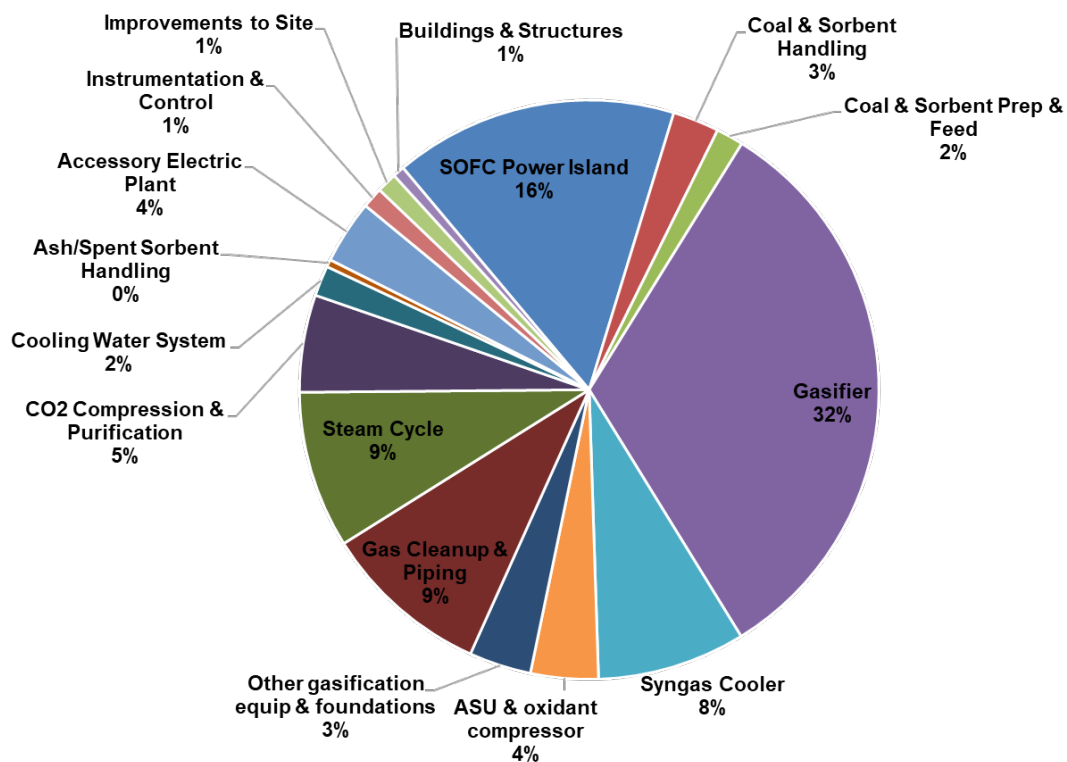


Exhibit 5-15. Case PCONVIGFC0B – Reference Pressurized IGFC Plant with CCS – BOP Cost Distribution

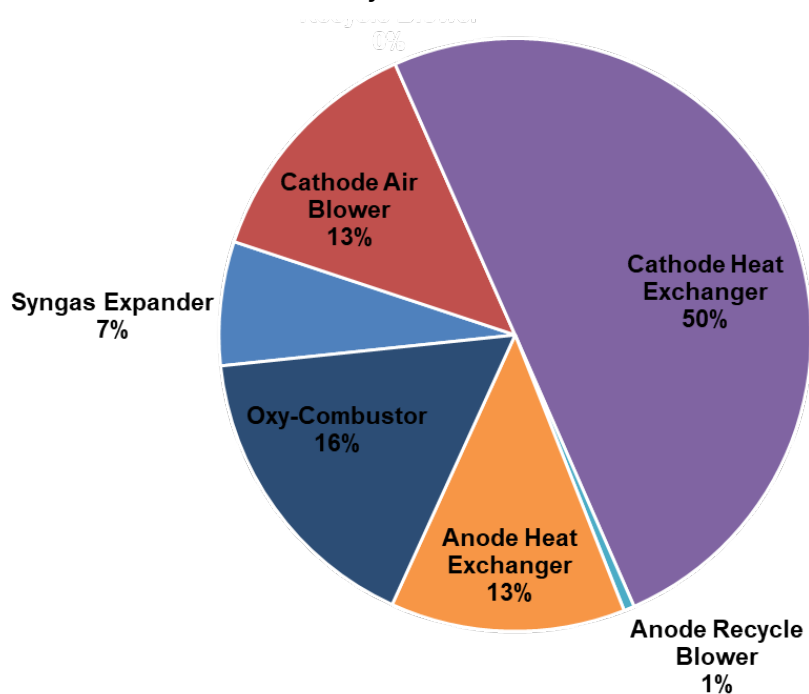


Exhibit 5-16. Case 0 – Reference pressurized IGFC plant with and without CCS – levelized cost of electricity

Levelized Cost of Electricity (2018\$/MWh)		
Carbon Capture	With CCS	No CCS
Component	\$/MWh	
Variable Costs	29.8	21.6
<i>Fuel Costs</i>	19.7	14.6
<i>Variable O&M Costs</i>	10.1	7.3
Fixed O&M Costs	16.1	13.9
Capital Costs	60.8	51.2
Total LCOE (excluding T&S)	106.7	87.0
T&S	7.9	0.0
Total LCOE (including T&S)	114.6	87.0

5.2 CASE 6: ADVANCED GASIFIER PRESSURIZED IGFC

Case 6 represents the culmination of the pressurized pathway for IGFC plants with the utilization of an advanced catalytic gasifier, resulting in a CH₄ content of ≈31.9 percent in the syngas entering the SOFC section. The SOFC technology parameters are listed in Exhibit 5-17 and include BOP enhancements as well as a reduction in stack cost.

Exhibit 5-17. IGFC Parameters Used for the Advanced Gasifier Case 6 BV

Parameter	Value
Gasifier Technology	Catalytic
Syngas CH ₄ content (mole %)	31.9
SOFC Operating Pressure, psia (atm)	44.1 (3)
Cell Technology	Advanced Cell
Fuel Utilization, %	85
Current Density, mA/cm ²	400
Degradation, %/1000 h	0.2
Inverter Efficiency (%)	98
Stack Cost (\$/kW)	200
Capacity Factor (%)	85

The relevant process data for the numbered streams in the block flow diagram (BFD) for the Case 6 plant with CCS shown in Exhibit 5-18 are tabulated in Exhibit 5-19. The higher syngas CH₄ content reduces the SOFC process air flow requirement by over 50 percent relative to the enhanced gasifier case.

Advanced Pressurized IGFC Performance

As shown in Exhibit 5-20, the Case 6 plant results in an efficiency of 57.8 percent, ≈19 percentage points higher than that of the reference pressurized plant, and 10–25 percentage points higher than conventional technologies, mainly due to the combined result of higher syngas CH₄ content. The Case 6 IGFC plant without CCS results in the highest efficiency of 68.3 percent amongst all the pathway cases considered in this report due to the higher proportion of syngas power generation (Exhibit 5-21). The compressor accounts for majority of the auxiliary load in both plants with and without CCS, as seen in Exhibit 5-22.

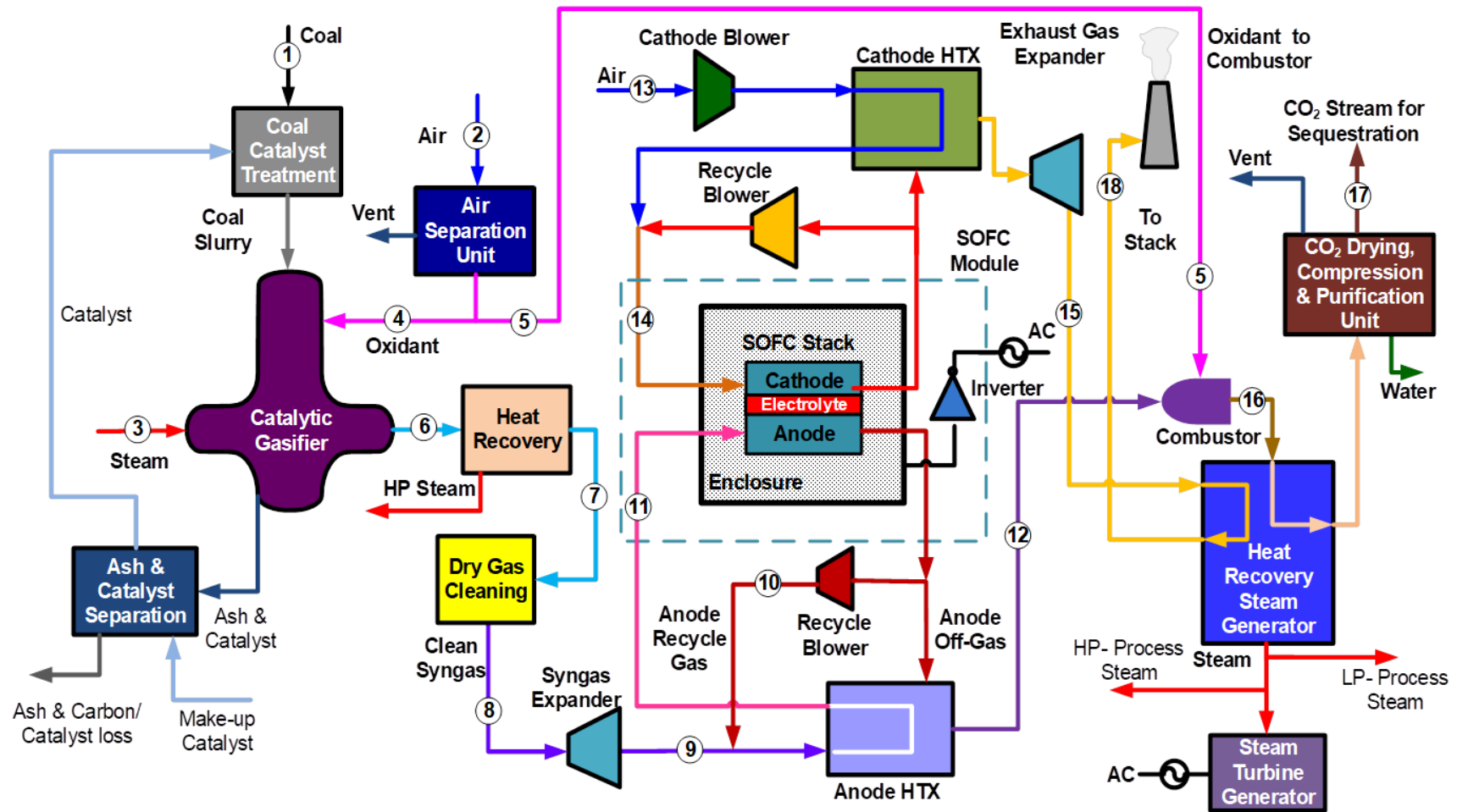
The salient material and energy balances for the plant with CCS are listed in Exhibit 5-23, which shows that the emission of the IGFC plant even without CCS is much lower than those of other conventional coal-based technologies. The IGFC plant water consumption is also significantly lower than the corresponding value for conventional fossil fuel power plant technologies. The heat and mass balance diagrams for the gasifier, ASU, and syngas clean-up, as well as the power island and CPU, are shown in Exhibit 5-24, Exhibit 5-25, and Exhibit 5-26, respectively.

Advanced Pressurized IGFC Costs

The capital and O&M costs for the Case 6 IGFC plant with CCS are itemized in Exhibit 5-27 and Exhibit 5-28, respectively. The lower cost of the catalytic gasifier, along with the highly efficient system, reduce the gasifier and clean-up cost contributions to 43 percent of the capital costs, as shown in Exhibit 5-29. The BOP cost distribution is shown in Exhibit 5-30. The corresponding cost distribution for the plants without CCS are similar and are shown in Exhibit 5-31 and Exhibit 5-32.

The LCOE for the Case 6 plant with CCS is \$62.4/MWh, as shown in Exhibit 5-33. The LCOE of the plant without CCS is \$55.4/MWh, which presents the lowest cost of electricity amongst all the IGFC plants considered in this report.

Exhibit 5-18. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Process Diagram



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 5-19. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
V-L Mole Fraction																		
Ar	0.0000	0.0094	0.0000	0.0001	0.0031	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0094	0.0094	0.0104	0.0003	0.0000	0.0104
CH ₄	0.0000	0.0000	0.0000	0.0000	0.0000	0.1974	0.1974	0.3190	0.3190	0.0000	0.0952	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0565	0.0565	0.0914	0.0914	0.0694	0.0759	0.0694	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO ₂	0.0000	0.0003	0.0000	0.0000	0.0000	0.2120	0.2120	0.3441	0.3441	0.3913	0.3772	0.3913	0.0003	0.0003	0.0003	0.4558	1.0000	0.0003
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	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	0.1475	0.1475	0.2384	0.2384	0.0777	0.1256	0.0777	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H ₂ O	0.0000	0.0104	1.0000	0.0000	0.0000	0.3742	0.3742	0.0004	0.0004	0.4576	0.3211	0.4576	0.0104	0.0104	0.0115	0.5297	0.0000	0.0115
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	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	0.0000	0.0071	0.0071	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N ₂	0.0000	0.7722	0.0000	0.0000	0.0019	0.0038	0.0038	0.0067	0.0067	0.0041	0.0049	0.0041	0.7722	0.7722	0.8553	0.0042	0.0000	0.8553
NH ₃	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007	0.0007	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	0.2077	0.0000	0.9999	0.9950	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.2077	0.1224	0.0100	0.0000	0.1224
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	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	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg _{mol} /hr)	0	4,464	11,194	776	1,372	16,136	16,136	8,471	8,471	23,417	33,383	16,324	69,975	69,975	63,174	16,496	7,416	63,174
V-L Flowrate (kg/hr)	0	128,810	201,658	24,816	43,923	347,100	347,100	199,050	199,050	648,134	882,313	451,805	2,019,085	2,019,085	1,801,459	495,728	326,377	1,801,459
Solids Flowrate (kg/hr)	149,130	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	15	291	157	27	704	204	371	114	726	649	569	15	569	116	1,321	37	132
Pressure (MPa, abs)	0.10	0.10	7.58	7.58	0.16	6.72	6.62	5.61	0.34	0.34	0.34	0.34	0.10	0.1	0.103	0.338	15.272	0.101
Enthalpy (kJ/kg) ^A	---	31.06	2,860.25	135.67	23.91	2,108.19	908.36	608.96	169.23	1,790.68	1,536.93	1,542.19	31.06	615.5	137.025	2,792.304	-207.541	153.564
Density (kg/m ³)	---	1.2	38.4	67.9	2.0	17.7	42.0	24.3	2.5	1.1	1.2	1.3	1.2	0.4	0.9	0.8	540.0	0.9
V-L Molecular Weight	---	28.854	18.015	31.999	32.016	21.512	21.512	23.498	23.498	27.678	26.430	27.678	28.854	29	28.516	30.052	44.010	28.516
V-L Flowrate (lb _{mol} /hr)	0	9,842	24,678	1,710	3,025	35,573	35,573	18,675	18,675	51,626	73,597	35,988	154,269	154,269	139,275	36,366	16,350	139,275
V-L Flowrate (lb/hr)	0	283,977	444,580	54,709	96,834	765,224	765,224	438,829	438,829	1,428,892	1,945,167	996,059	4,451,320	4,451,320	3,971,537	1,092,892	719,539	3,971,537
Solids Flowrate (lb/hr)	328,775	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	59	555	314	80	1,300	400	700	238	1,339	1,200	1,057	59	1,057	241	2,410	98	270
Pressure (psia)	14.7	14.7	1,100.0	1,100.0	23.0	975.0	960.0	814.0	49.1	49.1	49.1	49.0	14.7	15.8	14.9	49.0	2,215.0	14.7
Enthalpy (Btu/lb) ^A	---	13.4	1,229.7	58.3	10.3	906.4	390.5	261.8	72.8	769.9	660.8	663.0	13.4	264.6	58.9	1,200.5	-89.2	66.0
Density (lb/ft ³)	---	0.076	2.398	4.240	0.127	1.102	2.623	1.518	0.154	0.070	0.073	0.083	0.076	0	0.056	0.048	33.714	0.054

^A Reference conditions are 32.02 F and 0.089 psia

Exhibit 5-20. Advanced Gasifier Pressurized IGFC Plant with and without CCS – Performance Summary

Carbon Capture	With CCS	No CCS
Power Summary (Gross Power at Generator Terminals, kW_e)		
SOFC Power	631,000	534,500
Natural Gas/Syngas Expander Power	82,000	177,200
Steam Turbine Power	67,800	25,000
Total Gross Power, kW_e	780,800	736,700
Auxiliary Load Summary, kW_e		
Coal Handling	390	370
Coal Milling	480	406
Sour Water Recycle Slurry Pump	1,864	1,578
Ash Handling	1,026	869
Air Separation Unit Auxiliaries	136	115
Air Separation Unit Main Air Compressor	1,090	930
Oxygen Compressor	0	0
Nitrogen Compressors	5,151	4,402
Claus Plant TG Recycle Compressor	3,270	5,540
CO ₂ Compressor	0	0
CO ₂ Refrigeration	1,260	1,070
Boiler Feedwater Pumps	29,600	0
Condensate Pump	0	0
Syngas Recycle Compressor	1,177	434
Circulating Water Pump	88	32
Ground Water Pumps	1,280	-160
Cooling Tower Fans	353	299
Scrubber Pumps	670	-80
Quench Water Pump	226	191
Selexol Auxiliary Power	2,088	1,768
Steam Turbine Auxiliaries	28	10
Claus Plant/TGTU Auxiliaries	170	144
Gas Turbine Auxiliaries	160	135
Cathode air blower	74,120	62,780
Cathode recycle blower	0	0
Anode recycle blower	150	130
Miscellaneous Balance of Plant ^A	3,329	3,141
Transformer Losses	2,700	2,440
Total Auxiliaries, kW_e	130,803	86,546
Net Power, kW_e	649,997	650,154
Net Plant Efficiency, % (HHV)	57.8%	68.3%
CO₂ Capture Rate, %	98.6%	0.0%
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	6,226 (5,901)	5,272 (4,997)
Coal Feed Rate, kg/hr (lb/hr)	149,130 (328,775)	126,312 (278,470)
Thermal Input^B, kWt	1,124,071	952,080
Condenser Cooling Duty, 10⁶ kJ/hr (10⁶ Btu/hr)	401 (380)	148 (140)
Raw Water Withdrawal, m³/min (gpm)	4.5 (1,176)	1.1 (299)
Raw Water Consumption, m³/min (gpm)	3.3 (879)	1.3 (335)

^A Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads.

^B HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb).

Exhibit 5-21. Advanced Gasifier Pressurized IGFC Plant with and without CCS – Gross Power Distribution

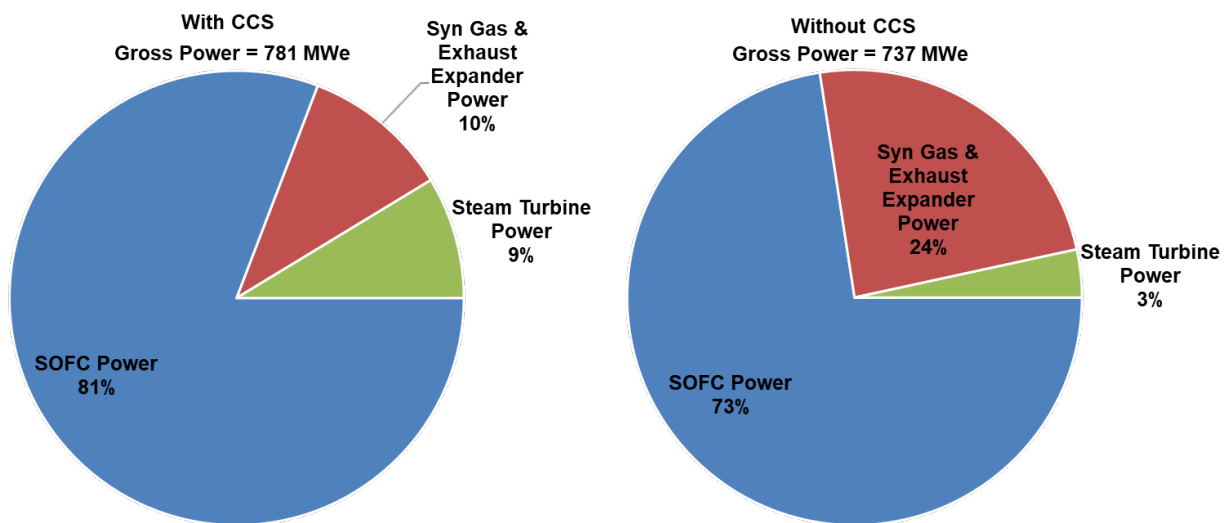


Exhibit 5-22. Advanced Gasifier Pressurized IGFC Plant with and without CCS – Auxiliary Load Distribution

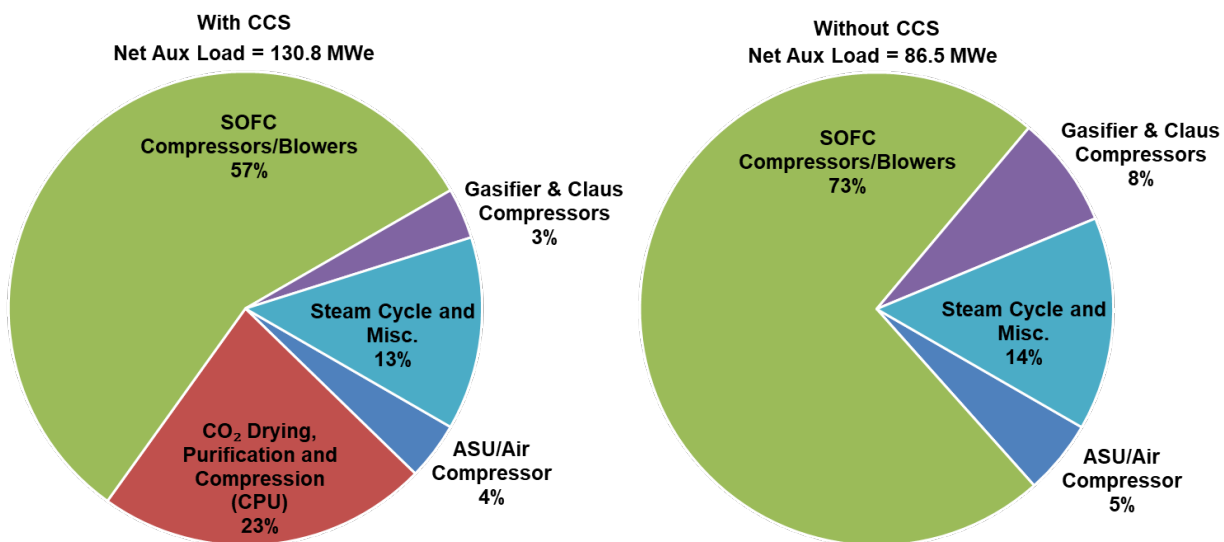


Exhibit 5-23. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Material and Energy Balances

Carbon Balance

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	95,062 (209,577)	Slag	4,924 (10,856)
Air (CO ₂)	268 (591)	Stack Gas	252 (556)
Cat Carbonator	169 (374)	CO ₂ Product	89,073 (196,373)
		N ₂ Product	16 (35)
		Carbonator Vent	3 (7)
		Vent Gas	1,234 (2,721)
		Convergence Tolerance	-3 (-8)
Total	95,500 (210,542)	Total	95,500 (210,542)

Sulfur Balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	3,738 (8,240)	Elemental Sulfur	3,736 (8,237)
		Polishing Sorbent	2 (4)
		Convergence Tolerance	0 (0)
Total	3,738 (8,240)	Total	3,738 (8,240)

Water Balance

Water Use	Water Demand m ³ /min (gpm)	Internal Recycle m ³ /min (gpm)	Raw Water Withdrawal m ³ /min (gpm)	Process Water Discharge m ³ /min (gpm)	Raw Water Consumption m ³ /min (gpm)
Slag Handling	0.46 (121)	0.46 (121)			
Slurry Water	0.05 (14)	0.05 (14)			
Quench/Wash	2.9 (765)	2.90 (765)			
Condenser Makeup	3.4 (910)	0.0 (0)	3.4 (910)	0.0 (0)	3.4 (910)
Gasifier Steam	3.4 (889)	0.0 (0)	3.4 (889)		
BFW Makeup	0.1 (21)	0.0 (0)	0.1 (21)		
Cooling Tower	5.0 (1,318)	3.98 (1052)	1.0 (266)	1.1 (296)	-0.1 (-31)
CO ₂ Dehydration	0.0 (0)	2.63 (694)	-2.63 (-694)		
SWS Excess Water	0.0 (0)	1.35 (358)	-1.35 (-358)		
Total	11.8 (3,128)	7.39 (1952)	4.5 (1,176)	1.1 (296)	3.3 (879)

Energy Balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MM Btu/hr)				
Coal	4,047 (3,835)	3.4 (3.2)		4,050 (3,839)
ASU Air		4.0 (3.8)		4 (4)
Fuel cell Air		62.7 (59.4)		63 (59)
Raw Water Makeup		16.7 (15.9)		17 (16)
Auxiliary Power			471 (446)	471 (446)
TOTAL	4,047 (3,835)	86.8 (82.3)	471 (446)	4,604 (4,364)
Heat Out GJ/hr (MM Btu/hr)				
Slag	161 (153)	15.7 (14.9)		177 (168)
Sulfur	35 (33)	0.4 (0.4)		35 (33)
CO ₂		-67.7 (-64.2)		-68 (-64)
Cooling Tower Blowdown		8.3 (7.9)		8 (8)
Exhaust		278.3 (263.8)		278 (264)
Condenser		396 (376)		396 (376)
Non-Condenser Cooling Tower Loads*		252 (239)		252 (239)
Process Losses		648 (614)		648 (614)
Difference**		66 (62)		66 (62)
Power			2,811 (2,664)	2,811 (2,664)
TOTAL	196 (186)	1,597 (1,514)	2,811 (2,664)	4,604 (4,364)

Emissions

	kg/GJ (lb/10 ⁶ Btu)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO ₂	0 (0)	0 (0)	0 (0)
NO _x	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO ₂	1 (3)	40,674 (44,835)	7 (15)

Exhibit 5-24. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – ASU, Gasifier, and Clean-up

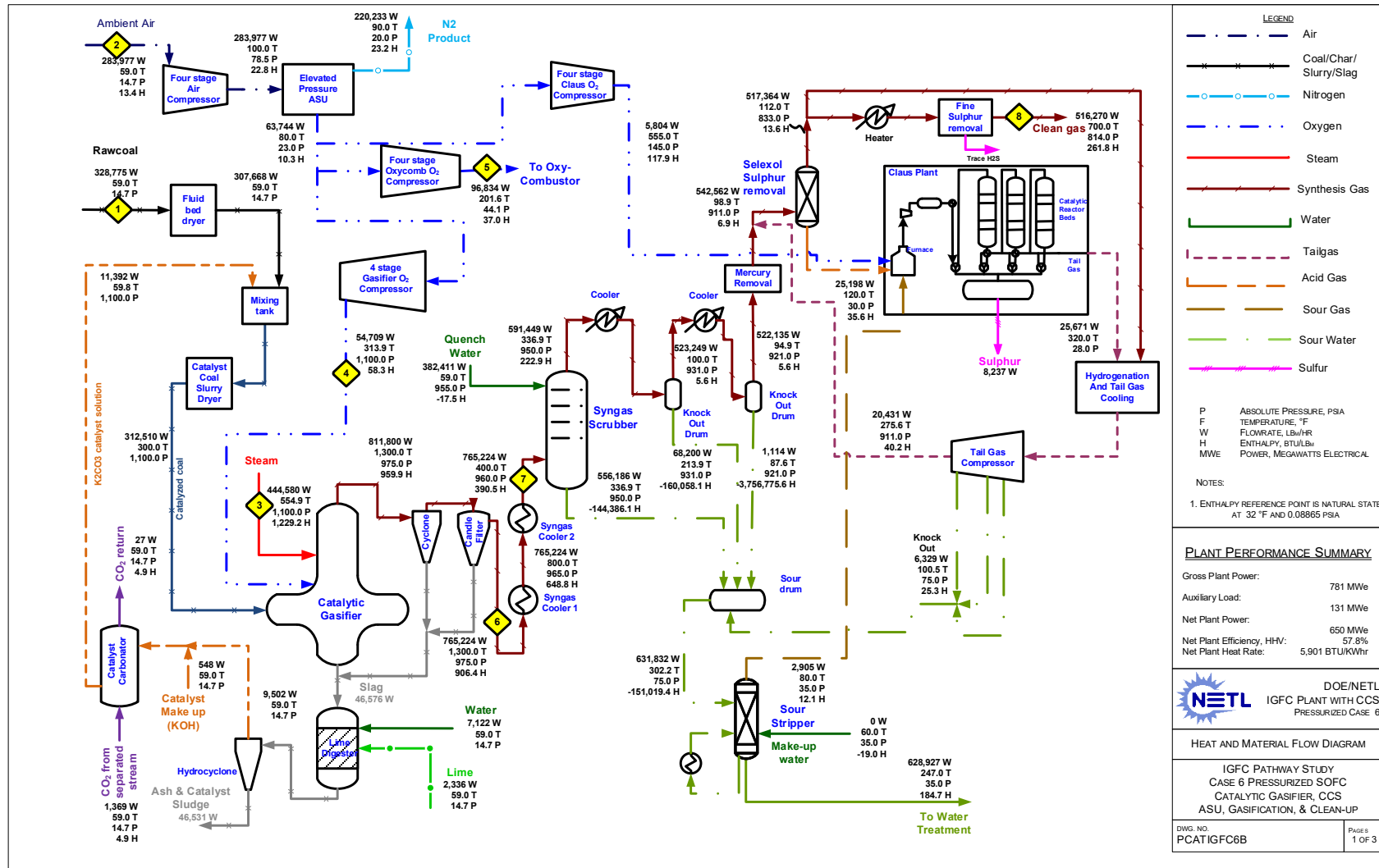


Exhibit 5-25. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – Power Block

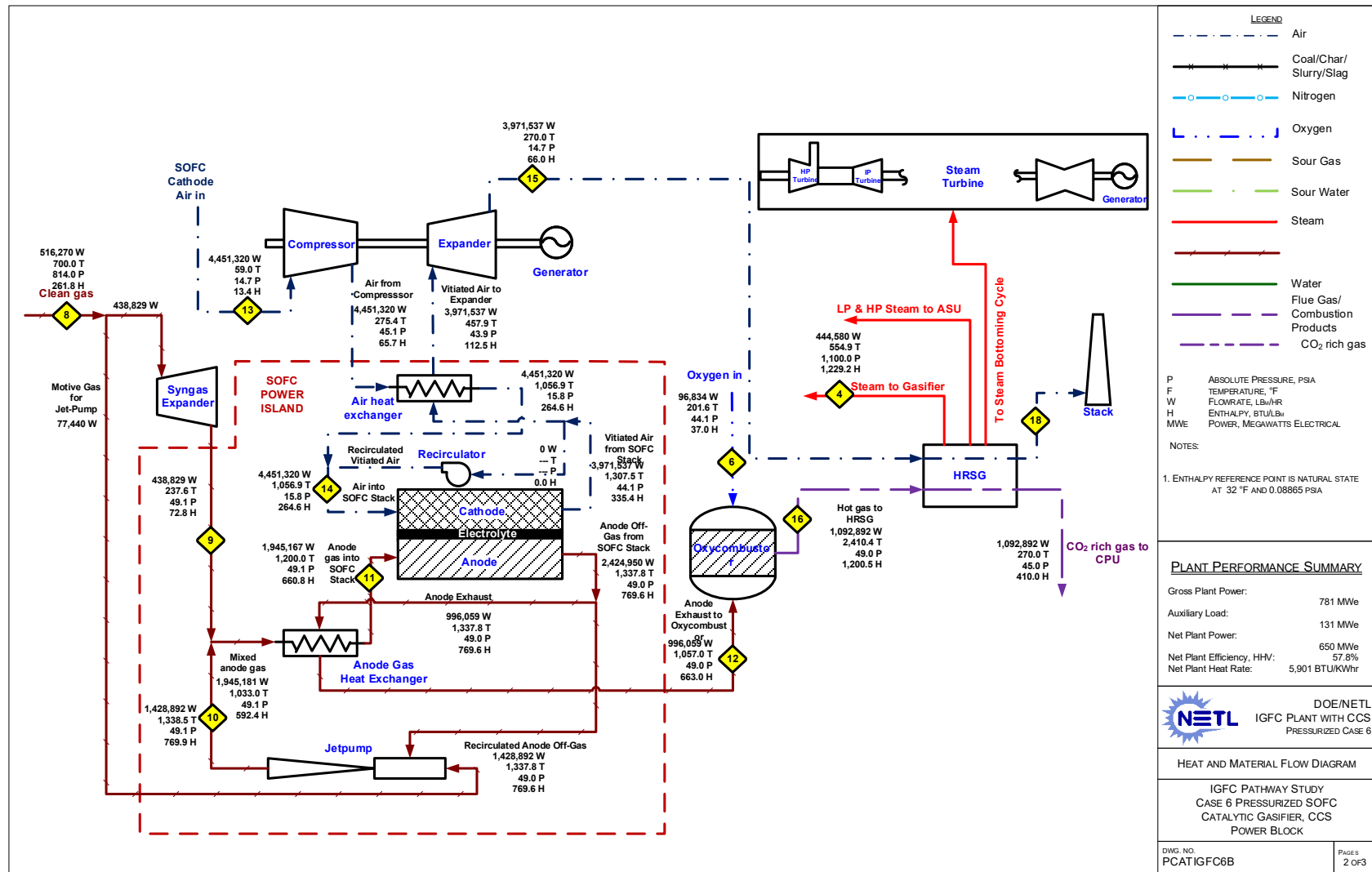
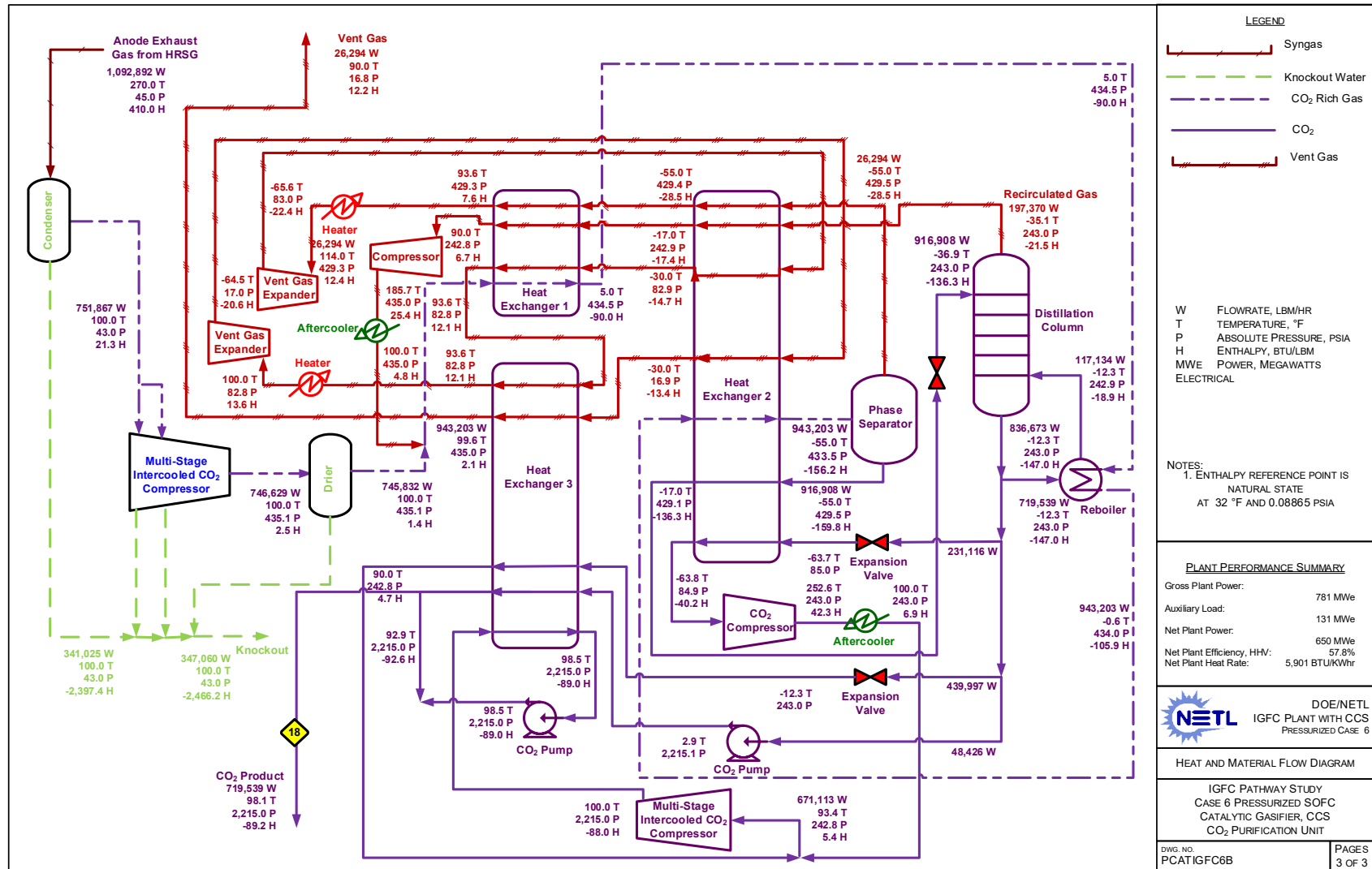


Exhibit 5-26. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Heat and Material Balance Diagram – CO₂ Dehydration, Purification, and Compression



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 5-27. Case PCATIGFC6BV – Advanced Gasifier Pressurized IGFC Plant with CCS – Capital Costs

Cost Component	Cost (\$1000)	Specific Cost (\$/kWe AC)
	2018\$	2018\$
SOFC Module		
SOFC Stack	125,868	194
Container	16,993	26
Insulation	2,517	4
Module Assembly	6,293	10
Air Distribution	6,293	10
Fuel Distribution	6,293	10
Module Current Collectors	3,147	5
Module I&C	3,147	5
Inverter	42,795	66
Total SOFC Module with 10 % Extra Installed Area	213,347	328
Syngas Expander	8,123	12
SOFC BOP		
Cathode Air Blower	10,490	16
Cathode Gas Recycle Blower	0	0
Cathode Heat Exchanger	63,844	98
Anode Recycle Blower	771	1
Anode Heat Exchanger	11,901	18
Oxy-Combustor	22,910	35
Total SOFC BOP	109,916	169
TOTAL SOFC POWER ISLAND	331,386	510
COAL & SORBENT HANDLING	54,848	84
COAL & SORBENT PREP & FEED	46,393	71
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	528,923	814
ASU & oxidant compressor	23,268	36
Other gasification equip & foundations	92,114	142
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	644,304	991
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	0	0
Single-Stage Selexol/MDEA	5,195	8
Claus Plant	128,658	198
Trace removal	3,310	5
COS Hydrolysis	15,226	23
Blowback, Piping, Foundations	6,662	10
Sulfur polishing/NG desulfurization	4,754	7
TOTAL GAS CLEANUP & PIPING	163,806	252
STEAM CYCLE		
HRSF, Ducting, and Stack	11,661	18
Steam Power System	41,492	64
Feedwater and Miscellaneous BOP systems	66,418	102
TOTAL STEAM CYLCE	119,570	184
CO₂ COMPRESSION & PURIFICATION	50,319	77
COOLING WATER SYSTEM	24,415	38
ASH/SPENT SORBENT HANDLING SYS	31,610	49
ACCESSORY ELECTRIC PLANT	61,105	94
INSTRUMENTATION & CONTROL	27,186	42
IMPROVEMENTS TO SITE	30,421	47
BUILDING & STRUCTURES	15,771	24
TOTAL PLANT COST (TPC)	1,600,867	2463
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	6,713	
1 Month Maintenance Materials	998	
1 Month Non-fuel Consumables	782	
1 Month Waste Disposal	44	
25% of 1 Months Fuel Cost at 100% CF	1,559	
2% of TPC	32,099	
Total Preproduction Costs	42,195	
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	13,911	
0.5% of TPC (spare parts)	8,025	
Total Inventory Capital	21,936	
Initial Cost for Catalyst and Chemicals	27,395	
Land	900	
Other Owner's Costs	240,745	
Financing Costs	43,334	
TOTAL OWNER'S COSTS	376,505	
TOTAL OVERNIGHT COST (TOC)	1,977,372	3042
TASC Multiplier	1.154	
TOTAL AS-SPENT COST (TASC)	2,281,888	3511

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 5-28. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – O&M Costs

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
OPERATING & MAINTENANCE LABOR		
Operating Labor Rate (base)	39.70	
Stack replacement O&M, \$/hr per stack kW	18.33	
Annual Operating Labor Cost	5,261,256	
Maintenance Labor Cost	5,479,120	
Administrative & Support Labor	2,685,094	
Property Taxes and Insurance	32,099,318	
TOTAL FIXED OPERATING COSTS	45,524,788	9.41
VARIABLE OPERATING COSTS		
Maintenance Material Cost	10,175,509	
Stack replacement		
SOFC stack replacement cost, \$/kW AC	198	
SOFC Stack replacement O&M, \$/yr per kW	20	
Stack Replacement Cost	12,841,335	2.65
CONSUMABLES		
Water		
Price (\$/1000 gal)	1.900	
Annual (\$)	633,858	
Chemicals		
MU & WT Chem.		
Price (\$/lb)	0.28	
Annual (\$)	430,245	
Carbon (Trace Removal)		
Price (\$/lb)	1.6	
Initial Cost (\$)	643,107	
Annual (\$)	273,320	
COS Catalyst		
Price (\$/m ³)	45,909	
Initial Cost (\$)	10,142,870	
Annual (\$)	2,155,360	
Selexol Solution		
Price (\$/gal)	38.00	
Initial Cost (\$)	14,273,030	
Annual (\$)	439,349	
Claus / DSRP Catalyst		
Price (\$/ft ³)	48	
Annual (\$)	20,291	
ZnO polishing sorbent		
Price (\$/lb)	1.5	
Initial Cost (\$)	223,761	
Annual (\$)	702,373	
KOH coal catalyst makeup		
Initial Fill (ft ³)	788,680	
Use (ft ³ /day)	13,145	
Price (\$/ft ³)	0.19	
Initial Cost (\$)	150,591	
Annual (\$)	778,682	
Lime for catalyst recovery		
Initial Fill (lb)	3,363,605	
Use (lb/day)	56,060	
Price (\$/lb)	0.12	
Initial Cost (\$)	403,633	
Annual (\$)	2,087,117	
Sub Total Chemicals		

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Cost Component	Cost	Cost (\$/MWh)
	2018\$	2018\$
Initial Fill (\$)	27,395,357	
Annual (\$)	7,345,079	
Waste Disposal		
Spent Trace Catalyst		
Price (\$/lb)	0.42	
Annual (\$)	76,060	
Ash + HCl Sorbent		
Price (\$/ton)	16.31	
Annual (\$)	175,098	
Spent sorbents		
Price (\$/lb)	0.42	
Annual (\$)	196,664	
Subtotal-Waste Disposal (\$)	447,823	
TOTAL VARIABLE OPERATING COSTS		
Initial Fill (\$)	27,395,357	
Annual (\$)	31,443,604	6.50
Fuel Coal		
Price (\$/ton)	51.96	
Annual (\$)	63,600,564	13.14

Exhibit 5-29. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – Capital Cost Components

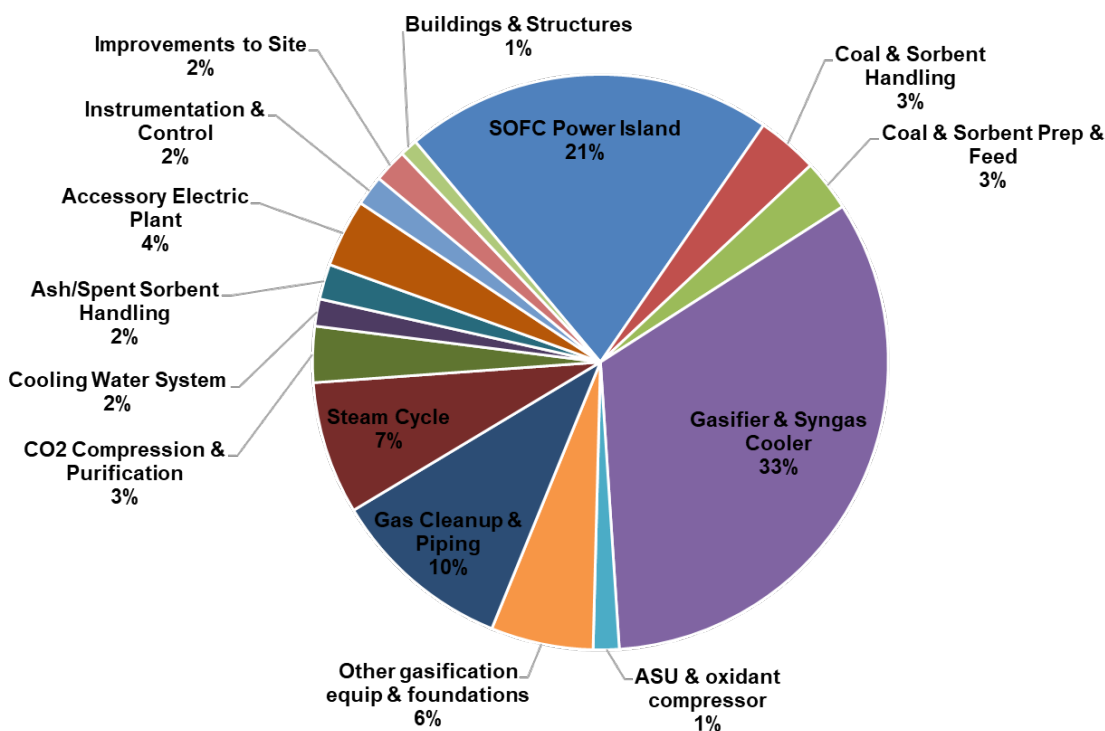


Exhibit 5-30. Case PCATIGFC6B – Advanced Gasifier Pressurized IGFC Plant with CCS – BOP Cost Distribution

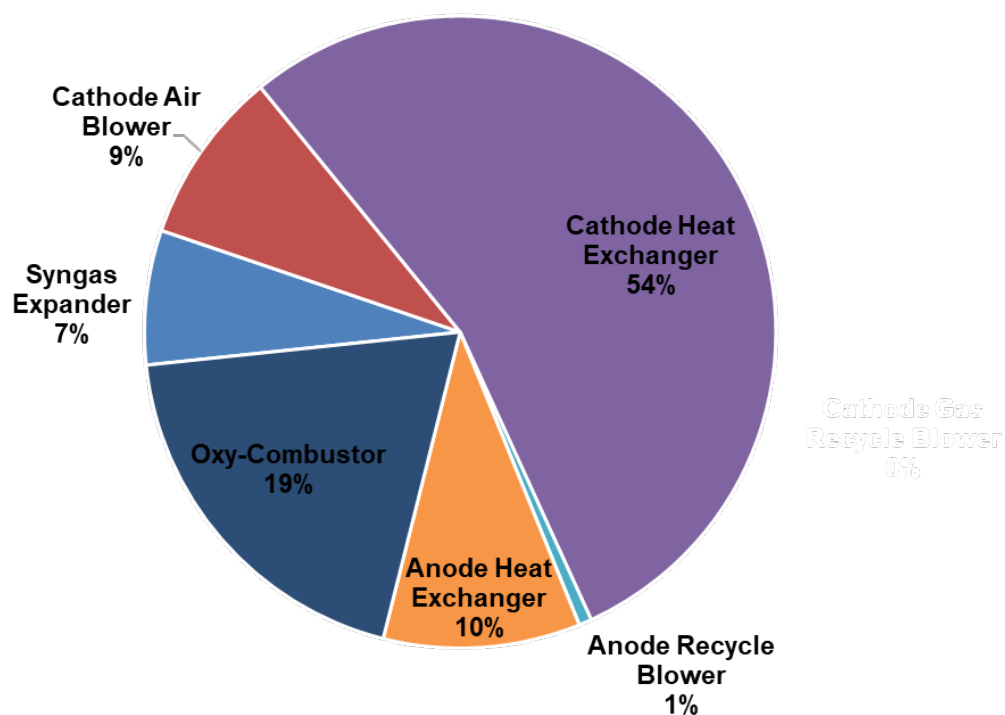


Exhibit 5-31. Case PCATIGFC6A – Advanced Gasifier Pressurized IGFC Plant without CCS – Capital Cost Components

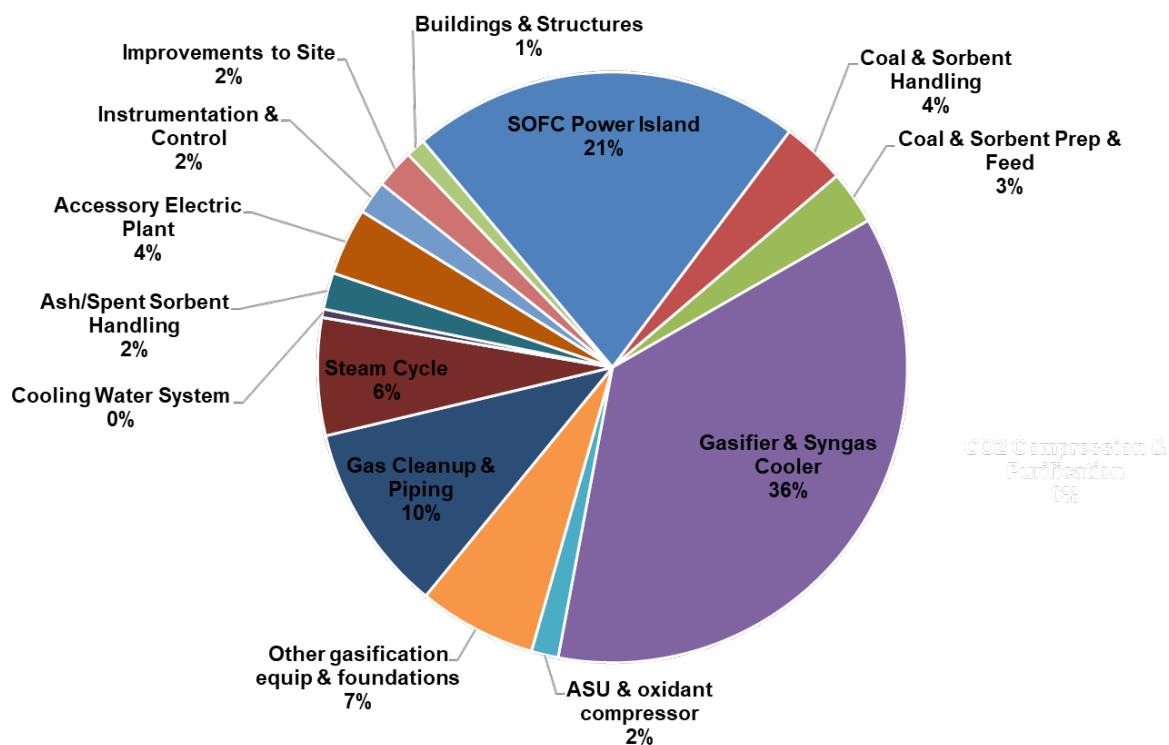


Exhibit 5-32. Case PCATIGFC6A – Advanced Gasifier Pressurized IGFC Plant without CCS – BOP Cost Distribution

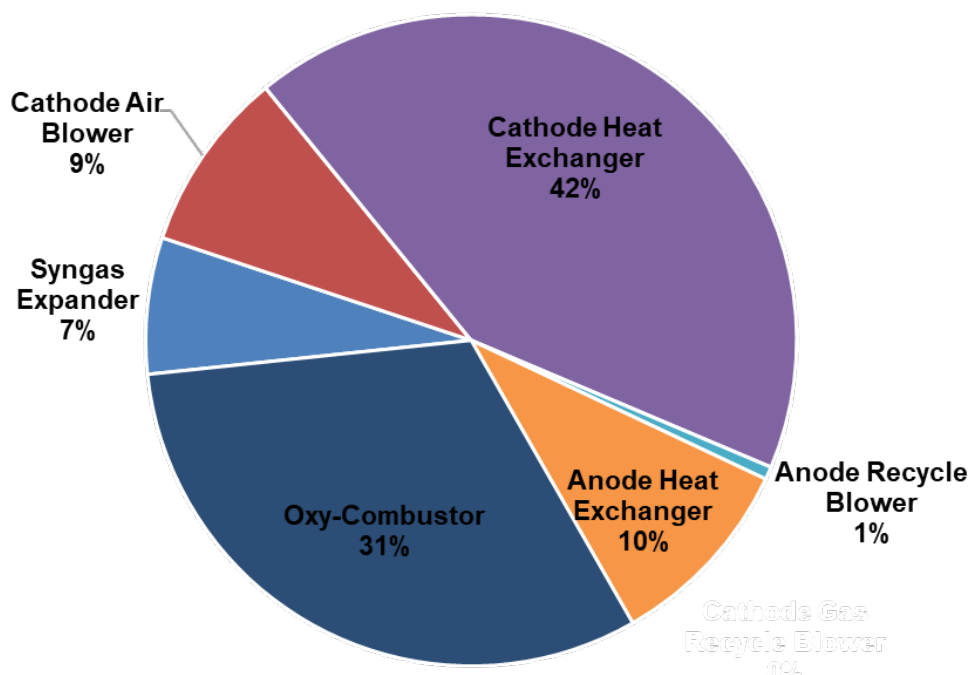


Exhibit 5-33. Case 6 – Advanced Gasifier Pressurized IGFC Plant with and without CCS – Levelized Cost of Electricity

Levelized Cost of Electricity (2018\$/MWh)		
Carbon Capture	With CCS	No CCS
Component	\$/MWh	
Variable Costs	19.6	17.3
<i>Fuel Costs</i>	13.1	11.1
<i>Variable O&M Costs</i>	6.5	6.2
Fixed O&M Costs	9.4	8.5
Capital Costs	33.4	29.6
Total LCOE (excluding T&S)	62.4	55.4
T&S	5.0	0.0
Total LCOE (including T&S)	65.5	55.4

5.3 PATHWAY 2 RESULTS

The performance and costs of the pressurized pathway IGFC plants are summarized in Exhibit 5-34, Exhibit 5-35, Exhibit 5-36, and Exhibit 5-37 for plants with and without CCS. The efficiency varies from 38.6 percent in the reference case to 57.8 percent for the atmospheric plants with CCS, while the LCOE declines from a value of \$106.7/MWh to a value of \$62.4/MWh, representing a 42 percent reduction in costs through technological advancements along the pathway. For plants without CCS, the pathway results in a 36 percent reduction in LCOE, along with an improvement in plant efficiency of 15 percentage points.

Exhibit 5-34. Pathway 2 – Pressurized IGFC Plants with CCS Performance Summary

Case		Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology		Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)		0.2						
Fuel Utilization (%)		80	85					
Capacity Factor (%)		80		85		80	85	
Inverter Efficiency (%)		97						98
SOFC Stack Cost (\$/kW)		225						200
Performance								
SOFC Parameters	Current Density (mA/cm²)	400	400	400	400	400	400	400
	Cell Potential (V)	0.878	0.869	0.869	0.874	0.883	0.883	0.883
	Power Density (mW/cm²)	351	348	348	349	353	353	353
Gross Power (MWe)		1,065	1,109	1,109	1,042	784	784	781
Auxiliary Loads (MWe)		415	459	459	393	134	134	131
Air Separation Unit (kWe)		49,451	46,259	46,259	39,577	6,511	6,511	5,151
CO ₂ Drying, Purification and Compression (CPU) (kWe)		45,890	45,830	45,830	42,320	29,960	29,960	29,600
SOFC Blowers (kWe)		259,070	306,220	306,220	259,400	75,160	75,160	74,270
Compressors (kWe)		33,910	33,860	33,860	27,680	7,812	7,812	7,717
Steam Cycle and Miscellaneous (kWe)		27,023	26,557	26,557	24,327	44,950	44,950	44,450
Net Power (MWe)		650	650	650	649	650	650	650
Coal Flowrate (lb/hr)		492,800	492,200	492,200	455,195	332,700	332,700	328,775
Net Electric Efficiency, HHV (%)		38.6	38.6	38.6	41.7	57.1	57.1	57.8
Net Plant Heat Rate, HHV (Btu/kWh)		8,844	8,831	8,831	8,185	5,971	5,971	5,901
CO ₂ Capture rate (%)		98.8	98.7	98.7	98.7	98.6	98.6	98.6
CO ₂ Captured (tonnes per year)		3,950,953	3,945,823	4,192,437	3,876,712	2,551,459	2,710,925	2,678,833
CO ₂ Emissions (lb/MWhgross)		20.0	20.5	20.5	19.6	15.6	15.6	15.4
CO ₂ Emissions (lb/MWhnet)		32.9	35.0	35.0	31.5	18.7	18.7	18.5
Raw Water Consumption (gpm/MWhnet)		2.15	1.79	1.79	1.51	1.37	1.37	1.35

Exhibit 5-35. Pathway 2 – Pressurized IGFC Plants with CCS Cost Summary

Case	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology	Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)	0.2						
Fuel Utilization (%)	80	85					
Capacity Factor (%)	80		85		80	85	
Inverter Efficiency (%)	97						98
SOFC Stack Cost (\$/kW)	225						200
Cost							
Total Plant Cost (TPC) (1000\$)	2,725,410	2,745,132	2,745,132	2,627,353	1,631,795	1,631,795	1,600,867
Total Overnight Cost (TOC) (1000\$)	3,378,063	3,402,160	3,402,027	3,253,949	2,015,238	2,015,160	1,977,372
Total As-Spent Cost (TASC) (1000\$)	3,898,284	3,926,092	3,925,939	3,755,057	2,325,585	2,325,495	2,281,888
Levelized Cost of Electricity (\$/MWh)							
Variable Costs	29.8	29.8	29.4	27.6	20.5	20.2	19.6
Fuel Costs	19.7	19.7	19.7	18.2	13.3	13.3	13.1
Variable O&M Costs	10.1	10.1	9.7	9.3	7.2	6.9	6.5
Fixed O&M Costs	16.1	16.2	15.2	14.7	10.2	9.6	9.4
Capital Costs	60.8	61.3	57.7	55.3	36.2	34.0	33.4
Total LCOE (excluding T&S)	106.7	107.3	102.3	97.5	66.8	63.8	62.4
T&S	7.9	7.9	7.9	7.3	5.1	5.1	5.0
Total LCOE (including T&S)	114.6	115.1	110.1	104.8	71.9	68.8	67.5

Exhibit 5-36. Pathway 2 – Pressurized IGFC Plants without CCS Performance Summary

Case		Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology		Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)		0.2						
Fuel Utilization (%)		80	85					
Capacity Factor (%)		80		85		80	85	
Inverter Efficiency (%)		97						98
SOFC Stack Cost (\$/kW)		225						200
Performance								
SOFC Parameters	Current Density (mA/cm²)	400	400	400	400	400	400	400
	Cell Potential (V)	0.878	0.869	0.869	0.874	0.883	0.883	0.883
	Power Density (mW/cm²)	351	348	348	349	353	353	353
Gross Power (MWe)		902	915	915	887	739	739	737
Auxiliary Loads (MWe)		252	264	264	237	89	89	87
Air Separation Unit (kWe)		25,296	23,284	23,284	20,150	5,556	5,556	4,402
CO ₂ Drying, Purification and Compression (CPU) (kWe)		0	0	0	0	0	0	0
SOFC Blowers (kWe)		185,600	202,200	202,200	181,450	63,560	63,560	62,910
Compressors (kWe)		24,310	22,480	22,480	19,440	11,637	11,637	11,514
Steam Cycle and Miscellaneous (kWe)		17,003	16,491	16,491	15,645	11,229	11,229	11,145
Net Power (MWe)		650	650	650	650	650	650	650
Coal Flowrate (lb/hr)		353,060	325,005	325,005	318,400	281,360	281,360	278,470
Net Electric Efficiency, HHV (%)		53.0	52.6	52.6	54.6	67.6	67.6	68.3
Net Plant Heat Rate, HHV (Btu/kWh)		6439	6487	6487	6247	5048	5048	4997
CO ₂ Capture rate (%)		0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Captured (tonnes per year)		0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO ₂ Emissions (lb/MWhgross)		920.1	878.7	878.7	878.9	847.5	847.5	841.1
CO ₂ Emissions (lb/MWhnet)		1277.0	1236.0	1236.0	1198.8	962.9	962.9	953.1
Raw Water Consumption (gpm/MWhnet)		0.72	0.70	0.70	0.55	0.52	0.52	0.52

TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 5-37. Pathway 2 – Pressurized IGFC Plants without CCS Cost Summary

Case	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Gasifier Technology	Conventional			Enhanced	Catalytic		
SOFC Degradation Rate (%/1000 h)	0.2						
Fuel Utilization (%)	80	85					
Capacity Factor (%)	80		85		80	85	
Inverter Efficiency (%)	97						98
SOFC Stack Cost (\$/kW)	225						200
Cost							
Total Plant Cost (TPC) (1000\$)	2,316,541	2,321,148	2,321,148	2,246,055	1,414,661	1,414,661	1,484,549
Total Overnight Cost (TOC) (1000\$)	2,833,529	2,836,867	2,836,778	2,745,191	1,749,245	1,749,176	1,832,967
Total As-Spent Cost (TASC) (1000\$)	3,269,892	3,273,745	3,273,642	3,167,951	2,018,628	2,018,549	2,115,244
Levelized Cost of Electricity (\$/MWh)							
Variable Costs	21.9	23.2	22.8	21.8	17.8	17.4	17.3
Fuel Costs	14.6	15.9	15.9	15.1	11.2	11.2	11.1
Variable O&M Costs	7.3	7.3	6.9	6.7	6.5	6.2	6.2
Fixed O&M Costs	13.9	14.0	13.1	12.8	9.1	8.5	8.5
Capital Costs	51.2	51.3	48.3	46.7	31.6	29.8	29.6
Total LCOE (excluding T&S)	87.0	88.4	84.2	81.3	58.4	55.7	55.4
T&S	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total LCOE (including T&S)	87.0	88.4	84.2	81.3	58.4	55.7	55.4

6 DISCUSSION OF PATHWAY RESULTS

The section discusses impact of the technological developments and the cost reduction assumptions considered in the present pathway studies on the performance and cost of an IGFC plant. The results from the various cases are consolidated to provide guidance to the fuel cell R&D program at NETL. Further, the IGFC system performance and costs are compared with conventional heat-engine technologies — specifically, the SC PC, IGCC, and F-class NGCC plants from the Baseline study [2], J-class NGCC plants [34], coal and NG-based sCO₂ systems [35], [36], [37] wherever applicable.

6.1 SYSTEM PERFORMANCE AND EFFICIENCY

The efficiencies of the various IGFC plants considered in the two pathways are shown in Exhibit 6-1. For an atmospheric IGFC plant without CCS, the HHV efficiency varies from a value of 53.0 percent for the reference plant to a value of 61.0 percent for the Case 6 advanced plant with a catalytic gasifier. Inclusion of CCS imposes an efficiency penalty of ≈ 3.5 –4.0 percentage points for atmospheric systems. The VGR configuration significantly reduces this penalty to nearly a percent in the case of the conventional gasifier, while the VGR systems with CCS have even a slight efficiency advantage over systems without CCS. Indeed, the highest efficiency of all the CCS systems considered, including pressurized systems, is achieved by the catalytic VGR system. The VGR concept enables operating at a fuel utilization of 97.5 percent while the in-stack utilization is maintained below 60 percent.

Pressurization generally results in an increase in efficiency over the atmospheric IGFC cases (without VGR) except for the conventional and enhanced gasifier cases with CCS due to excessive process air required (and an increase in the air compression losses) in these cases in the absence of cathode gas recirculation. The pressurized conventional and enhanced gasifier cases without CCS also include an efficiency advantage due to the injection of natural gas into the combustor to enhance cathode expander inlet temperatures.

As expected with the underlying SOFC technology advantages, the IGFC efficiencies are significantly higher than the values for comparable conventional technologies including, the SC PC and sCO₂ indirect coal plants with CCS [2, 35] shown in the Exhibit 6-1. The IGFC plants start to close in on the efficiency of even an H-class NGCC system with enhanced gasification technology, while the catalytic gasifier based IGFC systems result in efficiencies that are 15–20 percentage points higher relative to conventional systems depending on the inclusion of CCS.

The catalyst gasifier technology has the highest impact (≈ 9 –15 percentage points) on the plant efficiency with or without CCS, while the VGR concept, which requires minimal advances in technology, is the next-highest efficiency booster, resulting in ≈ 2.4 –4.7 percentage points improvement in performance, as indicated by the waterfall plots in Exhibit 6-2 and Exhibit 6-3. The BOP enhancements resulted only in modest gains, since the ASU parasitic is lower for the Catalytic Gasifier cases (Cases 4–6 for all the configurations), which have significantly lower O₂ requirement relative to the conventional and enhanced gasifier plants due to higher plant efficiency.

Exhibit 6-1. HHV Efficiency of the Pathway IGFC Plants

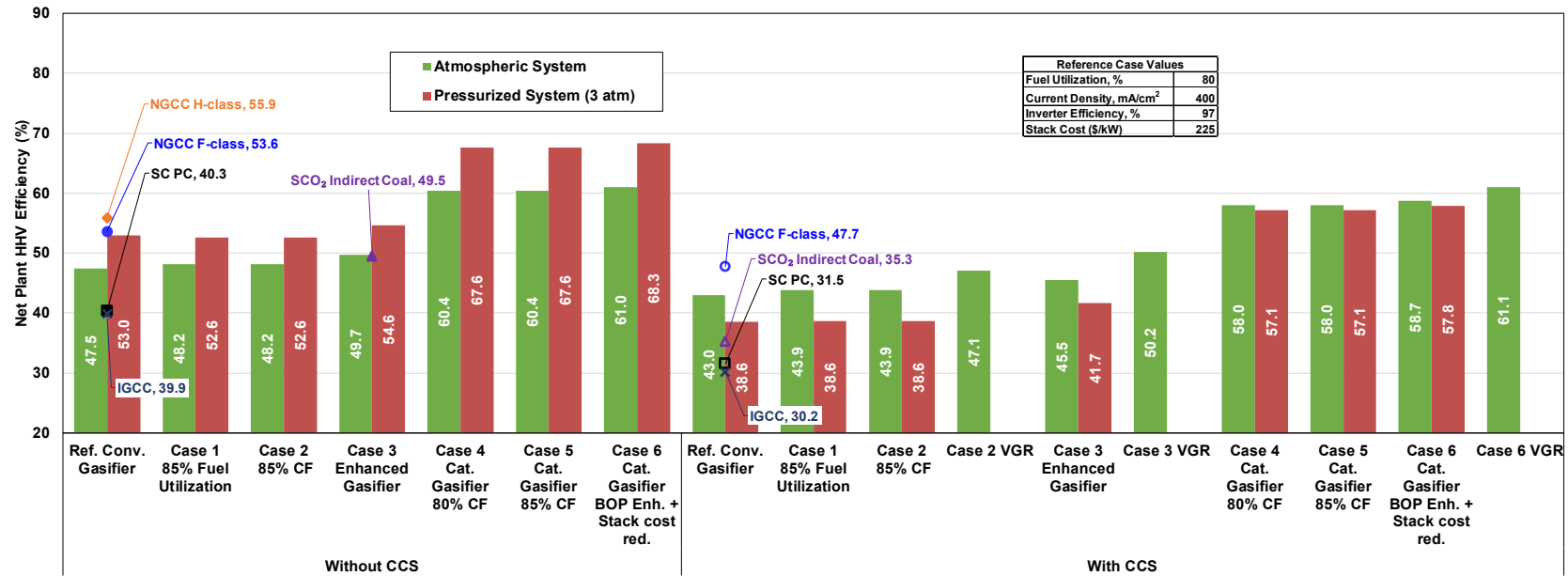


Exhibit 6-2. Technology Developments and Efficiency Improvements for IGFC Plants without CCS

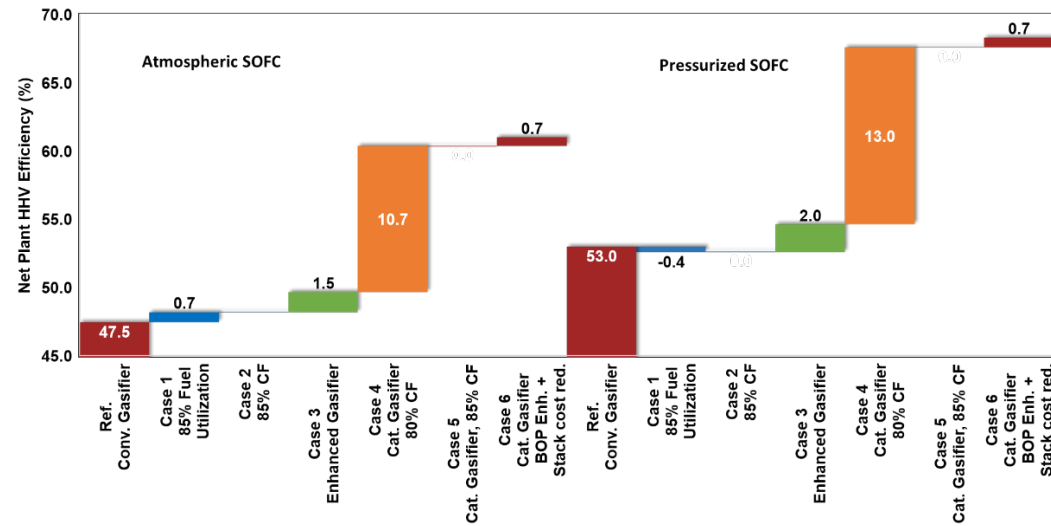
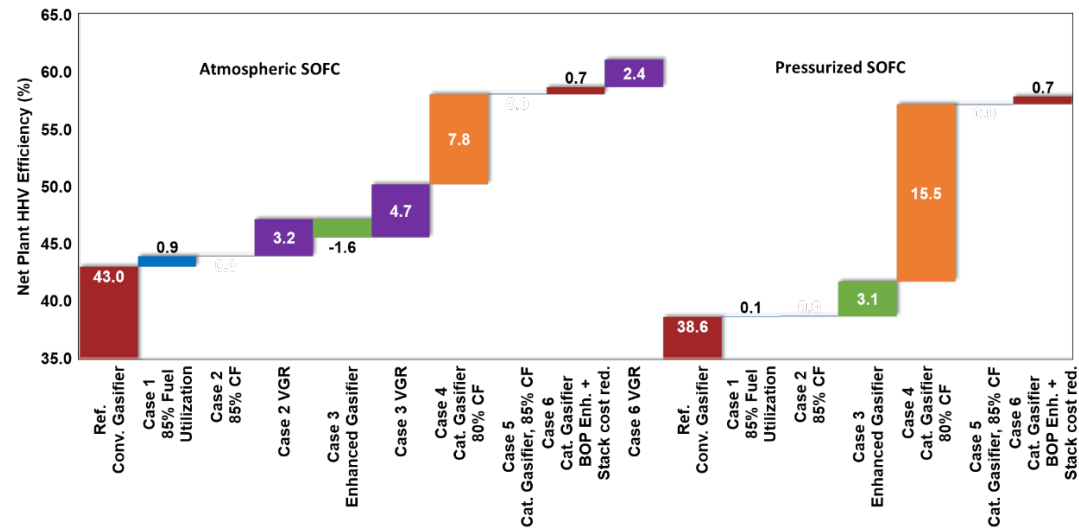


Exhibit 6-3. Technology Developments and Efficiency Improvements for IGFC Plants with CCS



6.2 IGFC PLANT COSTS

The catalytic gasifier cases result in the lowest TPC for the IGFC plants with and without CCS, as shown in Exhibit 6-4. This is attributable to the high CH₄ content in the syngas, which significantly reduces airflow requirements. The pressurized configurations generally result in a higher TPC relative to the atmospheric systems, primarily due to the increased enclosure (pressure vessel) costs and to the inclusion of cathode expander costs, which seem to negate any efficiency advantages of pressurization on the capital costs. The TPC of IGFC plants with conventional or enhanced gasification technology, are generally higher than conventional power generation technologies. A reduction in costs of conventional gasifiers, which dominates the plant costs for these cases, would significantly affect the relative comparison. The reduction in TPC for catalytic gasifier IGFC systems is mainly due to the projected lower costs for these types of gasifiers operating at temperatures lower than conventional technologies.

The LCOE (without T&S) of the reference atmospheric conventional gasifier IGFC plant with CCS is lower (by \$8/MWh) than the LCOE of SC PC plants with CCS, and is comparable to the LCOE of the sCO₂ indirect coal plant also with CCS, as shown in Exhibit 6-4.^{i, j} All the other IGFC plants with CCS have a lower LCOE than the coal-based conventional technologies with the catalytic gasifier IGFC plants, indicating a significant LCOE advantage even relative to F-class NGCC plants with CCS. In general, the LCOE of pressurized systems are slightly higher than the corresponding atmospheric systems. The VGR configuration substantially mitigates the LCOE penalty of CCS (by nearly half).

On the other hand, all the IGFC plants without CCS have higher costs relative to conventional coal-based technologies, mainly attributable to the higher capital costs of the gasifier, the cost of which has nearly doubled relative to the previous pathway study. [3] Indeed, the dominance of the capital portion of the plant is reflected by the breakdown of the LCOE with and without T&S shown in Exhibit 6-5 and Exhibit 6-6, respectively. A 10-percent reduction in conventional gasifier costs would reduce the LCOE by nearly 5-percent for the reference case. Although a reduced-cost gasifier was not included in the pathway, the catalytic gasifier serves to illustrate the effect of reducing gasifier costs.

The difference in LCOEs between an IGFC plant with and without CCS is significantly smaller than the corresponding penalties for conventional technologies. The SOFC is essentially an oxyfuel reactor, and along with the sealed design generally used to separate the air and fuel, it forms a highly effective inherent carbon separator (that produces power); it produces a concentrated CO₂ effluent that is ready for CCS with minimal incremental costs. This underscores the leading role played by SOFC-based systems in meeting DOE-FECM's environmental vision of reducing carbon emissions.

ⁱ The LCOE for H-class have been estimated to be ≈\$36/MWh in the study by Uysal. [35] However, these were not included on the charts since the costs are based on a 1 GW plant capacity and the financial parameters used are different than the baseline studies.

^j The LCOE for the sCO₂ indirect coal plant from the study of White et al. [36] has been adjusted to use the latest financial parameters.

Exhibit 6-4. Comparison of the TPC of the IGFC Plants

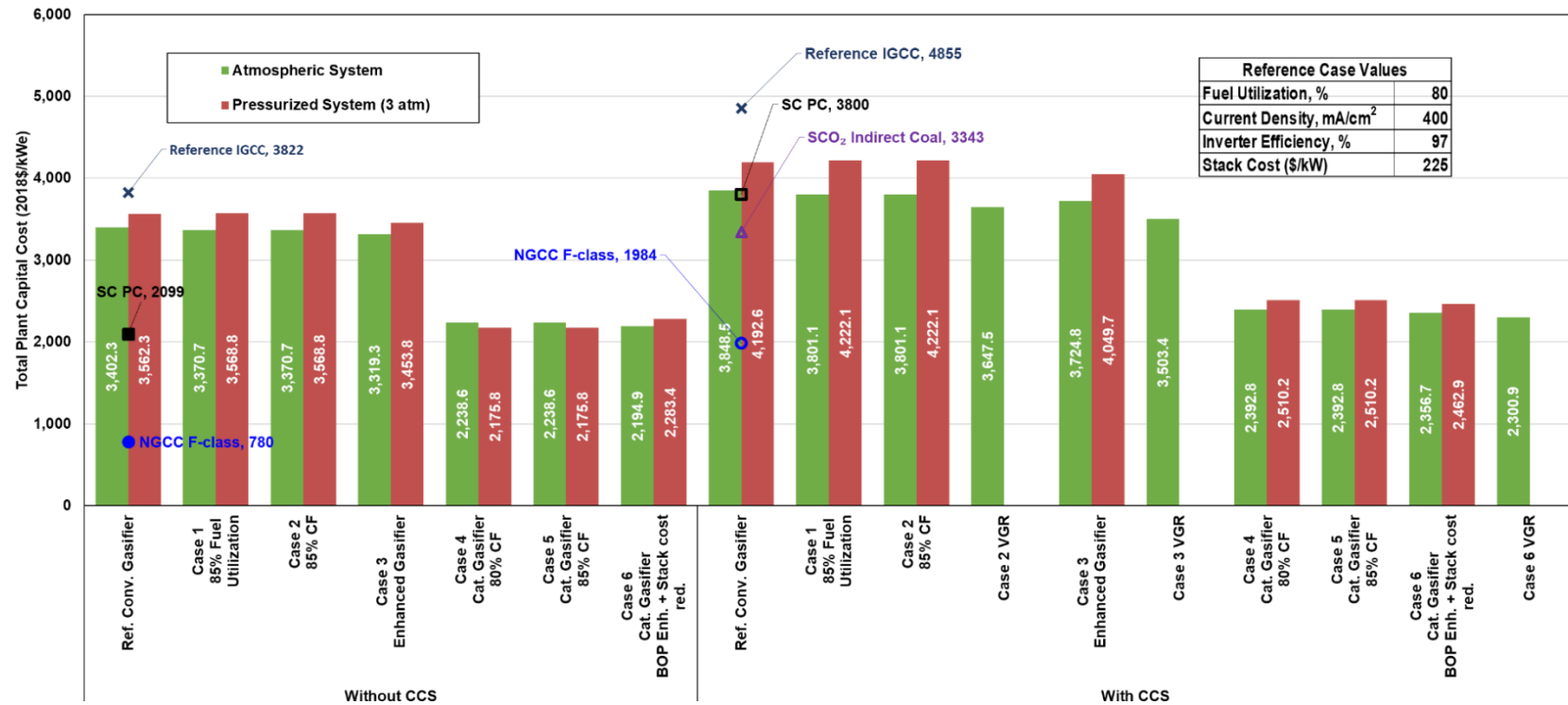
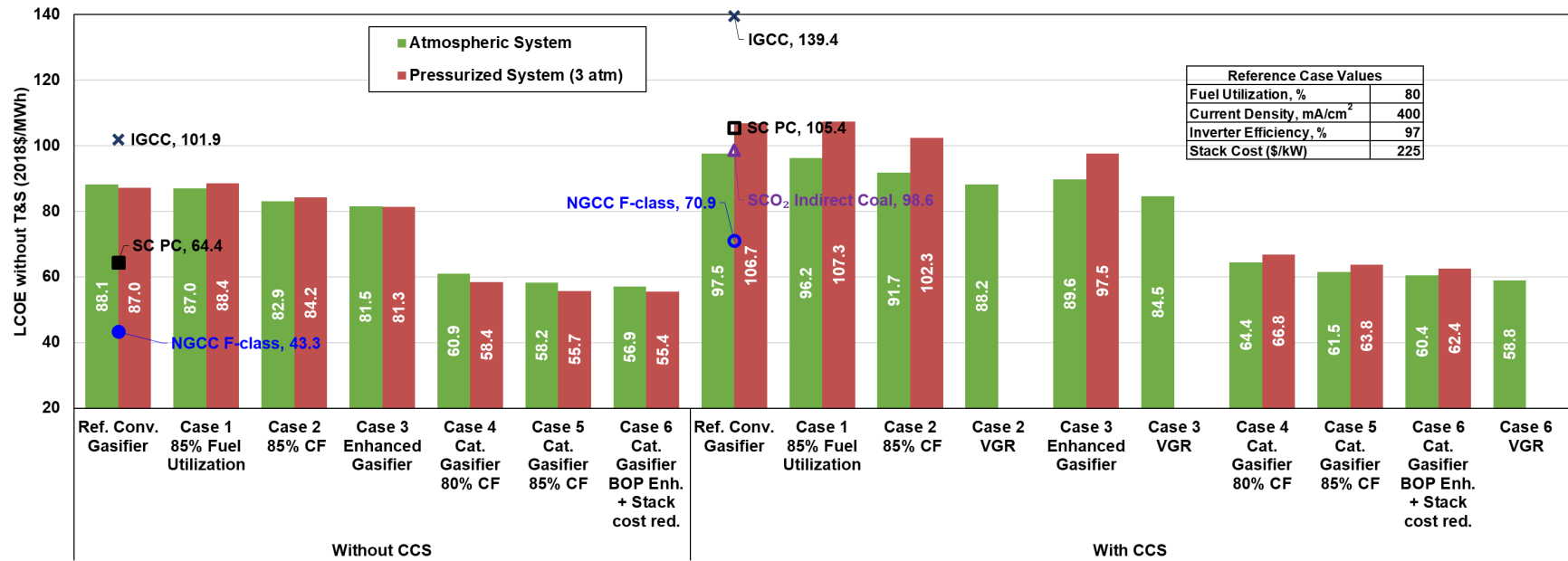


Exhibit 6-5. Comparison of the LCOE (without T&S) of the IGFC Plants



TECHNO-ECONOMIC ANALYSIS OF INTEGRATED GASIFICATION FUEL CELL SYSTEMS

Exhibit 6-6. LCOE Breakdown (with T&S) of the IGFC Plants

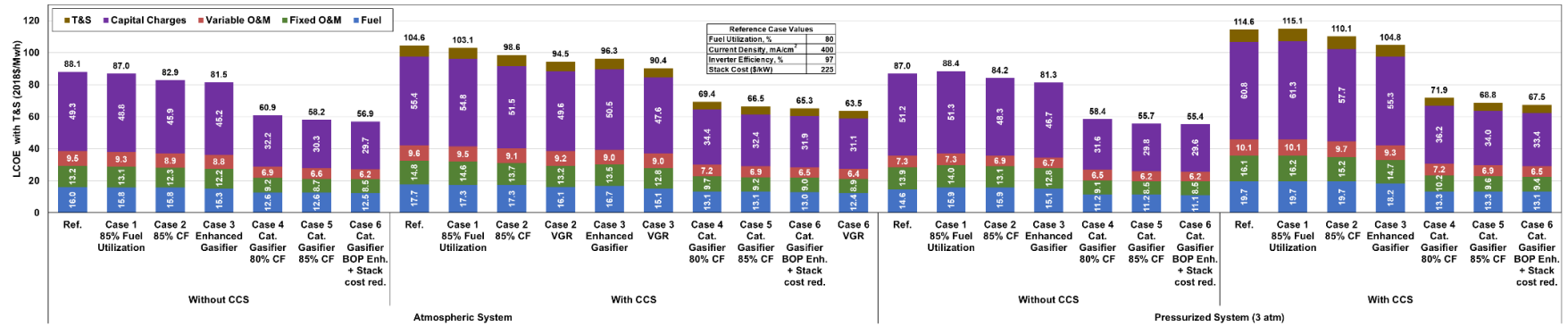
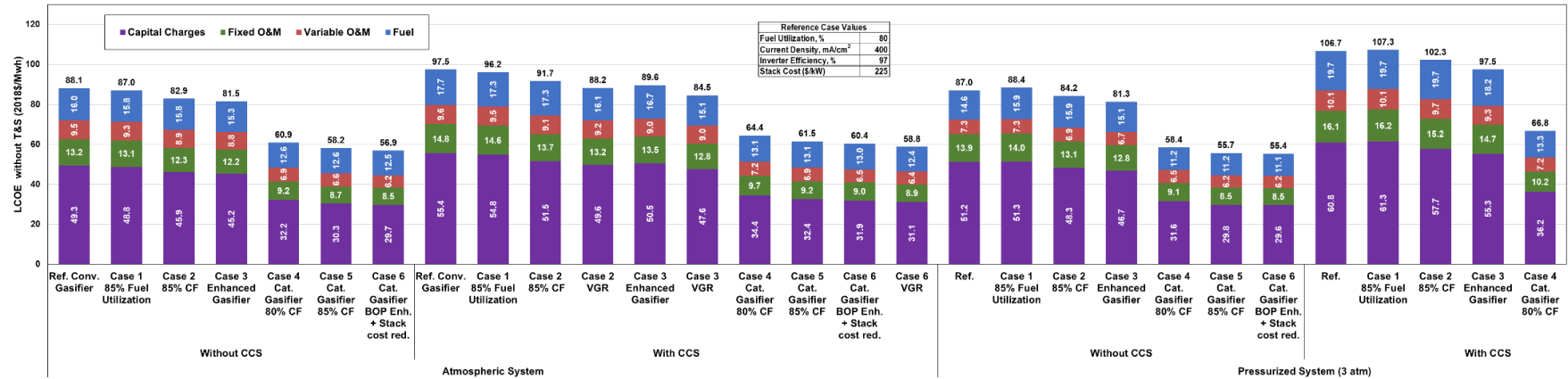
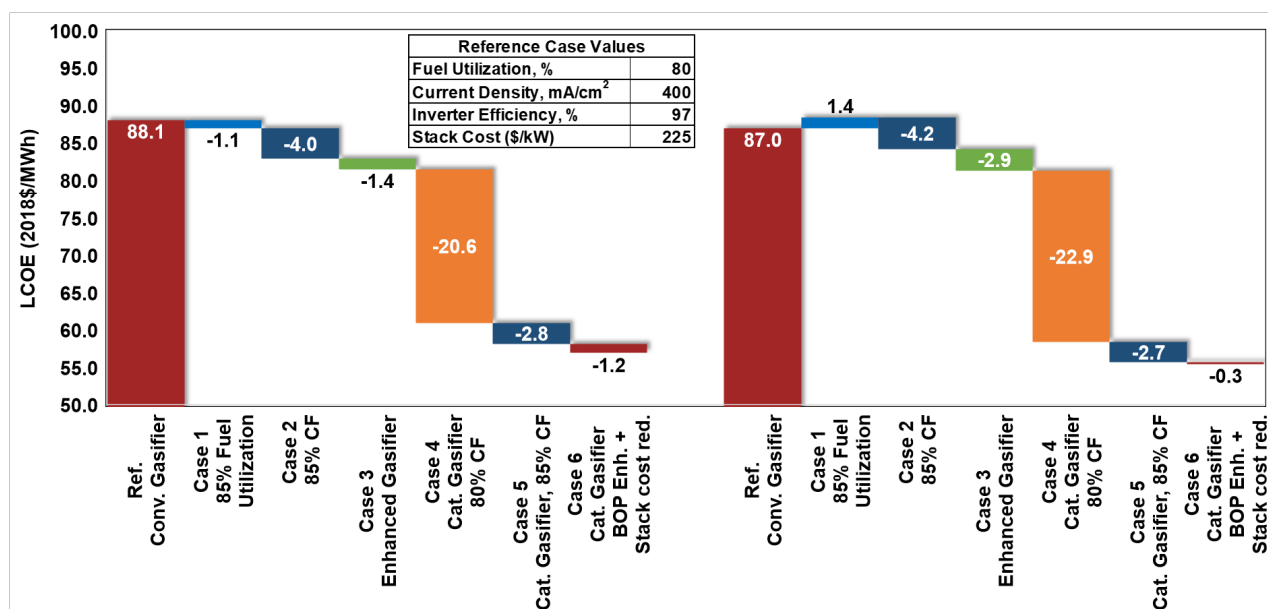


Exhibit 6-7. LCOE Breakdown (without T&S) of the IGFC plants

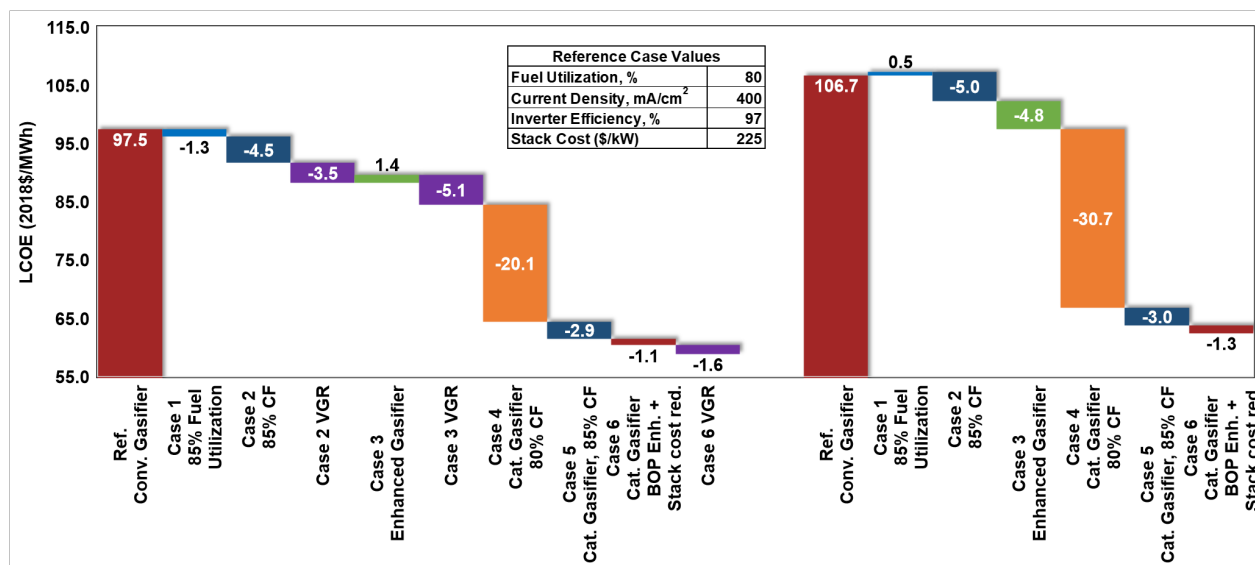


The progression of LCOE with pathway technology developments, cost reductions, and increased availability are shown in Exhibit 6-8 and Exhibit 6-9 for the IGFC plants without and with CCS respectively. In the case of IGFC plants with CCS, the largest reduction in LCOE, $\approx \$31/\text{MWh}$, is associated with the introduction of catalytic gasifier technology. Increase in system capacity factor and the VGR configuration also yield significant LCOE reductions, ranging between $\approx \$2\text{--}5/\text{MWh}$ depending on the configuration. The introduction of catalytic gasification technology and increase capacity factor affect the LCOE of the IGFC systems without CCS significantly.

Exhibit 6-8. Progression of Plant LCOE (without T&S) along the IGFC Pathway without CCS



The pathways, which represent practical and realizable steps consistent with the fuel cell program, lead to an IGFC system with capture that has a significantly low LCOE relative to conventional heat-engine based technologies with CCS. The pathways also lead to IGFC plants without CCS that are economically competitive with F-class NGCC plants. Although pressurization does not appear to have a significant advantage in the cases analyzed here, pressurized configurations could be attractive as a hybrid system where operational flexibility aspects may be advantageous.

Exhibit 6-9. Progression of Plant LCOE (without T&S) along the IGFC Pathway with CCS

6.3 COST OF CO₂ CAPTURED

The breakeven CO₂ sales price (the price of CO₂ that is required to pay for the difference in LCOE between a plant with and without CCS) for the pathway IGFC cases are shown in Exhibit 6-10. All the atmospheric IGFC plants and advanced pressurized plants considered herein have a breakeven CO₂ sales price that is well below the \$40/tonne of CO₂ generally considered to be an achievable selling price for pure CO₂ for EOR purposes. The atmospheric IGFC systems with the VGR and the atmospheric systems with catalytic gasifiers result in breakeven price that are even below \$10/tonne, which could make these systems highly competitive with other CO₂ sources.

6.4 WATER CONSUMPTION

The IGFC plants result in significantly lower water consumption compared to SC PC and NGCC plants, as shown in Exhibit 6-11. The IGFC plant water consumption is dominated by the gasifier and the cooling water makeup requirements of the steam cycle.

Exhibit 6-10. Breakeven Sale Price of CO₂ Captured for the Various IGFC Plants

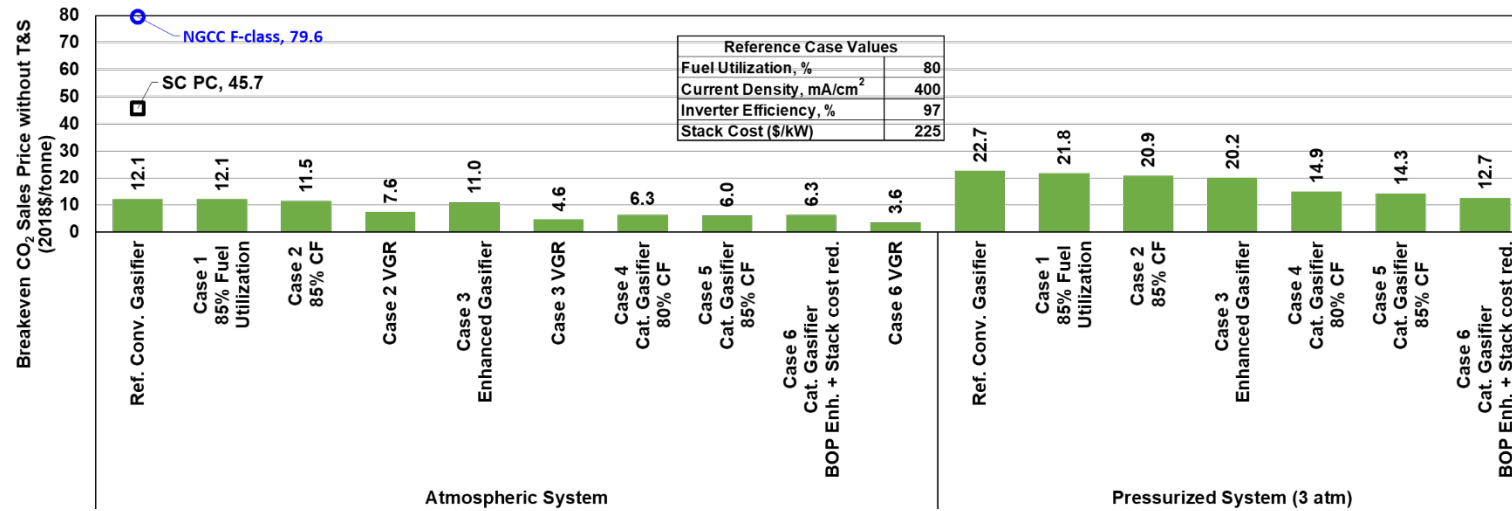
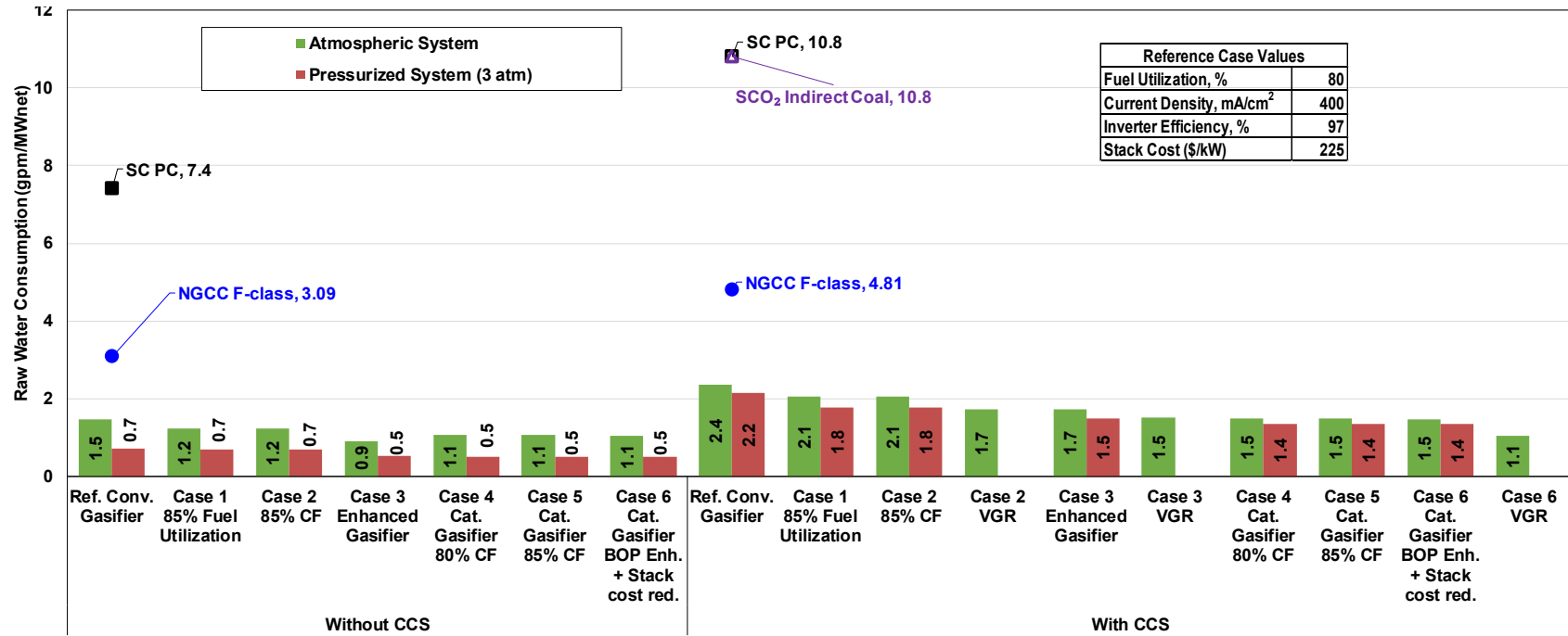


Exhibit 6-11. Water Consumption for the Various IGFC Plants



6.5 EFFECT OF NATURAL GAS PRICE

The LCOE of the reference IGFC plant (Atmospheric Conventional Gasifier Case 0) and the best case advanced IGFC plant (Case 6 with VGR) with CCS is shown in Exhibit 6-12 as function of natural gas price. The variation of the LCOE of a conventional F-class NGCC plant with CCS is also shown, along with the LCOE of an SC PC plant, which does not depend directly on the price of natural gas. With CCS, the reference IGFC plant costs become competitive relative to NGCC systems with CCS above a natural gas price of $\approx \$8/\text{MMBtu}$. In plants without CCS, natural gas prices above $\approx 12/\text{MMBtu}$, as shown in Exhibit 6-13, are required to make the IGFC plants competitive relative to conventional NGCC systems.

Exhibit 6-12. Influence of Natural Gas price on the LCOE (excl. T&S) of IGFC Plants with CCS

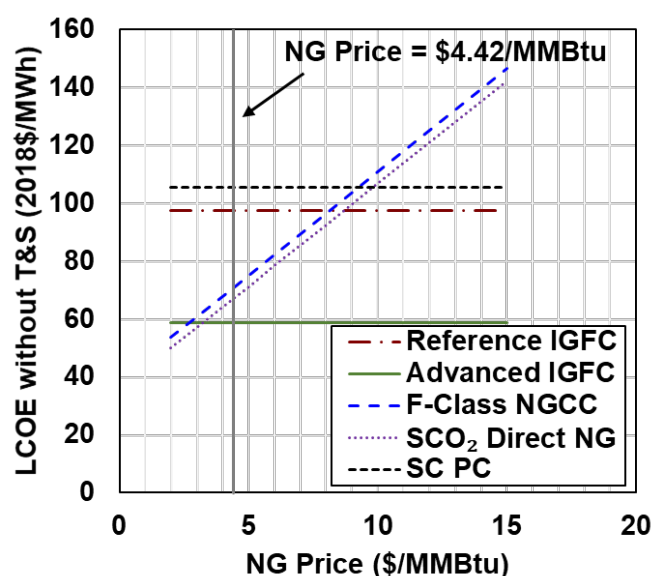
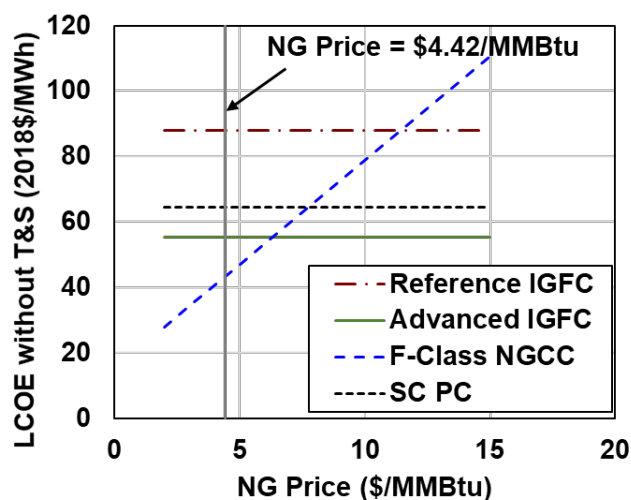


Exhibit 6-13. Influence of Natural Gas Price on the LCOE (excl. T&S) of IGFC Plants without CCS



7 CONCLUSIONS

The results of a pathway study for coal based, IGFC power systems with and without CCS are presented in this report. Two pathway scenarios are considered, based on the operating pressure of the SOFC. Three gasification technologies, including a conventional technology, differing in the syngas CH₄ content, which has a significant effect on the IGFC system performance, were considered. The performance and cost benefits were estimated for a series of projected gains made through the development of advances in the component technologies or improvements in plant costs and availability. The IGFC systems with CCS included a CO₂ purification unit that met EOR specifications for the product CO₂ stream.

The results indicate that

- IGFC systems have the potential to result in HHV efficiencies that exceed 60 percent even with CCS. The present pathway resulted in a system with efficiency as high as 68.1 percent without capture.
- While the IGFC plants with conventional technology have a significant cost advantage over coal-based conventional technologies, the IGFC economic competitiveness relative to natural gas-based heat engines (at current natural gas prices) are achievable only with significant advances and considerable cost reductions in gasification technology. In particular, the development of gasifiers with higher methane content in the syngas is desirable.
- The difference in LCOE between an IGFC plant with and without CCS is significantly smaller than the corresponding penalties for conventional technologies. The SOFC is, essentially, an oxyfuel reactor and, along with the sealed design generally used to separate the air and fuel, it forms a highly effective inherent carbon separator (that produces power); and it produces a concentrated CO₂ effluent that is ready for CCS with minimal incremental costs. Indeed, breakeven values of CO₂ prices below \$10/tonne are achievable.
- The VGR concept has the potential to significantly reduce the LCOE in the short-term even with conventional technologies.
- Pressurization does not appear to have a significant advantage in the cases analyzed here. However, pressurized configurations could be found to be attractive as a hybrid system where operational flexibility aspects may be attractive.
- The IGFC consumes 50 percent less raw water than even the water economical NGCC system with CCS.
- A pressurized SOFC provides no cost benefit over systems operating with an atmospheric-pressure SOFC. However, the IGFC plant configuration and operating conditions selected for the pressurized SOFC evaluation in this study have not been optimized and, thus, there are opportunities for further benefit. In particular, the cost of the pressure vessel is an important parameter that needs further refinement. This IGFC configuration requires development of the pressurized-SOFC technology.

There are other technological innovations that might also benefit the IGFC power plant performance and cost, such as humid gas cleaning, the ion transport membrane technology for oxygen separation incorporating integration with the pressurized SOFC cathode air compressor, and shock-wave CO₂ compression. Further, in the near-term, natural gas injection as a proxy for a catalytic gasifier could greatly increase the performance and cost potential of the IGFC plant using conventional or enhanced-conventional coal gasification. These technology advances will be explored in future IGFC pathway evaluations.

8 REFERENCES

- [1] National Energy Technology Laboratory, "Solid Oxide Fuel Cell," [Online]. Available: <https://www.netl.doe.gov/research/coal/energy-systems/fuel-cells>. [Accessed 25 May 2018].
- [2] National Energy Technology Laboratory, "Cost and Performance for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity," Revision 4, September 2019.
- [3] National Energy Technology Laboratory, "Techno-Economic Analysis of Integrated Gasification Fuel Cell Systems," DOE/NETL, November 24, 2014.
- [4] National Energy Technology Laboratory, "QGESS: Specifications for Selected Feedstocks," DOE/NETL, March 2012.
- [5] National Energy Technology Laboratory, Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases, DOE/NETL, 2012.
- [6] D. Denton, *Telephone communication*, Eastman Chemical Co.
- [7] National Energy Technology Laboratory, "QGESS: Carbon Dioxide Transport and Storage Costs in NETL Studies," DOE/NETL, September 30, 2013.
- [8] A. Khuri and S. Mukhopadhyay, "Response Surface Methodology," *WIREs Comp Sta*, vol. 2, no. 2, p. 128, 2010.
- [9] B. Koeppel, C. Lai, A. Iyengar, Z. Xu and G. Hackett, "Use of a Reduced Order Model (ROM) to Simulate SOFC Performance in System Models," in *Electrochemical Society (ECS) Transactions*, 2017.
- [10] K. Lai, B. Koeppel, K. Choi, K. Recknagle, X. Sun, L. Chick, V. Korolev and M. Khaleel, "A Quasi-two-dimensional Electrochemistry Modeling Tool for Planar Solid Oxide Fuel Cell Stacks," *J. Power Sources*, vol. 196, no. 6, p. 3204, 2011.
- [11] EG&G Technical Services, *Fuel Cell Handbook*, 7th Edition, 2004.
- [12] K. Huang and J. Goodenough, *Solid Oxide Fuel Cell Technology: Principles, Performance and Operations*, Woodhead Publishing, 2009.
- [13] National Energy Technology Laboratory, "QGESS: CO₂ Impurity Design Parameter," DOE/NETL, March 2012.
- [14] J. C. Hensley, Ed., *Cooling Tower Fundamentals*, Mission, Kansas: The Marley Cooling Tower Company, 1985.
- [15] "Top 20 Shipping Container Home Designs and their Costs 2018," [Online]. Available: <https://www.24hplans.com/top-20-shipping-container-home-designs-and-their-costs>. [Accessed August 2018].
- [16] National Energy Technology Laboratory, "The Impact of Scale-up and Production Volume on SOFC Manufacturing Cost," DOE/NETL, April 2, 2007.

- [17] National Renewable Energy Laboratory, "U.S. Solar Photovoltaic System Cost Benchmark: Q1 2017," NREL, September, 2017.
- [18] National Energy Technology Laboratory, "QGESS: Cost Estimation Methodology for NETL Assessments of Power Plant Performance," DOE/NETL, April 2011.
- [19] S. Vora, *SECA Program at Siemens Westinghouse*, Siemens Westinghouse, 2005.
- [20] National Energy Technology Laboratory, "Techno-Economic Analysis of Natural Gas Fuel Cell Plant Configurations," DOE/NETL, April 8, 2015.
- [21] T. Grant, D. Morgan and A. Poe, "Four Basin Study Update," DOE/NETL, Pittsburgh, PA, September 30, 2013.
- [22] A. Tsang and C. Keeler, "E-STR™ Technology Development for Lignite Gasification," Gasification Technology Conference, Washington, D.C, November 3, 2010.
- [23] M. Huttinger, "Alkali Metal Catalyzed Water Vapor Gasification of Carbon using Mineral Catalyst Raw Materials," in *Carbon and Coal Gasification: Science and Technology*, NATO, Kluwer, 1985.
- [24] Exxon, "Dynamic Simulation of Exxon's Catalytic Coal-Gasification Process," November 1982.
- [25] E. Levy, H. Bilirgen, U. Levy, J. Sale and N. Sarunac, "Use of Coal Drying to Reduce Water Consumed in Pulverized Coal Power Plants," Energy Research Center, Lehigh University, Bethlehem, PA, January 2006.
- [26] National Energy Technology Laboratory, "Cost and Performance for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity," DOE/NETL, Revision 2, November 2010.
- [27] D. Denton, Interviewee, *Telephone communication re: James Black*. [Interview]. November 2010 DOE/NETL-2010/1397.
- [28] J. Black, "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity," DOE/NETL-2010/1397, November 2010. [Online]. Available: http://www.calgoncarbon.com/bulletins/TYPE_HGR.htm.
- [29] SFA Pacific for NETL, "Process Screening Analysis of Alternative Gas Treating and Sulfur Removal for Gasification - Revised Final Report," December 2002.
- [30] G. Agrawal, "Advances in Fuel Cell Blowers," 10th Annual SECA Workshop, DOE/NETL, Pittsburgh, PA, July 16, 2009.
- [31] National Energy Technology Laboratory, "Advanced Oxy-combustion Technology for Pulverized Bituminous Coal Power Plants," DOE/NETL, April 20, 2015.
- [32] V. White and R. Allam (Air Products and Chemicals), "Purification of Carbon Dioxide". USA and Europe Patent US 656914 and EP 1 953 486, 2013.

- [33] B. Ha (Air Liquide), "CO₂ Separation Apparatus Process for Oxy Combustion Coal Power Plants". USA Patent US 8,268,050, 2012.
- [34] S. Uysal, "Sensitivity and Techno-economic Analysis for Turbine Cooling Technologies for Higher Combined Cycle Efficiency," DOE/NETL, Pittsburgh, December 19, 2018.
- [35] White, C.W., Gray, D., Shelton, W., Weiland, N., and Shultz, T., "Technoeconomic Analysis of Utility-Scale Power Plants based on Indirect sCO₂ Brayton Cycle," DOE/NETL, Pittsburgh, September 25, 2017.
- [36] R. Allam, S. Martin, B. Forrest, J. Fetvedt, X. Lu, D. Freed, W. Brown, T. Sasaki, M. Itoh and J. Manning, "Demonstration of the Allam Cycle: An update on the development status of a high efficiency supercritical carbon dioxide power process employing full carbon capture," *Energy Procedia*, vol. 114, p. 5948–5966, 2017.
- [37] N. Weiland and C. White, "Performance and Cost Assessment of a Natural Gas-Fueled Direct SCO₂ Power Plant," National Energy Technology Laboratory, March 15, 2019.
- [38] National Energy Technology Laboratory, "QGEES: CO₂ Impurity Design Parameter," DOE/NETL, March 2012.
- [39] S. D. Vora, W. L. Lundberg and J. Pierre, "Overview of U.S. Department of Energy Office of Fossil Energy's Solid Oxide Fuel Cell Program," 2017.

Albany, OR • Anchorage, AK • Morgantown, WV • Pittsburgh, PA • Sugar Land, TX

www.netl.doe.gov

(800) 553-7681

