Analysis of Carbon Dioxide Capture Retrofit Options

Duke Edwardsport Integrated-Gasification–Combined-Cycle Plant

1021656

Analysis of Carbon Dioxide Capture Retrofit Options

Duke Edwardsport Integrated-Gasification–Combined-Cycle Plant

1021656

Technical Update, September 2011

EPRI Project Manager

R. Schoff

ELECTRIC POWER RESEARCH INSTITUTE
3420 Hillview Avenue, Palo Alto, California 94304-1338 ▪ PO Box 10412, Palo Alto, California 94303-0813 ▪ USA
800.313.3774 ▪ 650.855.2121 ▪ askepri@epri.com ▪ www.epri.com

DISCLAIMER OF WARRANTIES AND LIMITATION OF LIABILITIES

THIS DOCUMENT WAS PREPARED BY THE ORGANIZATION(S) NAMED BELOW AS AN ACCOUNT OF WORK SPONSORED OR COSPONSORED BY THE ELECTRIC POWER RESEARCH INSTITUTE, INC. (EPRI). NEITHER EPRI, ANY MEMBER OF EPRI, ANY COSPONSOR, THE ORGANIZATION(S) BELOW, NOR ANY PERSON ACTING ON BEHALF OF ANY OF THEM:

(A) MAKES ANY WARRANTY OR REPRESENTATION WHATSOEVER, EXPRESS OR IMPLIED, (I) WITH RESPECT TO THE USE OF ANY INFORMATION, APPARATUS, METHOD, PROCESS, OR SIMILAR ITEM DISCLOSED IN THIS DOCUMENT, INCLUDING MERCHANTABILITY AND FITNESS FOR A PARTICULAR PURPOSE, OR (II) THAT SUCH USE DOES NOT INFRINGE ON OR INTERFERE WITH PRIVATELY OWNED RIGHTS, INCLUDING ANY PARTY'S INTELLECTUAL PROPERTY, OR (III) THAT THIS DOCUMENT IS SUITABLE TO ANY PARTICULAR USER'S CIRCUMSTANCE; OR

(B) ASSUMES RESPONSIBILITY FOR ANY DAMAGES OR OTHER LIABILITY WHATSOEVER (INCLUDING ANY CONSEQUENTIAL DAMAGES, EVEN IF EPRI OR ANY EPRI REPRESENTATIVE HAS BEEN ADVISED OF THE POSSIBILITY OF SUCH DAMAGES) RESULTING FROM YOUR SELECTION OR USE OF THIS DOCUMENT OR ANY INFORMATION, APPARATUS, METHOD, PROCESS, OR SIMILAR ITEM DISCLOSED IN THIS DOCUMENT.

REFERENCE HEREIN TO ANY SPECIFIC COMMERCIAL PRODUCT, PROCESS, OR SERVICE BY ITS TRADE NAME, TRADEMARK, MANUFACTURER, OR OTHERWISE, DOES NOT NECESSARILY CONSTITUTE OR IMPLY ITS ENDORSEMENT, RECOMMENDATION, OR FAVORING BY EPRI.

THE FOLLOWING ORGANIZATIONS PREPARED THIS REPORT:

Electric Power Research Institute (EPRI)

Gas Processing Solutions, LLC

Process Power Plants, LLC

Sorensenergy

Syngas Consultants, LTD

This is an EPRI Technical Update report. A Technical Update report is intended as an informal report of continuing research, a meeting, or a topical study. It is not a final EPRI technical report.

NOTE

For further information about EPRI, call the EPRI Customer Assistance Center at 800.313.3774 or e-mail askepri@epri.com.

Electric Power Research Institute, EPRI, and TOGETHER…SHAPING THE FUTURE OF ELECTRICITY are registered service marks of the Electric Power Research Institute, Inc.

Copyright © 2011 Electric Power Research Institute, Inc. All rights reserved.

ACKNOWLEDGMENTS

The following organizations prepared this report:

Electric Power Research Institute (EPRI)
1300 West W.T. Harris Blvd.
Charlotte, NC 28262

Principal Investigators
R. Schoff
J. Phillips
G. Booras
N. Holt

Gas Processing Solutions, LLC
12 2nd Ave.
Isle of Palms, SC 29451

Principal Investigator
D. Kubek

Process Power Plants, LLC
6003 Parkis Mills Rd.
Galway, NY 12074

Principal Investigator
D. Todd

Sorensenergy
620 N. Muhlenberg St.
Allentown, PA 18104

Principal Investigator
J. Sorensen

Syngas Consultants, LTD
Top Floor, 14 Athol St.
Douglas, Isle of Man, IM1 1JA

Principal Investigator
C. Higman

This report describes research sponsored by EPRI.

EPRI acknowledges the support of David Julius and Frank Carchedi of Duke Energy in the preparation of this report.

This publication is a corporate document that should be cited in the literature in the following manner:

Analysis of Carbon Capture Retrofit Options: Duke Edwardsport Integrated-Gasification–Combined-Cycle Plant. EPRI, Palo Alto, CA: 2011. 1021656.

ABSTRACT

This report summarizes the results of a project supported by Duke Energy using tailored collaboration funds to study the potential impact to plant performance of retrofitted carbon dioxide (CO_2) capture on the Duke Edwardsport integrated-gasification–combined-cycle (IGCC) plant. The Duke Edwardsport IGCC plant is under construction and scheduled to begin operation in September 2012. Details on the project have been published in a 2010 Electric Power Research Institute (EPRI) report, *Duke Edwardsport Generic Design Specification Version 2* (1019668).

The study was conducted in 2009 and was based on the original design of the Edwardsport plant. Because the design evolved as the project progressed through detailed engineering, the base case performance used in this study is slightly different from that of current projections. Nevertheless, the relative impact of the levels of CO_2 capture retrofit studied in this project would likely be unchanged had the study been initiated after the design changes had been made.

The study focused on three levels of CO_2 capture retrofit: “skimming,” or removal of CO_2 from unshifted syngas; natural gas–combined-cycle equivalent CO_2 emissions (single-stage water-gas shift); and “full” CO_2 capture (two stages of water-gas shift). Power plant performance results are provided for each case in addition to discussion by the CoalFleet IGCC technical experts group of the plant systems that would require further evaluation when moving beyond conceptual design (as done here) through pre–front-end engineering design (FEED), FEED, and detailed engineering.

Keywords

IGCC
Gasification
Combined cycle
Edwardsport IGCC
CO_2 capture

EXECUTIVE SUMMARY

This report summarizes the results of a project supported by Duke Energy using tailored collaboration funds to study the potential impact to plant performance of retrofitted carbon dioxide (CO_2) capture (CC) on the Duke Edwardsport integrated-gasification–combined-cycle (IGCC) plant. The Duke Edwardsport IGCC plant is under construction and scheduled to begin operation in September 2012. Details on the project have been published in a 2010 Electric Power Research Institute (EPRI) report, *Duke Edwardsport Generic Design Specification Version 2* (1019668).

The study was conducted in 2009 and was based on the original design of the Edwardsport plant. Because the design evolved as the project progressed through detailed engineering, the base case performance used in this study is slightly different from that of current projections. Nevertheless, the relative impact of the levels of CO_2 capture retrofit studied in this project would likely be unchanged had the study been initiated after the design changes had been made.

This study contains analysis of three retrofit options for the Duke Energy Edwardsport IGCC plant related to various levels of CC. The study also includes analysis of the base plant design and its readiness for CC retrofit in addition to an investigation of CO_2 capture from unshifted syngas, syngas exiting a single stage of water-gas shift (WGS) reactors, and syngas exiting two stages of WGS reactors.

Project Assumptions

Analyses of the Edwardsport IGCC plant design and the three retrofit CC options were based on a normal operating condition (NOC) heat and material balance provided through the GE/Bechtel Alliance website using the design coal. The heat and material balance for the base case was used in this analysis because it is more representative of an annual average than the performance case is. Although Edwardsport is designed to operate on a range of coal qualities, this study did not attempt to evaluate any except the design coal.

Using these criteria, the EPRI model of the Edwardsport IGCC plant operating on Indiana #5 coal (25,930 kJ/kg [11,150 Btu/lb] higher heating value [HHV]) will produce 635 MWe (net) at a net plant heat rate of 9,408 kJ/kWhr (8,920 Btu/kWhr) or 38.3% HHV. As noted, the plant performance used in this project reflects an earlier stage of design for the Edwardsport IGCC plant. It is now estimated that the plant will produce 618 MWe (net) at a net plant heat rate of 9,823 kJ/kWhr (9,313 Btu/kWhr) or 36.6% HHV. The reasons for the change are explained in Section 3 of EPRI report 1019668, which is based on public testimony made by Duke Energy to the Indiana Utility Regulatory Commission.

Key Technical Results

Three retrofit CC cases were included in the analysis:

1. CC Case 1: “Skimming” or Removal of CO_2 from Unshifted Syngas
2. CC Case 2: NGCC Equivalent CO_2 Emissions (Single-Stage Water-Gas Shift)
3. CC Case 3: “Full” CC (Two Stages of Water-Gas Shift)

CC Case 1, or the “skimming” case, is designed to remove CO_2 from syngas produced by the gasifier without inclusion of water-gas shift reactors in the syngas train. The amount of CC possible depends on the coal quality, which impacts coal slurry concentration and gasifier efficiency. The slurry-fed, single-stage entrained-flow GE gasifier at Edwardsport uses a mid-quality bituminous fuel, Indiana #5. This is mixed with water to create a pumpable slurry and produces a saturated syngas at the exit of the water scrubber containing 20.5% by volume CO_2 on a dry basis, which represents ~35% of the carbon in the feedstock. Designing the Selexol unit for aggressive yet economically reasonable levels of CO_2 capture results in 29% CC in Case 1, which reduces the CO_2 emissions of the plant from 807 kg/MWe (1,780 lb/MWe) to ~590 kg/MWe (~1,300 lb/MWe) net. The installation of CO_2 absorption, flash recovery, and compression equipment is required in this case.

CC Case 2, or the natural gas–combined-cycle (NGCC) equivalent emissions case, includes a single stage of WGS reaction on each of the two operating trains in the plant and the installation of CO_2 absorption, flash recovery, and compression equipment. A CO_2 capture level of 59% results in a reduction in CO_2 emissions from 807 kg/MWe (1,780 lb/MWe) to 363 kg/MWe (~800 lb/MWe) net.

CC Case 3, or the full capture case, includes two stages of WGS reactors on each of the two operating trains in the plant and the installation of CO_2 absorption, flash recovery, and compression equipment. A CO_2 capture level of 86% results in a reduction in CO_2 emissions from 807 kg/MWe (1,780 lb/MWe) to 132 kg/MWe (~290 lb/MWe) net. Because of sufficient water in syngas exiting the gasifier island, the thermal impact of the full carbon capture retrofit was minimized. Figure ES-1 compares the performance of each case and CO_2 emissions.

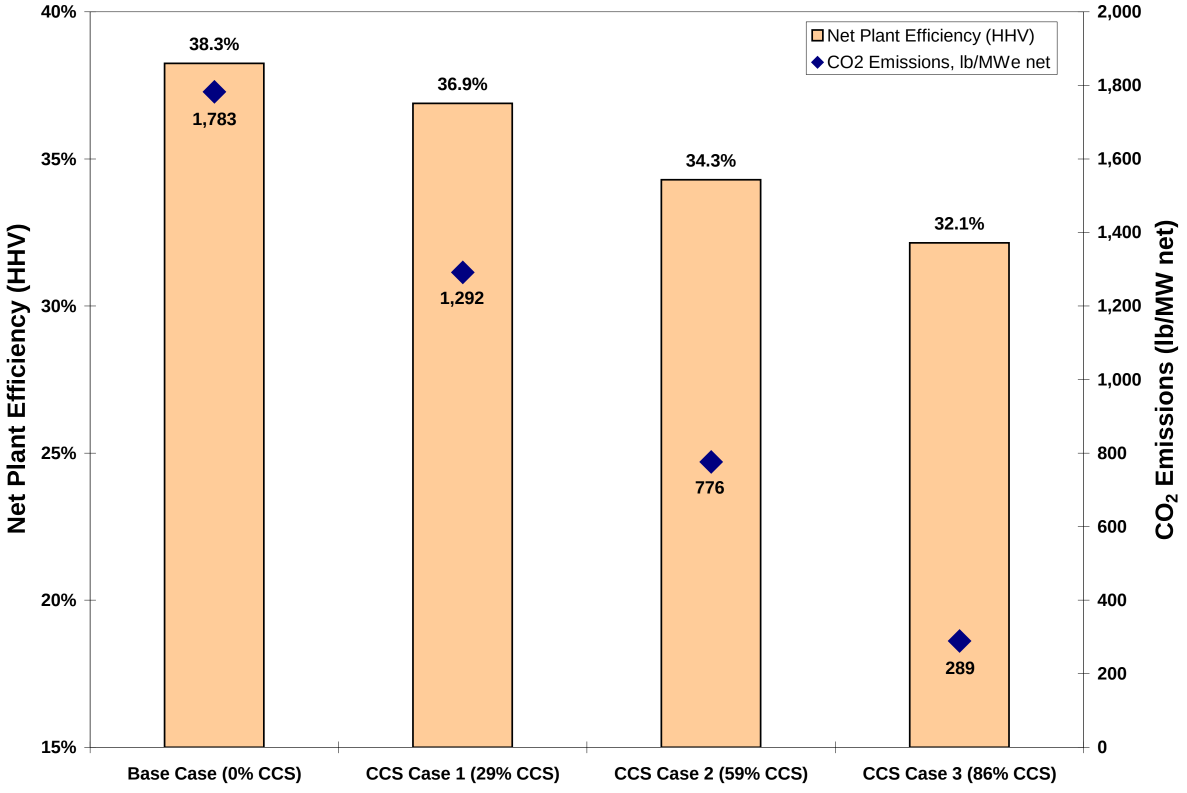


Figure 1
Case Performance and CO_2 Emissions

The inclusion of CO_2 capture for the Edwardsport IGCC plant is expected to have a positive impact on the plant's emissions profile. NOx emissions are expected to be maintained; SOx emissions should decrease because the inclusion of CO_2 absorption in the Selexol unit increases sulfur capture in that unit. Further, the capture of CO_2 reduces overall carbon emissions from the plant.

ACRONYMS

ABS	ammonium bisulfate
AGR	acid gas removal
ALS	ammonium lignosulfonate
ANSI	American National Standards Institute
API	American Petroleum Institute
ASCE	American Society of Civil Engineering
ASHRAE	American Society of Heating, Refrigerating and Air-Conditioning Engineers
ASL	site elevation above mean sea level
ASME	American Society of Mechanical Engineers
ASTM	ASTM International, formerly the American Society for Testing and Materials
ASU	air separation unit, an oxygen/nitrogen plant
BACT	best available control technology
barg	bars, gauge
BAT	Best Available Technology
BEC	bare erected cost
BFD	block flow diagram
BOP	balance of plant
Btu	British thermal unit
°C	temperature degrees Celsius
CC	combined cycle
CCF	levelized carrying charge factor
CCPI	Clean Coal Power Plant Initiative of the US DOE
CCS	carbon capture and storage (or carbon sequestration)
CEMS	continuous emissions monitoring system
CF	capacity factor
CEI	carbon emission intensity
CFR	Code of Federal Regulations
Cl	chlorine or chloride
CLSTK	converted lump-sum turnkey
CO	carbon monoxide
CO_2	carbon dioxide
CO_2e	100-year CO_2 equivalents; IPCC 2007 Global Warming Potential
COC	cycles of concentration
COD	chemical oxygen demand
COE	levelized busbar cost of electricity, often expressed in $/MWh
COI	chemicals of interest
COS	carbonyl sulfide
CT	acronym for combustion turbine, used interchangeably with gas turbine
CWS	coal/water slurry
DAF	dry and ash-free; synonym for MAF
DCS	digital control system
DGAN	diluent gaseous nitrogen
DI	de-ionized
DIPA	diisopropanolamine
DME	di-methyl ether
DOE	United States Department of Energy
dscf	dry standard cubic feet
dscm	dry standard cubic meter
EAF	equivalent availability factor
EHS	Environmental, health and safety
EIA (definition 1)	Electronic Industries Association

EIA (definition 2)	Energy Information Agency (of the U.S. DOE)
EIS	Environmental Impact Statement
EOR	enhanced oil recovery
EPA	United States Environmental Protection Agency
EPC	engineering procurement and construction contract or contractor
EPCM	engineering procurement and construction management
EPRI	Electric Power Research Institute
ºF	temperature, degrees Fahrenheit
FAC	flow accelerated corrosion
FC	fixed carbon
FCR	field change request
FD	forced draft (fan)
FDA	flash dryer absorber
FOF	forced outage factor
FOR	forced outage rate
fps	feet per second
FT	fluid temperature
FW	Foster Wheeler
GADS	Generating Availability Data System
GAN	gaseous nitrogen
GC	geological sequestration
GE	the General Electric Company
GEE	GE Energy
GEP	good engineering practice
GJ	gigajoules, i.e. 10^9 joules
GOX	gaseous oxygen
GT	acronym for gas turbine, a combustion turbine
GTAW	gas tungsten arc welding
GTC	Gasification Technologies Council
GTCC	gas turbine combined cycle
HAZOP	Hazard & Operability
H_2	hydrogen
H_2S	hydrogen sulfide
HDPE	high density polyethylene
HEI	Heat Exchange Institute Standards
Hg	mercury
HGI	Hardgrove Grindability Index, the relative hardness of the coal
HHV	higher heating value (heating value of fuel if heat of condensation of water vapor from its combustion products in a bomb calorimeter at 77ºF is included)
HI	Hydraulics Institute
HMI	Hoist Manufacturers Institute Standards
HMI	human/machine interface
HP	high pressure
HPSS	high pressure solids settler
HRSG	heat recovery steam generator
HSS	heat stable salts
HTHP	high temperature-high pressure particulate filter
HVAC	heating ventilation and air conditioning
ID	induced draft (fan)
IDC	Interest during construction
IEEE	Institute of Electrical and Electronics Engineers
IFR	interim final rule
IGCC	integrated gasification combined cycle (a type of electric power plant that uses syngas from a gasifier as fuel for a combustion turbine topping cycle and heat

	recovered from the combustion turbine exhaust and gasification process to produce steam for a steam-turbine bottoming cycle)
IGV	inlet guide vane
IOU	investor-owned utility
IP	intermediate pressure
IPP	independent power producer
IPPC	United Nations Intergovernmental Panel on Climate Change
I.R.	infrared radiation
ISA	Instrumentation, Systems, and Automation Society
ISBL	Inside Battery Limits
IOS	International Organization for Standardization
ISO	International Standards Organization
ITM	ion transport membrane, a technology for oxygen production, a type of ASU
K	temperature, degrees Kelvin
KBR	Kellogg, Brown and Root
K.O.	knockout drum
kW, kWe	kilowatt electric
kWt	kilowatt thermal
LAER	lowest achievable emission rate
LAR	liquefied argon
LBtu	low Btu content syngas
LCA	Life cycle analysis
LHV	lower heaving value (fictitious heating value of fuel if the heat of condensation of water vapor is ignored; i.e., if the water in the combustion products were assumed to remain in the vapor state in the bomb calorimeter)
LIN	liquefied nitrogen
LOI	loss on ignition
LORS	laws, ordinances, regulations, and standards
LOX	liquefied oxygen
LP	low pressure
LPG	propane gas from liquefied propane
LSTK	lump-sum turnkey
LTPD	long tons per day
MAC	main air compressor
MAF	moisture and ash-free; synonym for DAF
MCR	maximum continuous rating
MDEA	methyldiethanolamine
MDT	mean down time
MHI	Mitsubishi Heavy Industries, Ltd.
MM&V	measurement, monitoring, and verification
MMBtu	10^6 Btu (U.S. million Btu)
MP	medium pressure
MPTA	Mechanical Power Transmission Association
MSD	Master Scoping Drawing
MSS	Manufacturers Standardization Society
MTBF	mean time between failures
MTPD	metric tons/day
MW, MWe	megawatt electric, equivalent to 1000 kW or 10^6 watts
MWt	megawatt thermal
NA, N/A	not applicable
NAAQS	National Ambient Air Quality Standards
NCR	non-conformance report
ND	not detected
NEMA	National Electrical Manufacturers Association

NEPA	National Environmental Policy Act
NERC	North American Electric Reliability Corporation
NETL	DOE National Energy Technology Laboratory
NEPA	National Environmental Policy Act
NFPA	National Fire Protection Association
NGCC	natural gas-fired combined cycle
NO_x	nitrogen oxides
NPDES	National Pollutant Discharge Elimination System
NSPS	New Source Performance Standards
NSR	New Source Review
NYPA	New York Power Authority
O&M	operating and maintenance
OCA	Off-site Consequence Analysis
OEM	original equipment manufacturer
OEHI	Occidental of Elk Hills, Inc.
ORAP	Operational Reliability Analysis Program (registered trademark of Strategic Power Systems Inc.)
OSHA	Occupational Health and Safety Administration
OUC	Orlando Utilities Commission
P&ID	piping and instrumentation diagram
Part.	particulate emissions
PC, pc	pulverized coal
PCD	particulate control device (filter)
PCFB	pressurized circulating fluidized bed
PDI	Plastic Drum Institute Standards
PEL	Permissible Exposure Limit
PFI	Pipe Fabrication Institute
PGM	partial gasifier module
PHA	process hazard analysis
PM	particulate matter
PPA	power purchase agreement
ppmv	parts per million by volume
ppmvd	ppmv, dry basis
ppmvw	ppmv, wet basis
ppmw	parts per million by weight
PRB	Powder River Basin
PSA	pressure-swing absorber
PSD	prevention of significant deterioration
psia	lb/square inch absolute (14.696 psia = 1 atm)
psid	lb/square inch difference, used for pressure drop
psig	lb/square inch gauge [(psia) – (local atmospheric pressure in psia)]
PSM	process safety management
PSSR	pre-startup safety review
PTC	Power Test Code
PVC	polyvinyl chloride
Q	heat
ºR	temperature, degrees Rankine
RBD	reliability block diagrams
RCRA	Resource Conservation and Recovery Act
RMP	the federal Clean Air Act Risk Management Program
RMS	root-mean-square
RO	reverse osmosis
S	sulfur content of fuel or feedstock
SAMA	Scientific Apparatus Manufacturers Association

SCADA	supervisory control and data acquisition
scf	standard cubic feet (or feet)
scfd, scf/d	standard cubic feet per day
SCGP	Shell Coal Gasification Process
SCOT	Shell-Claus off-gas treating unit
SCR	selective catalytic reduction (for NO_x control)
SGC	syngas cooler
SMACNA	Sheet Metal and Air Conditioning Contractors National Association
SNCR	selective non-catalytic reduction (for NO_X control)
SNG	synthetic natural gas
SO_2	sulfur dioxide
SO_x	sulfur oxides
SRU	sulfur recovery unit
SSPC	Society for Protective Coatings
SST	side-stream treatment
ST	steam turbine
STL	Short Term Exposure
STOMP	subsurface transport over multiple phases
STQ	screening threshold quantity
SWS	sour water stripper
syngas	from “synthesis gas”, the name given to gases of varying composition, usually containing carbon monoxide and hydrogen, that are generated from the gasification of coal, lignite, petroleum coke, heavy oils and other feedstocks
t	short ton (2,000 lbs)
T&D	transmission and distribution (electrical)
t/h, tph	short tons per hour (2000 lb/h)
t/y, tpy	short tons per year (2000 lb/y)
T_{250}	the temperature at which the slag viscosity is 250 poise
TAR	Transmission Access Review
TBD, tbd	to be determined
TBtu	a “tera Btu”, a “trillion (U.S.) Btu”; that is, 1 TBtu = 1 tera Btu = 10^{12} Btu
TCLP	toxicity characteristic leaching procedure
TECO	Tampa Electric Company
TEMA	Tubular Exchanger Manufacturers Association
TG	turbine-generator
TGT	tail-gas treatment
TLV	Threshold Limit Value
TOC	total organic carbon
ton	short ton (2000 lb)
tonne	metric tonne (1000 kg or 2205 lb)
TPC	total plant cost
TRIG™	Transport Integrated Gasification (KBR/Southern)
TS&M	Transportation, Storage & Monitoring (of CO_2)
UIC	Underground Injection Control
UDBS	User Design Basis Specification (this document)
UHC	unburned hydrocarbons
UL	Underwriters Laboratory
US, U.S.	United States
USACE	US Army Corps of Engineers
USC	ultra-supercritical
USD, US$	United States dollar
USDOE	United States Department of Energy
USDW	underground sources of drinking water
USEPA	United States Environmental Protection Agency

UTS	universal treatment standards
VAR	Volt-ampere reactive
VM	volatile matter
VOC	volatile organic compounds
VR	vacuum residue
WAG	water-alternating-gas flood process for water/CO_2 injection
WGDS	warm gas desulfurization system
WGCU	a warm gas desulfurization process
y, yr	year
ZLD	zero liquid discharge

CONTENTS

LIST OF FIGURES

LIST OF TABLES

1

DESIGN BASIS

This section includes selected parts of the GE scope process design basis for the engineering design of the nominal 630 MW Edwardsport IGCC Plant for Duke Energy. It specifies the nominal and design conditions for the sub-systems of the plant. The Edwardsport IGCC plant is designed to gasify a range of coals from Indiana and to operate over a wide range of ambient conditions. This pre-FEED study was performed only on the design fuel for the normal operating condition (NOC) at average ambient conditions that relates to the base load maximum capacity case.

The Edwardsport IGCC plant will be designed to gasify coal to produce syngas in sufficient quantity to fully load two GE 7F Syngas IGCC combustion turbines configured in combined cycle on the performance coal at ISO conditions. The nominal net power output is expected to be approximately 630 MW when the design coal is used. The objective of the project is to develop a design that is cost competitive with Rankine-cycle pulverized coal power generation while meeting more stringent emission standards, providing high overall plant availability, and high efficiency.

1.1 Coal

The Duke Edwardsport IGCC Plant will be designed to be operable over a defined coal range. Indiana #5 Coal has been identified as the performance coal / NOC coal. Table 1-1 shows the coal analysis that was used for plant performance evaluations in this study.

If fluxant is determined to be required for a particular fuel feed, the fluxant and coal are assumed to be blended offsite, as no blending capacities are included in the Plant design. The total of any blended fuel should not exceed the below listed coal range.

The Duke Edwardsport IGCC Plant will include two coal grinding trains. These two trains will share two common coal bins. Each train is comprised of a weigh feed conveyor, grinding mill, mill discharge tank and agitator. Each train will be designed to process 55% of the design coal feed rate.

There will also be two trains of slurry storage. Each train will be designed for 50% of the design case slurry rate with an eight-hour working capacity.

Table 1-1
Edwardsport Performance Coal Analysis – Indiana #5

Moisture	As-Received	Dry
Proximate Analysis (wt. % As-Received)		
Moisture	13.50	0.00
Ash	10.00	11.60
Fixed Carbon	35.90	41.50
Volatile Matter	40.60	46.90
Total	100.00	100.00
Ultimate Analysis (wt. % As-Received)		
Moisture	13.5	0.00
Carbon	61.00	70.50
Hydrogen	4.60	5.30
Nitrogen	1.30	1.50
Chlorine	0.00	0.00
Sulfur	3.70	4.30
Ash	10.00	11.60
Oxygen (by difference)	5.90	6.80
Total	100.00	100.00
Heating Value		
HHV kJ/kg (Btu/lb) as-received	26,205 (11,268)	30,296 (13,027)
LHV kJ/kg (Btu/lb) as-received	24,898 (10,706)	28,968 (12,465)

1.2 Technical Data

Common technical data include:

- The design fuel for the Edwardsport IGCC plant is Indiana #5 coal. The characteristics and analyses of this coal are presented in Table 1-1.
- Backup fuel is natural gas (NG).
- The main products and by-products of the plant are the following:

<u>Electric Power:</u>

Voltage:	345 kV
Frequency:	60 Hz

<u>Carbon Dioxide</u>:

For the cases where CO_2 capture is included, the design specification at the plant battery limits (B.L.) are the following:

Phase: Supercritical
Pressure: 15.18 MPa (2,200 psig)
Temperature: 30°C (86°F)
Purity: 96.0 % wt (min – Industrial Grade)
H_2S content: 100 ppmv (max)
CO content: 0.1 % wt (max)
Moisture: 100 ppmv (max)
N_2 Content: 40,000 ppmv (max)
Hydrocarbons: 20,000 ppmv (max)
Oxygen: 100 ppmv (max)
Glycol: 0.3 gal/MMcf

The carbon recovery level varies on a case-by-case basis up to 86% with respect to the carbon entering the plant, based on the maximum practical removal percentage resulting from optimized CO_2 capture plant components. Residual carbon in the gasifier by-products is not considered as carbon emission.

In addition to a full capture case, two sensitivity cases were performed: one with capture of the CO_2 in the syngas outlet from the gasifier (no shift) and one with a single CO shift stage plus sufficient capture in the AGR to meet CO_2 emissions levels similar to a state-of-the-art 2x1 F-class gas turbine-based natural gas combined cycle (NGCC).

1.3 Environmental Conditions

Information related to the Duke Edwardsport air permit for the base plant excluding any CO_2 capture, is included in EPRI report 1019668 for reference. Section 1.3.1 provides some indicative environmental data based on the EPRI User Design Basis Specification (UDBS) environmental design 2.

1.3.1 EPRI UDBS Environmental Design 2

The EPRI UDBS includes environmental profiles that reflect emissions targets relating to IGCC power plants, including those that include CO_2 capture. Environmental design 2 is the design for the emissions that would be typical for a plant burning the bituminous coal, with SCR for NOx control. The emissions criterion for environmental design 2 is discussed below, and summarized in Table 1-2. This information is included for informational purposes only and does not reflect the Duke Edwardsport IGCC plant air permit data.

Table 1-2
Guarantee Emissions Level Chart for Environmental Design 2

Pollutant	**Unit of measurement**	**Data**
NOx[1]	ppmv, dry	3.5
SOx[1]	ppmv, wet	2.0
PM (filterable – front half of sampling train)	kg/hr (lb/hr)/GT	8.2 (18)
Total PM (including condensables – back half of sampling train)	kg/hr (lb/hr)/GT	16.8 (37)
CO	ppmv, dry	25
Unburned Hydro Carbons (UHC) as CH_4	ppmv, wet	7
Volatile Organic Compound (VOC)[1]	ppmv, wet	1.4
Mercury (Hg)	-	No more than 10% of mercury in coal as air emission
Slip Ammonia[1]	ppmv, dry	5.0

Notes:
1. Referenced to 15% O_2.

- NO_x – When firing syngas or hydrogen rich fuel, HRSG stack emissions limited to 3.5 ppmvd or lower in the flue gas (at 15 percent O_2 dry) are based on the use of selective catalytic reduction (SCR). The gas turbine design limit with syngas or hydrogen rich fuel and nitrogen dilution or fuel humidification is 15 ppmvd (at 15 percent O_2 dry). Back-up natural gas emissions are limited to 3.5 ppmvd (at 15% O_2 dry) with use of a diluent, such as steam, and with SCR
- SO_2 – Sulfur emission levels (H_2S + COS) designed for 2 ppmvw SO_2 in the flue gas to avoid excess ammonium sulfate / bisulfate deposits in the HRSG downstream of the SCR, to meet a requirement of maximum one outage for washing /year. This corresponds to approximately 15 ppmvw or less in the fuel outlet from the AGR, depending on the gas turbine type, diluents, etc.
- Particulate Matter (PM)
  - Filterable – Particulate matter from the front half of the sampling train without condensables, limited to 8.2 kg/hr (18 lbs/hr)/GT. In this case, the potential to emit additional PM may occur due to the use of SCR, which is likely to produce additional ammonium bisulfate and sulfate particles in the exhaust gas stream.
  - Total PM – With condensables included from the back half of the sampling train, limited to less than 16.8 kg/hr (37 lbs/hr)/GT.
  - CO limited to 25 ppmvd at actual %O_2, without CO catalyst.

  - UHC as CH_4 limited to 7 ppmw.
  - VOC limited to 1.4 ppmw, referenced to 15%O_2.
  - Mercury air emission no more than 10% of mercury in coal using sulfur- impregnated activated carbon upstream of the AGR process.
  - Slip Ammonia- 5 ppmvd.
- Water Discharge.
- Non-Hazardous Dry Solid Waste Disposal. The plant equipment and facilities are designed so that all ash or slag from the gasification facility and sludge produced from water treatment are disposed of off-site in accordance with the non-hazardous waste disposal guidelines of the Resource Conservation and Recovery Act (RCRA). Appropriate temporary onsite storage facilities are specified in accordance with the guidelines.
- Hazardous Dry Solid Waste Disposal. The spent carbon used for Mercury removal is considered a hazardous waste and accordingly provision is made for its disposal in accordance with this classification.
- Noise Limitations. In-plant noise levels not exceed 90 dBA for an 8-hour exposure. Plant perimeter noise levels not exceed 65 dBA during the day and 55 dBA during the night.

2

PROCESS DESCRIPTIONS

The Edwardsport IGCC plant uses GE's gasification technology to convert coal feedstock into syngas, which is then cleaned and burned in a combined cycle power plant to produce electricity.

The grinding and slurry feed section includes coal handling systems to meter, grind, and slurry the coal for feed to the gasifier. Coal slurry reacts with oxygen from the air separation unit (ASU) to produce syngas. Facilities are included to scrub particulates from the syngas as well as remove coarse and fine slag from the scrubbing water.

Facilities are designed with the necessary provisions for the addition of carbon capture equipment. The site arrangement has been developed assuming that carbon capture equipment will be added.

Wet raw syngas from the gasification and scrubbing section is cooled and sent to the acid gas removal (AGR) section. The clean syngas is then sent to the power block section as fuel for the gas turbines. Nitrogen from the ASU is sent to the combustion turbines in the power block section for use as diluent.

The power block uses two dual-fueled GE 7F Syngas gas turbine generators (GTGs) to generate electricity. Steam generated within the gasification block is combined with steam from the power block to generate additional electricity in a model G13 steam turbine generator (STG). The steam conditions used in the STG are 11.6 MPa/557°C/544°C (1,680 psig/1,035°F/1012°F). The LP turbine section employs two dual-flow units with 87.6 cm (34.5 in) last stage buckets, with the LP units exhausting to condensers operating at different pressure (4.6 cm [1.8 in] Hga and 3.6 cm [1.4 in] Hga).

2.1 Coal Grinding and Slurry Preparation System and Recycle Solids Mixing and Slurry Run Tanks

In the coal grinding section, coal is conveyed from storage to feed bins via belt conveyors on a one shift per day basis. Weigh feeders weigh and regulate the coal fed to the grinding mill. Chemical additives from the slurry additive tank are fed to the grinding mill by the slurry additive feed pump. The slurry additive system is common to both coal grinding trains.

The coal is converted into slurry in the grinding mill. The grinding mill, which is rod-type with an overflow discharge, grinds the coal to the required particle size distribution. After discharge from the grinding mill, the slurry passes into the mill discharge tank. Slurry from this tank is then pumped into the slurry run tank.

Makeup water and recycle fines from the coarse and fine slag handling sections are mixed in the recycle solids tank and fed to the grinding mill to slurry the coal.

There are two slurry run tanks, (one per coal grinding train). The recycle solids tank is common to both coal grinding trains. Slurry from the grinding section is fed to the slurry run tanks via the mill discharge pumps. The slurry run tanks have feed capability from either grinding mill and feed the slurry charge pumps that feed the gasifiers.

A sump is located below grade within the grinding and slurry preparation section to handle drains and spills in this area. Water collected in the sump is reused in the slurry process. This sump is common to both the coal grinding trains and the slurry preparation trains.

2.2 Gasification, Radiant Syngas Cooling and Syngas Scrubbing

There are two gasification trains. Each train is a radiant-only gasifier type, with high-pressure steam production, followed by cooling by a direct water quench. Each train is designed to process 50 percent of the design case syngas production.

The gasifier is a refractory-lined vessel capable of withstanding high temperatures and pressures. The coal slurry is pumped from the slurry run tank to the gasifier by the slurry charge pump. This high-pressure metering pump supplies a steady, controlled flow of slurry to the feed injector. The slurry and high-pressure oxygen from the ASU react in the gasifier at high temperatures (approximately 1,200°C [2,500 °F]) and in a reduced oxygen environment, to produce syngas.

Syngas consists primarily of hydrogen (H_2), carbon monoxide (CO), water vapor (H_2O), and carbon dioxide (CO_2). The syngas also contains small amounts of hydrogen sulfide (H_2S), carbonyl sulfide (COS), methane (CH_4), argon (Ar), ammonia (NH_3), and nitrogen (N_2). Ash, which was present in the coal, and unconverted carbon form a liquid melt called slag.

The tip of the feed injector is protected from the high temperatures in the gasifier by a water jacket and cooling coils through which cooling water is continuously circulated. The feed injector cooling water is supplied by the feed injector cooling water system. The water flows through the feed injector water jacket and cooling coil and returns to the feed injector cooling water system.

There are two trains of radiant syngas cooling and syngas scrubbing. Each train is designed to process 50 percent of the design case syngas production.

From the gasifier vessel, the hot syngas and slag from the reaction chamber flow down into the radiant syngas cooler (RSC). The RSC is a high-pressure steam generator equipped with a water wall to protect the vessel shell. Heat is transferred primarily by radiation from the hot syngas to the boiler feedwater circulating in the water wall.

The high-pressure steam flows to the syngas cooler high pressure steam drum for liquid disengagement, and the dry steam flows to the high-pressure steam header. Liquid is pumped back to the RSC by the syngas cooler high-pressure circulation pump.

In the RSC system, the raw syngas exiting from the radiant section is first cooled by direct contact with water and then sent to the syngas scrubber for cooling, condensation of water vapor, and removal of particulates by scrubbing with water. The saturated raw syngas from the direct quench is fed to the scrubber nozzle, where it is mixed with a portion of the syngas scrubber bottoms stream to wet the entrained solids so they can be removed in the syngas scrubber. The syngas scrubber circulating pump sends some of the syngas scrubber bottoms to the scrubber nozzle. The syngas scrubber RSC quench pump sends the syngas scrubber bottoms to the bottom of the RSC vessel to cool the gas and solidify the molten slag. The water also wets the slag solids and assists with removal to the lockhopper.

The syngas from the overhead of the syngas scrubber is routed to the low temperature gas cooling (LTGC) section.

Process condensate is pumped to the syngas scrubber from the LTCG 0.5 MPa (75 psi) steam generator knockout (KO) drum in the LTGC section. Some of this water is pumped to the syngas scrubber trays to wash the syngas. The remaining process condensate flows into the syngas scrubber to maintain liquid level in the vessel. The syngas scrubber bottoms contain solids removed from the syngas. As previously described, a portion of the scrubber bottoms stream is used to provide the water for contacting in the nozzle scrubber and for recycle back to the RSC bottoms.

2.3 Coarse Slag Handling

The coarse slag handling section removes coarse solid material from the gasifier. These solids are made of ash and unconverted coal components that exit the gasifier in the solid phase. There are two trains of coarse slag handling equipment. Each train is dedicated to one gasification and scrubbing train with the ability to be cross connected in times of equipment outage and is designed to process 50 percent of the maximum slag production and 50 percent of the maximum fines production.

Coarse slag exiting the bottom of the RSC is crushed by the slag crusher and flows into the lockhopper. Flow into the lockhopper is assisted by the lockhopper circulation pump, which takes water from the top of the lockhopper and returns it to the liquid sump in the RSC vessel. After the solids enter the lockhopper, the particles settle to the bottom. Thus, the lockhopper acts as a clarifier by separating solids from the water.

Solids collection is performed using a batch process. When the solids collection period is complete, the lockhopper is isolated from the RSC and depressurized. The solids that have accumulated in the lockhopper are water-flushed into the slag sump. When the water flush is finished, the lockhopper is filled with water and pressurized. Then, the next solids collection period begins.

In the slag sump, slag settles onto a submerged slag drag conveyor, which separates the slag from the water. The slag is washed and passed over a screen, which allows surface water to drain. The coarse slag conveyor then transfers the slag to owner’s material handling area. The water removed from the slag is pumped by the slag sump pump to the recycle solids tank in the grinding and slurry preparation section and to the vacuum flash drum in the black water flash system.

Water used to flush collected solids from the lockhopper is supplied to the lockhopper flush drum from the ammonia stripper bottoms pump, from cooled low pressure boiler feed water, and from grey water.

2.4 Black Water Flash

There is a single train to handle black water, except for the LP flash drums, which are configured as two parallel trains, each sized for 50 percent of the maximum design flow.

In this system, the black water from the sumps of the two radiant syngas coolers goes to the LP flash drum. The pressure is let down to 0.2 MPa (30 psig) through a tungsten-carbide-lined letdown valve at the inlet to the LP flash drum. At this pressure, a significant amount of steam and virtually all dissolved gases flash out of the black water.

The overhead of the LP flash drums is cooled in the LP flash overhead exchanger and flows to the condensate ammonia stripper. Grey water pumped from the grey water tank by the HP grey water pump to the syngas scrubbers is heated in this exchanger. The bottoms stream from the LP flash drum is let down into the vacuum flash drum. Any remaining dissolved gases flash out of the black water in this drum. The bottoms stream from the vacuum flash drum is pumped to the two parallel settler tanks in the fine slag handling section.

The overhead vapor from the vacuum flash drum enters the vacuum flash KO drum. The vapor from this drum is cooled in the vacuum flash overhead condenser using cooling water, and the condensate is returned to the vacuum flash KO drum. The bottoms stream from this drum is pumped by the vacuum flash condensate pump to the grey water tank. The uncondensed vapor is sent to the sour gas vacuum pump.

The vacuum in the vacuum flash drum is produced by the sour gas vacuum pump. Any net liquid condensed from the vapor stream is removed in the vacuum pump KO drum (included as part of the vacuum pump unit package) and sent to the grey water tank. The vapor discharge from the sour gas vacuum pump is sent to the sulfur recovery unit.

2.5 Fine Slag Handling and Grey Water Handling

There is one fine slag handling equipment train, which contains two parallel solid settler tanks, each equipped with settler bottoms pumps.

In the fine slag handling system, the bottoms of the vacuum flash drum flow to the settler tanks, where the solids are concentrated. The settler tanks contain a slow-moving rake that keeps the concentrated solids moving toward the bottom outlet. The overflow water from the settler (grey water) flows to the grey water tank. The settler bottoms are pumped to the grinding section to recycle fines. A small amount of anionic and/or cationic flocculent can be added upstream of the settler tank to improve the concentration efficiency of the settler.

There is one grey water handling equipment train. The water in the grey water tank is mostly free of particulates. The majority of the grey water is sent by the LP grey water pumps to the grey water blowdown pretreatment section to remove remaining particulates greater than 2 microns before it travels to the deep well injection system.

Grey water from the tank is also sent by the HP grey water pumps through the LP flash overhead exchanger to the syngas scrubbers as mentioned in the black water section above.

2.6 Condensate Ammonia Stripper

One condensate ammonia stripper train is included.

This section provides process condensate stripping to prevent the build-up of ammonia in the system. The overhead stream of the LP flash drum is used to strip the process condensate coming from the low temperature gas cooling section. The overhead reflux of the stripper acts as the cooling duty needed to condense out excess water vapor for the LP flash overhead of train 1 and the LP flash overhead of train 2. The stripped condensate is then pumped to the lockhopper flush drum after cooling.

The ammonia stripper overhead vapor is sent to the sulfur recovery unit (SRU).

2.7 Low Temperature Gas Cooling System

There are two LTGC trains, designed for 50 percent of the design case syngas production. This section is divided into steam generation and trim cooling equipment.

First, the scrubbed syngas from the syngas scrubber is cooled and the heat is recovered by heating water from the syngas saturator and generating LP steam. The condensate that falls out of the syngas through these exchangers is knocked out in the LTGC LP steam generator KO drum and pumped back to the syngas scrubber.

A COS hydrolysis reactor is located downstream of the LTGC LP steam generator KO drum to reduce the COS content of the syngas by 95 percent vol. The syngas entering the catalyst bed must be superheated to 10°C (50°F) above its dew point temperature to ensure no liquid moisture is present. This is achieved by heating the syngas in the COS preheater upstream of the reactor.

Following the COS hydrolysis reactor, the heat remaining in the syngas is recovered by heating the syngas saturation circulation water and the steam condensate. Finally, the remaining heat in the syngas is removed in the trim cooler using cooling water. Most of the water in the syngas is condensed and knocked out in the trim cooler KO drum following the trim cooler.

An upstream jacketed pipe for preheating and a mercury guard drum are provided to adsorb most of the mercury present in the syngas that is present in the coal feedstock.

Part of the condensate blowdown from the trim cooler KO drum is sent through the condensate ammonia stripper bottoms exchanger to the condensate ammonia stripper for ammonia removal while the remainder is recycled by the trim cooler KO pumps to the LTGC LP steam generator KO drum.

2.8 Acid Gas Removal System and Transient Emissions Handling System

This section briefly describes UOP's typical Selexol acid gas removal process. The details of this design are provided by UOP with further detailed descriptions.

UOP's Selexol process is used to remove acid gas from syngas. Scrubbed syngas from the mercury removal drum in the LTGC train is cooled in the feed/product exchanger. The syngas then enters the bottom of the H_2S absorber tower and contacts the lean solvent as it rises through the tower. The treated gas exits the top of the tower and is heated by the incoming solvent from the Selexol stripper in the pre-saturator drum. The treated gas is then heated further by the feed/product exchanger before going to the syngas saturation section.

The rich solvent contains dissolved H_2S, CO_2, and lesser amounts of other gases, including CH_4, N_2, CO, and COS. The rich solvent leaves the bottom of the H_2S absorber and is warmed by exchanging heat with the warm stripped solvent from the stripper bottoms. The rich solvent then flows into a flash drum, where some of the dissolved gases are flashed off. A substantial portion of the CO_2 and H_2S is flashed since Selexol is a selective physical solvent, and the amount of dissolved gas directly depends on the gas partial pressure of each component of the gas.

Gas flashed from the flash drum goes to the CO_2 recycle compressor for recycling upstream in the process. Flashed solvent from the flash drum goes to the stripper, where the remaining H_2S and CO_2 in the solvent are stripped by vapor rising through the tower. Heat is provided by the stripper bottoms reboiler using LP steam.

The overhead gases from the stripper go to an air-cooled condenser and reflux drum. The net overhead vapor stream is sent to the SRU. The stripped lean solvent is pumped through the lean/rich exchangers and the lean/treated gas exchanger to the top of the absorber. This section provides three major outlet streams. A clean syngas stream is sent as fuel to the power block section for saturation and then combustion. A concentrated acid gas stream is sent to the SRU. A CO_2 recycle stream is sent to the CO_2 recycle compressor.

In the event of an SRU trip, HP compressor trip and shutdowns (depressuring from 1.73 MPa [250 psig] to 0.7 MPa [100 psig]) the feed to the respective units can be diverted to the transient emissions handling system. The system comprises of a low-pressure absorber where H_2S from the flash gas, acid gas or syngas is removed by contacting the gas with solvent before flaring. The rich solvent is stored in a drum, from where it is fed to the main stripper to regenerate the solvent. The rich solvent drum has a capacity to store rich solvent for a period of one hour during a HP CO_2 compressor trip scenario, which is the design case.

2.9 CO_2 Recycle Compressor

Two 50 percent CO_2 recycle compressors are included.

The CO_2 recycle stream from the acid gas removal unit and tail gas recycled from the TGU are compressed by the CO_2 recycle compressors and recycled upstream.

2.10 Sulfur Recovery System and Tail Gas Unit

The two SRUs, the tail gas unit (TGU), and the thermal oxidizer (TO) are referred to collectively as the sulfur block.

The SRUs are configured as two parallel trains, each sized to handle 50 percent of the design feed. The primary feed to the SRUs is the acid gas from the acid gas removal unit. Secondary feeds to the SRUs are the condensate ammonia stripper overhead product and the black water flash vacuum pump discharge. These secondary feeds are referred to as ammonia acid gas.

The SRU design is based on Claus sulfur recovery technology. Acid gas and ammonia acid gas are fed to a two-chamber reaction furnace. The ammonia acid gas and some of the acid gas are fed to the burner in the first chamber. Oxygen for the combustion reaction is supplied from a combination of ambient air from an air blower and LP oxygen from the air separation unit (ASU). The LP oxygen flow rate is adjusted so that the air delivered to the burner contains 50 percent vol. oxygen.

The first chamber of the reaction furnace converts NH_3 to N_2. The effluent from the first chamber is mixed with the remaining acid gas in the second chamber, where some of the H_2S in the feed streams is converted to SO_2. Some of the H_2S and SO_2 react to form elemental sulfur.

The effluent from the second chamber of the reaction furnace is cooled in a waste heat boiler that generates medium pressure steam at approximately 600 psig. The effluent is further cooled in the first condenser by generating low pressure steam at approximately 75 psig.

Liquid sulfur is condensed during the cooling process and separated from the gas at the outlet end of the first condenser. The condensed sulfur drains to the sulfur pit through a steam-jacketed sulfur seal pot.

The cooled gas from the first condenser flows to the first reheater and is reheated using medium pressure steam. The reheated gas flows downward through the first-stage converter, where more sulfur is formed from the reaction of H_2S and SO_2 over a fixed bed of catalyst.

The process gas leaving the first-stage converter enters the second condenser and is cooled by generating low pressure steam. Liquid sulfur is condensed during the cooling process and separated from the gas at the outlet end of the second condenser. The condensed sulfur drains to the sulfur pit through a steam-jacketed sulfur seal pot.

The cooled gas from the second condenser flows to the second reheater and is reheated using medium pressure steam. The reheated gas flows downward through the second-stage converter, where more sulfur is formed.

The process gas leaving the second-stage converter enters the third condenser and is cooled by generating 15 psig steam (SLL). Liquid sulfur is condensed during the cooling process and separated from the gas at the outlet end of the third condenser. The condensed sulfur drains to the sulfur pit through a steam-jacketed sulfur seal pot.

The SRU tail gas leaves the third condenser and is routed to the TGU.

The sulfur pit collects the condensed sulfur from all three condensers. The sulfur pit floor is lined with low pressure steam heated steam coils to maintain the sulfur in a liquid state. The head space in the sulfur pit is continually swept with ambient air using a low pressure steam-powered eductor on the vent. The vent is sent to the thermal oxidizer (TO) for disposal. The sulfur is pumped out of the sulfur pit to a truck-loading rack.

The TGU is configured as a single train sized to handle 100 percent of the design flow. In the TGU, unconverted sulfur in the SRU tail gas is reduced to H_2S for recycling back to the gasification process. Clean syngas is the source of reducing gas. SRU tail gas is mixed with a small amount of clean syngas and preheated with medium pressure steam before passing through a catalyst bed in the hydrogenation reactor, where the SO_2, COS, and any other sulfur compounds are converted to H_2S. Reactor effluent is cooled in the reactor effluent cooler by generating low pressure steam. This is followed by a second cooling step in the contact condenser tower, a pump-around direct-contact cooler. The pump-around water is cooled by cooling water in the contact condenser water cooler, and a bleed stream is taken off the tower bottoms to maintain water inventory. The pressure of the overhead stream from the contact condenser tower is boosted by the tail gas blower and is recycled back to the gasification area via the CO_2 recycle compressors.

The TO provides a sink for SRU tail gas and/or TGU tail gas if either of these streams cannot be dispositioned normally. The vent from the SRU sulfur pits is also continuously routed to the TO. The TO is a natural gas-fired, natural-draft device designed to combust the various components in these streams prior to releasing them to the atmosphere.

2.11 Syngas Saturation

The clean syngas from the acid gas removal unit is contacted with heated water in the syngas saturator column to add moisture content to the syngas. The saturated syngas is then heated through the syngas performance heater using HP boiler feedwater and is then sent to the clean syngas scrubbers to knock out any remaining liquids. The saturated clean syngas is then routed to the GT fuel control skids.

The water used to saturate the syngas is a circulating loop of water that is heated in the LTGC section and then sent to the syngas saturator column to contact the dry syngas. The circulating water is moved by the syngas saturator circulation pumps at the bottom of the column. The makeup source for the circulating water is demineralized water.

2.12 Deep Well Injection Pretreatment

Grey water is transferred by pumping it through a cooling heat exchanger. A tank functions as the grey water feed tank for subsequent filtration before deep well injection and as a diversion/storage tank during non-filtration periods. Permanganate can be added between the heat exchanger and the feed tank to oxidize soluble iron for precipitation and removal. The feed tank is mixed to keep influent solids or precipitated solids in suspension. Grey water from the feed tank is pumped to one of two pre-coat candle filters (each rated at 100 percent of flow) for initial filtration and removal of solids. One pre-coat tank is used to pre-coat either candle filter. The candle filters operate on a cycle consisting of candle cloth pre-coating, filtration, vessel draining, and solids drying and discharge. During pre-coat and cake discharge periods, the water level in the tank may increase slightly. The processing filtration rate and cycle times balance the continuous influent flow to the feed tank.

The feed to a candle filter is positioned at the lower portion of the vessel to allow upflow of the incoming slurry, which is imperative to maintain fast-settling solids in a homogeneous suspension and ensure that uniform cakes are formed. Each candle core serves as a filtrate passage and supports the filter medium. The core is a bundle of perforated tubes contained in a coarse mesh cloth screen sleeve.

Once the pre-coating stage is completed, process slurry is pumped into the filter, a cake of captured solids is formed and retained on the candles, and filtrate flows out of the filter. The filtrate outlet from each row of candles is connected to a horizontal header. All headers deliver the filtrate through valves to a collecting manifold for further processing. The header arrangement allows the flow in each row of candles to be individually isolated to determine if a filter cloth is ever torn. A sight glass on each header enables the quality of the filtrate to be monitored.

Once the filtration cycle is completed, nitrogen is blown into the vessel and the slurry heel surrounding the candles is pushed and displaced downward by nitrogen until it reaches the lowest part of the candle stack. At this point, the remaining heel slurry is evacuated back to the feed tank by a dip pipe located at the very bottom of the filter to ensure the filter is empty of slurry. The heel slurry is transferred from the filter to the feed tank through a small transfer tank. After heel draining, nitrogen continues to pass through the cake until the captive moisture is reduced to a minimum and the cake is considered to be dry (in practical terms). At this point, the nitrogen pressure is released, the cake outlet is opened and a brief surge of nitrogen is applied through the outlet registers in a counter current direction, i.e. from the filtrate side thus, into the inside of the filter elements. The filter medium is momentarily billowed outward by this surge and vertical cracks form in the filter cake, which is immediately discharged. The cake outlet opening is interlocked with a pressure sensor to avoid opening until surging pressure is vented.

Filtrate leaving the candle filter enters a transfer tank. The contents of the transfer tank are pumped through a flow control valve (flow cascaded off of the tank level) through two polishing filters and two stage pH acidification adjustment (using hydrochloric acid) for storage in a filtered water storage tank before deep-well injection. The battery limit for grey water filtration ends at the storage tank.

2.13 Diluent Nitrogen Heating and Extraction Air Cooling

In the diluent nitrogen air exchanger, extraction air from the gas turbines preheats nitrogen from the ASU for use as diluent in the turbines. Extraction air then preheats the demineralized water used as makeup for the syngas saturation circulation water through the extraction air/makeup water exchanger. Finally, in the extraction air trim cooler, cooling water trim cools the extraction air before it is sent to the ASU.

There are two diluent nitrogen/extraction air exchangers, one extraction air/makeup water exchanger and one extraction air trim cooler.

2.14 Condensate and Feedwater Systems

The condensate system delivers water from the condenser hotwell to the condensate heaters (part of the low temperature syngas cooling system) via two condensate pumps. Makeup water for the condenser is provided from condensate water storage tank.

A stand-alone deaerator is provided to deaerate the condensate. The deaerator receives condensate from the condensate heater and LLP condensate return from the process block. The main source of the pegging steam is steam extraction from the steam turbine. LP steam from the LP superheater of the HRSG is the secondary source of deaerator pegging steam. All LLP steam from the process block is imported to the deaerator to augment pegging steam. Three LP boiler feedwater pumps supply feedwater from the deaerator to the HRSG LP economizer as well as to the LP boiler feedwater header in the process block.

Two combined HP/IP boiler feedwater pumps are provided for each HRSG. The HP/IP feed pump takes suction from the LP drum of the HRSG. The IP bleed from the HP/IP feedwater pump is used to supply boiler feedwater to the process block, attemperation water for the hot reheat steam, and natural gas performance heaters when firing natural gas. Discharge from the HP section of the HP/IP feedwater pump supplies water to the HP economizer and attemperation water for the main steam. HP feedwater for the SRU is supplied after the first pass of the HP economizer. HP boiler feedwater for the RSC and syngas heater is supplied from last HP economizer before the HP drum.

Condensate and boiler drum water conditioning chemicals are controlled by skid-mounted, packaged chemical injection unit(s).

System designs are based on the heat balance for the maximum calculated HRSG load, including additional requirements for HRSG blowdown.

The air removal system consists of vacuum pumps to both hog and hold the condenser vacuum. Each pump has the capacity to hold vacuum during operation, but both pumps are required to hog to startup vacuum within thirty (30) minutes.

2.15 Main Steam System

Main steam from the RSC is superheated in the HRSG for use by the HP section of each STG. Pressure loss has been allowed for the saturated steam line and backpressure control valve connecting the RSC and the HRSG main steam drums.

The reheater section of the HRSGs reheats the steam for re-entry into the IP section of the STG.

The IP section of the HRSG provides an economizer section for preheating water supplied to the IP boiler feedwater process header.

The HP economizer section of the HRSG preheats water that provides boiler feedwater for the SRU, RSC HP feedwater, and HP boiler feedwater for clean syngas heating.

The HRSG provides diluent steam for NOx control in the gas turbine when burning natural gas or co-firing natural gas, and as a backup of diluent nitrogen for syngas operation.

The LP steam section of the HRSGs comprises the LP drum and LP superheater. The LP steam drum serves as the water storage vessel and a water source for the boiler feedwater pumps associated with each HRSG. Excess steam generated in the LP evaporator section is superheated and admitted to the STG.

2.16 Gas Turbine Generators

Two 50 percent GE MS7001FB gas turbine generators (GTG) are included. These are also called 7F Syngas throughout this report.

The gas turbine was developed specifically for combined cycle applications and has features that enable it to be integrated with a gasification system. High combined cycle efficiency results from the high specific power of the gas turbine, which is achieved by its efficient compressor and turbine and its F-class high firing temperature. The gas turbine air extraction provides supplemental air to the ASU.

2.17 Heat Recovery Steam Generators (HRSG)

Two HRSGs shall be provided. Each HRSG will be equipped with a selective catalytic reduction (SCR) system and a spool section for future CO catalyst installation.

2.18 Steam Turbine

One 100 percent steam turbine is included.

2.19 Air Separation System

The Duke Edwardsport IGCC Plant includes a dual cold-box type ASU. The ASU will be designed to produce 100% of the design case HP oxygen rate required by the two gasification trains plus 100% of the LP oxygen required for oxygen enrichment at the two SRU trains.

The feed air requirement for the ASU is satisfied in part by air supplied by extraction from the combustion turbines within the power block. The remaining air required will be supplied by six motor-driven main air compressors (MACs) within the ASU. The MACs are configured as two 50% units with a common extraction air compressor.

The ASU also produces LP nitrogen for purging and HP nitrogen for gasification usage. The remaining nitrogen will be utilized as diluent in the combustion turbine

The Duke Edwardsport IGCC Plant design will include liquid oxygen storage equivalent to four hours of oxygen production at the design case rate. Production of back-up oxygen supply from liquid oxygen storage must be available upon demand. The ASU vendor will determine the need for any gaseous high-pressure oxygen storage based on their assessment of the reliability and availability of their ASU design. For this liquid-pumped design, gaseous HP oxygen storage is not expected to be required.

Liquid nitrogen storage is included for low as well as high pressure purging of gasification equipment mainly during start up and shut down.

3

EDWARDSPORT IGCC PLANT DESIGN EVALUATION

This section will include some overview information related to the Edwardsport IGCC plant design, particularly focused on plant performance, economics, and readiness of the design to accommodate retrofitted CC at various levels. The approach taken by the EPRI team was to create a model of the Edwardsport plant based on GE/Bechtel heat and mass balance data and to compare the overall performance of the plant predicted by the model to the GE/Bechtel values. In some cases, the output from the model is slightly different than that produced by GE/Bechtel, though the values are generally close.

3.1 Overview

The Edwardsport IGCC plant is designed by GE and Bechtel for Duke Energy. A slurry of Indiana #5 coal and water is fed with high purity oxygen (95% by vol.) to two GE designed entrained flow gasifiers to produce synthesis gas that, once cleaned of impurities, may be used as fuel in two GE designed gas turbines (7F Syngas model).

3.2 Plant Performance

Table 3-1 displays performance data derived from the EPRI model of the Duke Edwardsport IGCC power plant. The EPRI model was based on heat and material balance information generated by the GE/Bechtel alliance and represents a normal operating condition (NOC) on the performance coal. The heat rate estimated by EPRI was matched to the Edwardsport heat rate by varying the "miscellaneous" auxiliary loads to match those shown in the GE/Bechtel performance summary. Close agreement was reached, which allows for critical analysis to be conducted on the Edwardsport IGCC plant design vis-à-vis the ease of CO_2 capture and storage (CC) retrofit for various levels of CC.

The EPRI model shows that the anticipated power output of the Duke Edwardsport IGCC plant is 635 MWe generated at a net heat rate of 9,409 kJ/kWhr (8,920 Btu/kWhr) or 38.3% HHV. The design includes two GE 7F Syngas gas turbines each producing approximately 234.2 MWe and a single steam turbine producing ~327 MWe with 11.7 MPa/557°C/544°C (1,700psig/1035°F/1012°F) steam conditions.

As noted in the Introduction, the plant design used in this study reflects an earlier design iteration of the Edwardsport plant. The current design is expected to produce 618 MWe generated at a net heat rate of 9,823 kJ/kWhr (9,313 Btu/kWhr) or 36.6% HHV. The reasons for the change are explained in Chapter 3 of EPRI report 1019668, which is based on public testimony made by Duke Energy to the Indiana Utility Regulatory Commission.

Table 3-1
Base Case Performance Summary: EPRI Model of Edwardsport IGCC Power Plant

Plant Output		
Gas Turbine Power	468,300	kW_e
Steam Turbine Power	326,900	kW_e
Total	**795,200**	kW_e
Auxiliary Load		
Coal Handling	380	kW_e
Coal Milling	3,720	kW_e
Coal Slurry Pumps	1,780	kW_e
Slag Handling and Dewatering	1,780	kW_e
Miscellaneous Gasification Island Auxiliaries	510	kW_e
Syngas Treatment and Conditioning Line	200	kW_e
Selexol Unit Auxiliaries	6,000	kW_e
CO_2 Recycle Compressor	6,300	kW_e
Claus Plant/TGTU Auxiliaries	690	kW_e
CO_2 Compression and Dehydration	0	kW_e
Air Separation Unit Auxiliaries	11,000	kW_e
Air Separation Unit Main Air Compressor	49,330	kW_e
ASU Boost Compressor	11,040	kW_e
Oxygen Pump	500	kW_e
Nitrogen Compressor	40,460	kW_e
Gas Turbine Auxiliaries	1,000	kW_e
Steam Turbine Auxiliaries	1,500	kW_e
Boiler Feedwater Pump	5,850	kW_e
Condensate Pump	300	kW_e
Circulating Water Pump	9,000	kW_e
Cooling Tower Fans	4,000	kW_e
Miscellaneous Balance-of-Plant	1,000	kW_e
Deep Well Injection	1,600	kW_e
Transformer Losses	2,000	kW_e
Total Auxiliaries	159,940	kW_e
Net Plant Power	635,260	kW_e
Net Plant Efficiency (HHV)	38.3%	
Net Plant Heat Rate (HHV)	8,920	Btu/kWhr
Coal Feed Flowrate	508,235	lb/hr
Thermal Input, HHV[1]	1,660,781	kW_{th}
Condenser Duty	1,690.0	MMBtu/hr

Note: ASU auxiliaries contain MWe's not accounted for in the EPRI model.

3.3 Evaluation of Edwardsport Plant Design Readiness for CC Retrofit

The following sections describe key process areas of the Duke Edwardsport IGCC plant and the retrofit-ability of the design.

3.3.1 Air Separation Unit

3.3.1.1 Introduction

The ASU specified for the Edwardsport IGCC facility appears to be able to support a possible future conversion of the facility to any of the three retrofit cases studied - CC at nominally 20%, 50% and 80-90%. This discussion will speak to additional equipment needs and other considerations associated with the potential retrofits.

Required ASU changes for the retrofit scenarios are primarily:

1. The need for supplemental air compression, as the availability of gas turbine extraction air decreases with increasing CO_2 capture
2. The need for diluent nitrogen booster compression for the cases where a portion of the diluent nitrogen must be delivered at a higher pressure upstream of the gas turbine section control valve in order to enable compliance with the maximum heating value limit, or the maximum H_2 limit of 60%.

If some loss of IGCC power production with lower quality coals or at high ambient temperatures can be tolerated, the ASU cryogenic section capacity should be sufficient for the three retrofit cases. Such analyses can be performed during the FEED, as desired. It is recommended that at least three such analyses be performed on retrofit cases of interest: one to confirm operation on the base case, one on a lower quality coal of choice at 15°C (59°F) ambient temperature, and one on Indiana #5 coal to determine how high the ambient temperature can be before power production begins to decline. Ramping and turndown performance should be examined on selected operating cases, to determine any loss of operating flexibility for retrofit cases of interest.

Because the ASU industry is very competitive and, in general, very capable of designing ASU’s to exacting tolerances in order to achieve specified requirements, it is important to use a thorough specification of required flows, purities and pressures, as well as other performance characteristics; e.g., turndown, ramping, availability, etc. such as has been developed for the Edwardsport project for the various cases of interest, and not to expect or depend upon unspecified design margins.

3.3.1.2 Base Case Process Description (no CC)

The function of the air separation unit (ASU) is to provide oxygen for the gasifier and sulfur recovery unit (SRU), and nitrogen for syngas dilution and other uses within the facility. The ASU design is based on an elevated pressure, pumped liquid oxygen (LOX) process cycle, in which the cryogenic distillation columns are operated at elevated pressures as part of the optimized configuration for delivering a large quantity of compressed nitrogen for fuel gas dilution. The pumped LOX feature of the cycle involves taking liquid oxygen from the cryogenic distillation section of the ASU and pumping it through coldbox heat exchangers in which the refrigeration of the liquid oxygen is recovered by cooling down a portion of the ASU main air compressor discharge stream which has been further boosted in pressure for efficient heat exchange.

The Edwardsport base case design also provides for a portion (nominally up to 40%) of the feed air to the ASU being supplied by air extracted from the gas turbine air compressor. The available extraction air is determined by the desire to maintain the maximum output from the gas turbines, and thus decreases for increasing percentages of CO_2 capture and for increasing ambient operating temperatures.

3.3.1.3 Edwardsport ASU Design Basis and Operating Cases (no CC)

A variety of design and operating cases have been specified by the GE/Bechtel team and their subcontractors which collectively define ASU's functional requirements. The major design cases provide for margins in ASU cryogenic section capacity and main air compressor capacity that can be considered in determining the equipment additions necessary to support the retrofit options, and will be further discussed in the sections for those individual cases. These cases are listed below.

- Maximum O_2 case, -18°C (0°F). This case will size the cryogenic section.
- Maximum air case, 38°C (100°F). This case will govern the MAC frame size.
- Cryogenic design flowrate case (flow rates will not be achieved at all ambients)
- Maximum LOX case. This case will govern an aspect of process cycle design.
- Case 2: Indiana #5, 18°C (59°F) (Same as commercial guarantee case, except that Case 2 total N_2 is about 4.4% higher than in the Guarantee Case). This case will be used to determine the design of the MAC for maximum efficiency.
- Indiana #5, 18°C (59°F), 50% turndown (one GT at 100%) case
- Indiana #5, 18°C (59°F), 50% turndown (two GT's @ 50% each) case
- Case 1: Indiana #5, -7°C (20°F)
- Case 5: Indiana #5, 32°C (90°F)
- Duke Additional Case 1, 18°C (59°F)
- Duke Additional Case 2, 18°C (59°F) (Total O_2 about 3% less than for Additional Case 1)
- Ramping cases (50 to 100% of design rates in 35 minutes maximum with plus/minus limit of 2% on O_2 purity)
- Extraction air booster case---one GT at baseload, one at 40%. This case is expected to size the extraction air booster compressor.
- Start-up transient cases (depend on availability of extraction air)
- High chloride coal case
- High ash coal case

- High sulfur coal case
- High moisture coal case
- High solids coal case
- Reference Plant poor coal case
- Reference Plant 50% turndown case, -18°C (0°F) (2 GT's @ 50% each)
- Reference Plant 50% turndown case, 38°C (100°F) (1 GT @ 100%)

The many operating cases can be used by the supplier to test an ASU model comprised of equipment pieces individually sized by the various major sizing cases. In this way, the operation of the plant for the various operating requirements can be verified, and any constraints identified and dealt with as appropriate.

3.3.2 Gasification Area

The gasification area encompasses coal slurry preparation through post-gasification syngas scrubbing. Coal slurry and oxidant are fed to the gasifier and outputs are saturated synthesis gas, slag, and wastewater. The coal slurry is comprised of Indiana #5 coal and water and is fed with high purity oxygen (95% by vol.) to two GE-designed entrained flow gasifiers to produce synthesis gas that, once cleaned of impurities, may be used as fuel in two GE designed gas turbines (7F Syngas model).

The gasifiers produce a syngas from particulate scrubbers with ~240 Btu/scf on a dry basis. The syngas contains 20.5% CO_2 on a dry basis, which represents ~35% of the carbon in the feedstock.

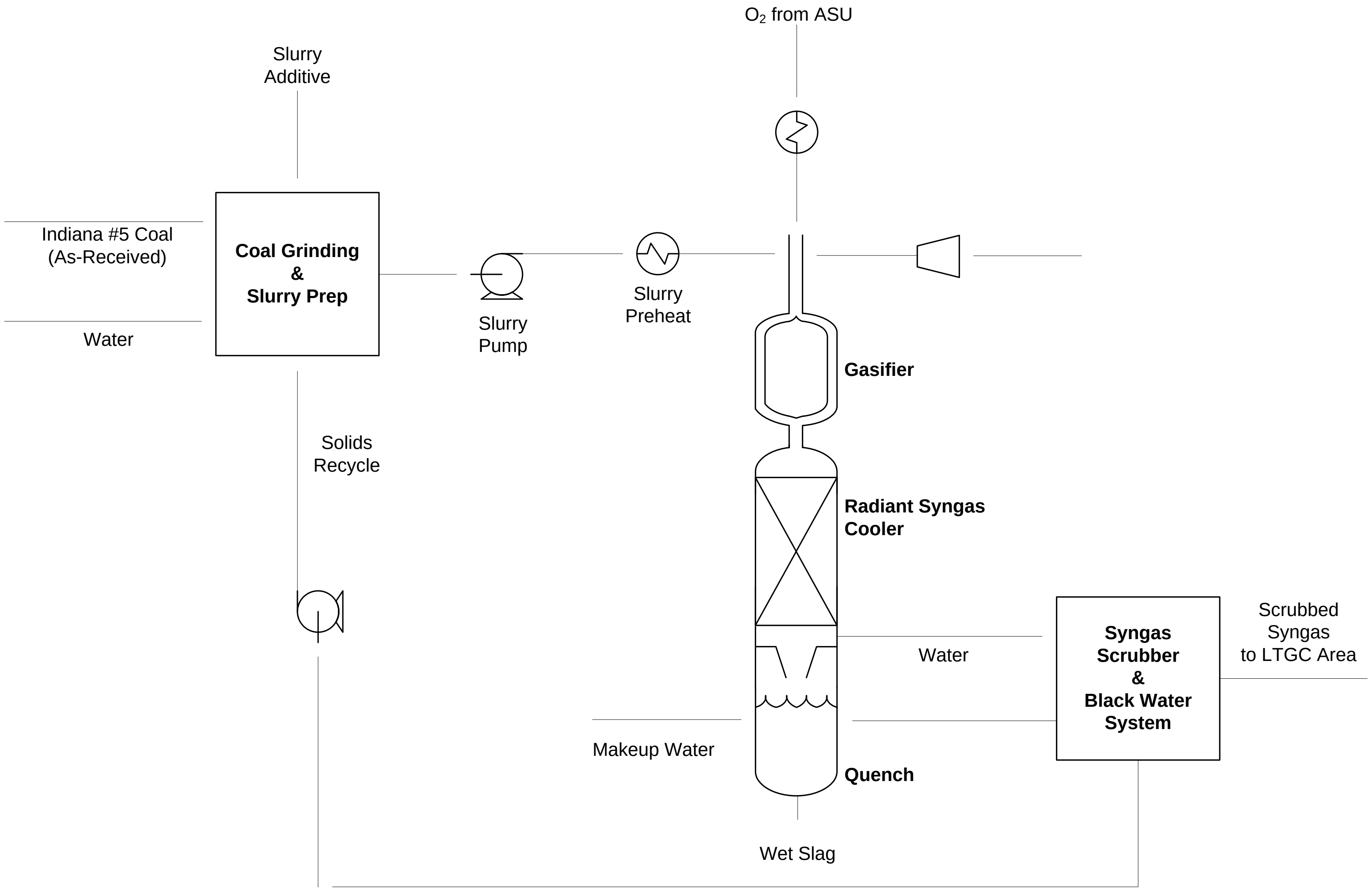


Figure 3-1
Gasification Area Process Flow Diagram and Performance Summary

3.3.3 Low Temperature (LT) Gas Cooling

The LT gas cooling section of the plant can be retrofitted with the equipment required for carbon capture. For CC case 1 (20% CC) no changes are anticipated. For CC cases 2 (50%) and 3 (80-90%) a CO shift section will have to be inserted. Some of the existing equipment may be reused in the new situation.

For the base case the low temperature gas cooling (from scrubber outlet to trim cooler knockout drum) includes a COS hydrolysis to convert the bulk of the COS to H_2S, which improves the economics of the downstream Selexol AGR unit. The heat in the scrubber exit gas is used to generate low pressure steam and preheat circulation water for the syngas saturator and steam condensate from the steam cycle. A superheater operating on MP steam is provided to ensure that adequate superheat above the dew point is available at the catalyst bed inlet.

EPRI has modeled this cooling train to generate a start point for modeling the retrofit cases. The results show good consistency with the GE-Bechtel material balance supplied.

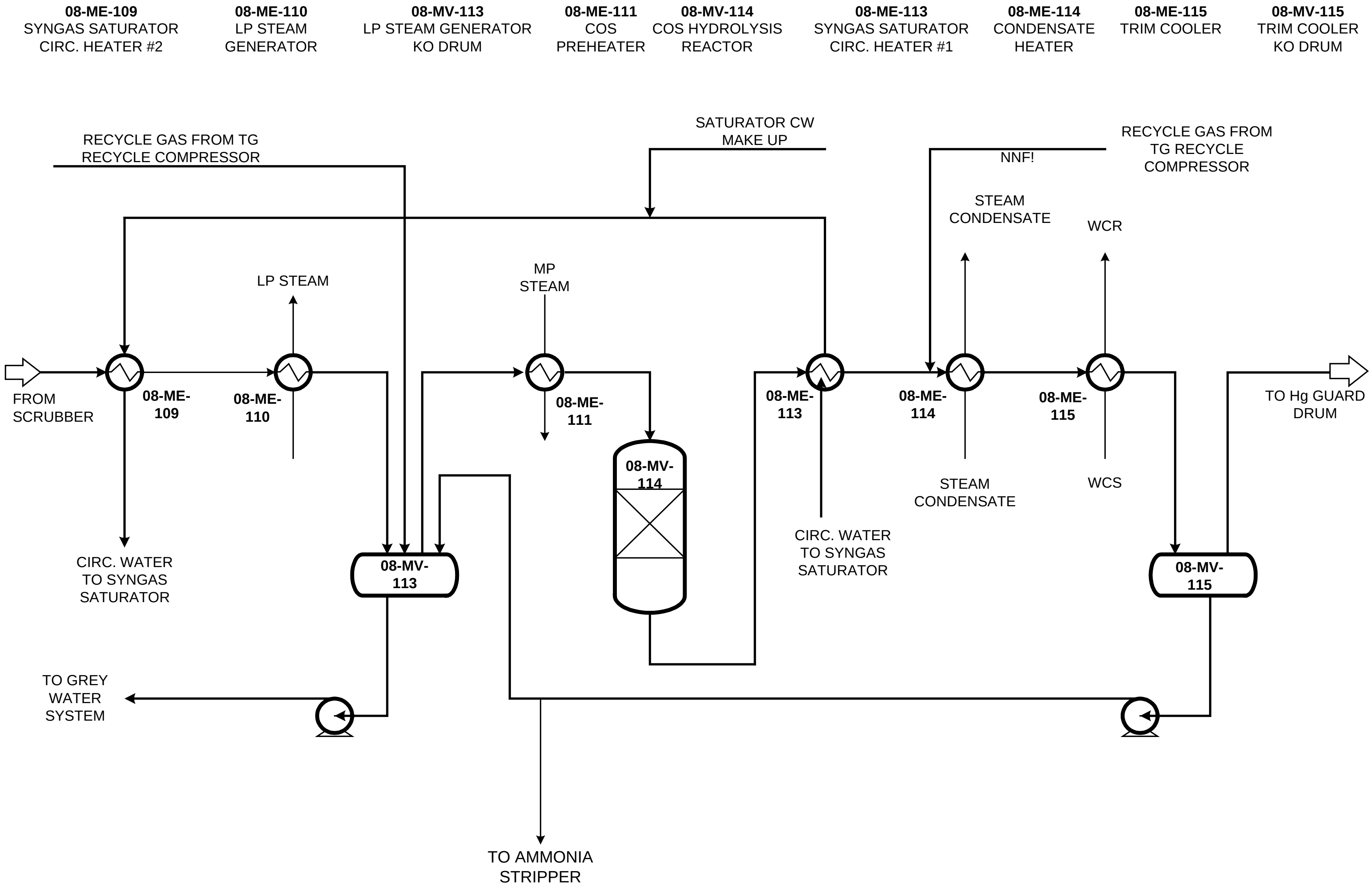


Figure 3-2
Low Temperature Gas Cooling System Process Flow Diagram

It should be noted that in the documentation supplied, the gas outlet temperature of the trim cooler is shown as 33°C (91°F), while on the utility summary the cooling water supply temperature is shown as 31°C (88°F). Since this combination would lead to unreasonably large exchanger surfaces, it is assumed that these two figures originate from different operating cases, particularly since in other documentation relating to the power block a cooling water supply temperature of 22°C (71°F) is mentioned. No attempt has been made to reconcile these figures or to deduce an exchanger surface area from them.

3.3.4 Acid Gas Removal (Selexol Process)

3.3.4.1 Basic Selexol Configuration (2-Tower)

The basic Selexol process configuration for H_2S/COS capture without CO_2 capture is depicted in Figure 3-3. This configuration is quite similar to the UOP Selexol Process for the Duke IGCC project, but differs as follows:

- Duke recycles the flash gas from the rich solvent flash upstream, as opposed to the standard process of recycle back to the Selexol absorber. Both serve the same purpose for the Selexol unit.
- The Duke Selexol process for de-sulfurization employs refrigerated solvent for absorption, while the standard process utilizes above-ambient operation via cooling water. Both routes are acceptable, and it is an economic analysis decision of reduced power for solvent flow and steam for reboiler duty versus added power for refrigeration.
- The Duke Selexol process uses pre-loading of the lean solvent via mixing and then separation with product gas, while the standard process does not. Both routes are acceptable, and it is an economic analysis decision of incremental equipment versus reduced solvent flowrate and thus reboiler duty, which reduces size of existing equipment.

This flow scheme is employed when the CO_2:H_2S ratio in the raw syngas feed is generally <10/1, since this flow scheme produces an acceptable H_2S concentration in the acid gas (generally >35%V H_2S) for good Claus SRU design/operation/economics. The CO_2:H_2S ratio in the Duke design for the performance coal (Case 1) is ~ 9/1, and for the base case is ~ 10/1.

Another observation on feed gas composition is the COS concentration, which is 55 ppmv in the performance coal case and 70 ppmv in the design conditions (maximum flow) case. These are quite high values, even for end-of-life conditions for the COS hydrolysis catalyst. Since COS is a more difficult species to remove than H_2S (while trying to selectively reject CO_2), this places an additional processing burden on the Selexol process design to achieve product synthesis gas specification while simultaneously generating a quality acid gas to the SRU in the GEE design.

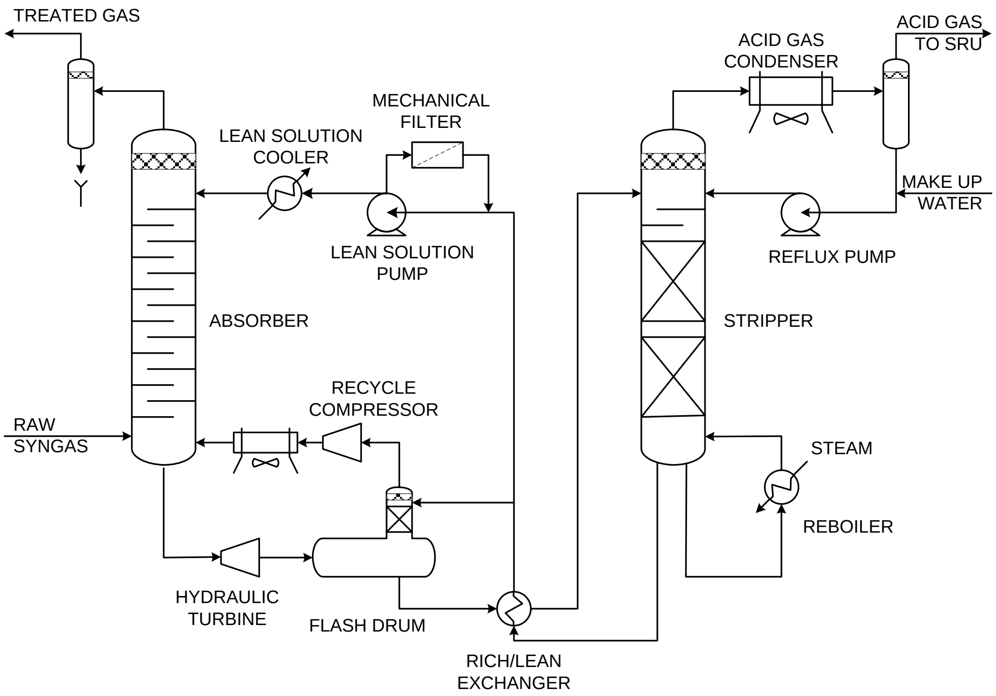


Figure 3-3
Selexol Flow Diagram (2-Tower Configuration) for H_2S/COS Capture

3.3.4.2 Selexol Configuration with H_2S Enrichment in the Acid Gas

A modification is made to the aforementioned configuration when greater enrichment of the acid gas in H_2S is required. This alternate configuration is employed when the CO_2:H_2S ratio in the raw syngas feed is generally > 15/1, since this configuration produces a quality H_2S concentration in the acid gas (generally >35%V H_2S) for good Claus SRU design / operation / economics. This CO_2:H_2S ratio will be exceeded when either partial or complete CO-Shift is added to the upstream processing of the synthesis gas.

In these situations, selective absorption of H_2S in the Absorber is insufficient, so selective regeneration is also required to produce a sufficiently enriched acid gas. The main change is the transformation of the flash drum into an H_2S-concentrator tower, as depicted in Figure 3-4. The rich solvent discharging the absorber bottom is heated in the lean/rich heat exchanger, let down in pressure, and then enters the top of the H_2S-concentrator tower. A flash occurs, but then the solvent passes counter-currently down the column where it meets an upward flowing stripping gas stream of N_2. The N_2 not only removes all the remaining H_2 and CO, but its main function is to selective strip CO_2 over H_2S from the solvent. The overhead stream from the H_2S-concentrator is compressed and recycled to the absorber, while the H_2S-enriched rich solvent from the bottoms optionally undergoes additional degassing steps before it is sent to the Stripper for solvent regeneration. All other aspects of the Selexol configuration conform to the process description for Figure 3-3.

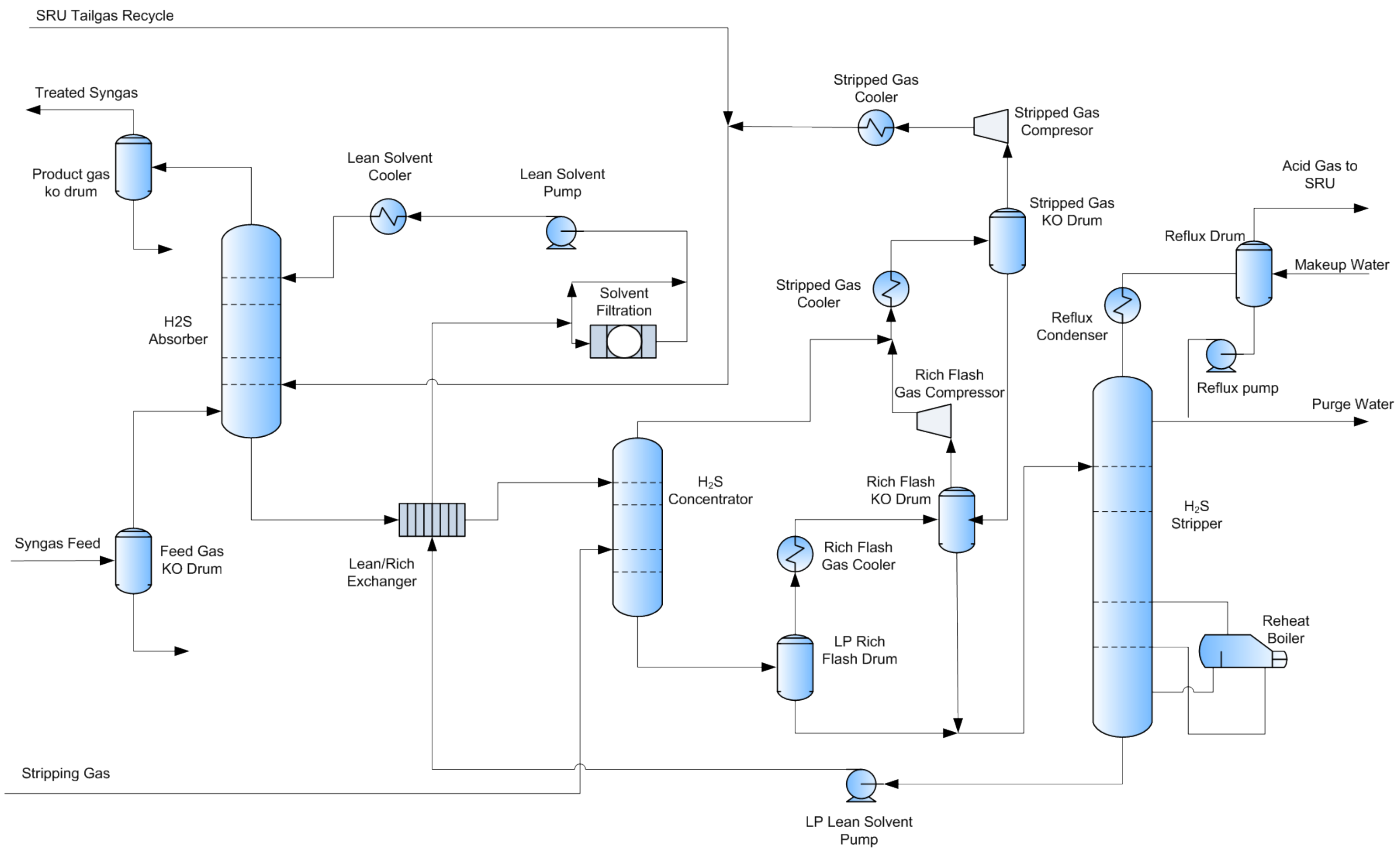


Figure 3-4
H_2S removal (base case): AGR process flow diagram

3.3.4.3 Selexol Process for Capturing H_2S/COS and CO_2 Separately

When CO_2 capture is required, this generally means the raw syngas has undergone some level of CO-shift, and thus high levels of CO_2 are present in the syngas. Due to the resulting high CO_2:H_2S ratio in the syngas feed, and the need for CO_2 capture, a CO_2 removal section has to be added to the base configuration in Figure 3-5. While it could be sequential in nature and completely separate from the H_2S removal section of the Selexol process, there are inherent process advantages from integrating the two sections.

- Figure 3-5 presents a general integrated Selexol process flow diagram for separate H_2S and CO_2 capture. The lower section of the flowsheet is the H_2S removal section, and is identical to the 3-tower configuration, while the upper section is the added CO_2 removal section. The CO_2 is captured and removed in a CO_2 absorber and multiple rich solvent flash drums, in order to recover the CO_2 at various elevated pressures. Virtually all of the CO_2 is removed by flash regeneration of the solvent, thereby minimizing overall thermal heat input to the system.
- The H_2S removal section and the CO_2 removal section are integrated in the following manner:
  - Part of the rich CO_2-laden solvent discharging the CO_2 absorber bottom is used as lean solvent for the H_2S absorber. This preloading of the solvent (with CO_2) assists in H_2S selectivity and solvent flowrate minimization in the H_2S absorber
  - The regenerated lean solvent from the stripper is sent to the CO_2 absorber top for incremental CO_2 recovery

A succinct summary of the process description goes as follows:

- H_2S & CO_2 absorbers operate at ~ $4^{\circ}C$ (~$40^{\circ}F$)
- Same solvent for H_2S and CO_2 sections
- H_2S absorber removes all the H_2S and most of the COS
- CO_2 absorber is divided into two sections
  - Bulk CO_2 removal section using flash-regeneration
  - Trim CO_2 removal section in which regenerated solvent is used
- From the H_2S absorber, the rich solvent is heated, flashed to an intermediate pressure, and back-stripped with a slipstream of N_2 (or alternatively a slipstream of product synthesis gas) in the H_2S concentrator (or optionally a slipstream of N_2)
- Solvent is then sent to the regenerator for complete sulfur removal
- Acid gas sent to the Claus SRU.
- Thermally-regenerated solvent is sent to the top of the CO_2 absorber for trim removal
- Flash-regenerated solvent is sent to the bulk CO_2 removal section of the CO_2 absorber

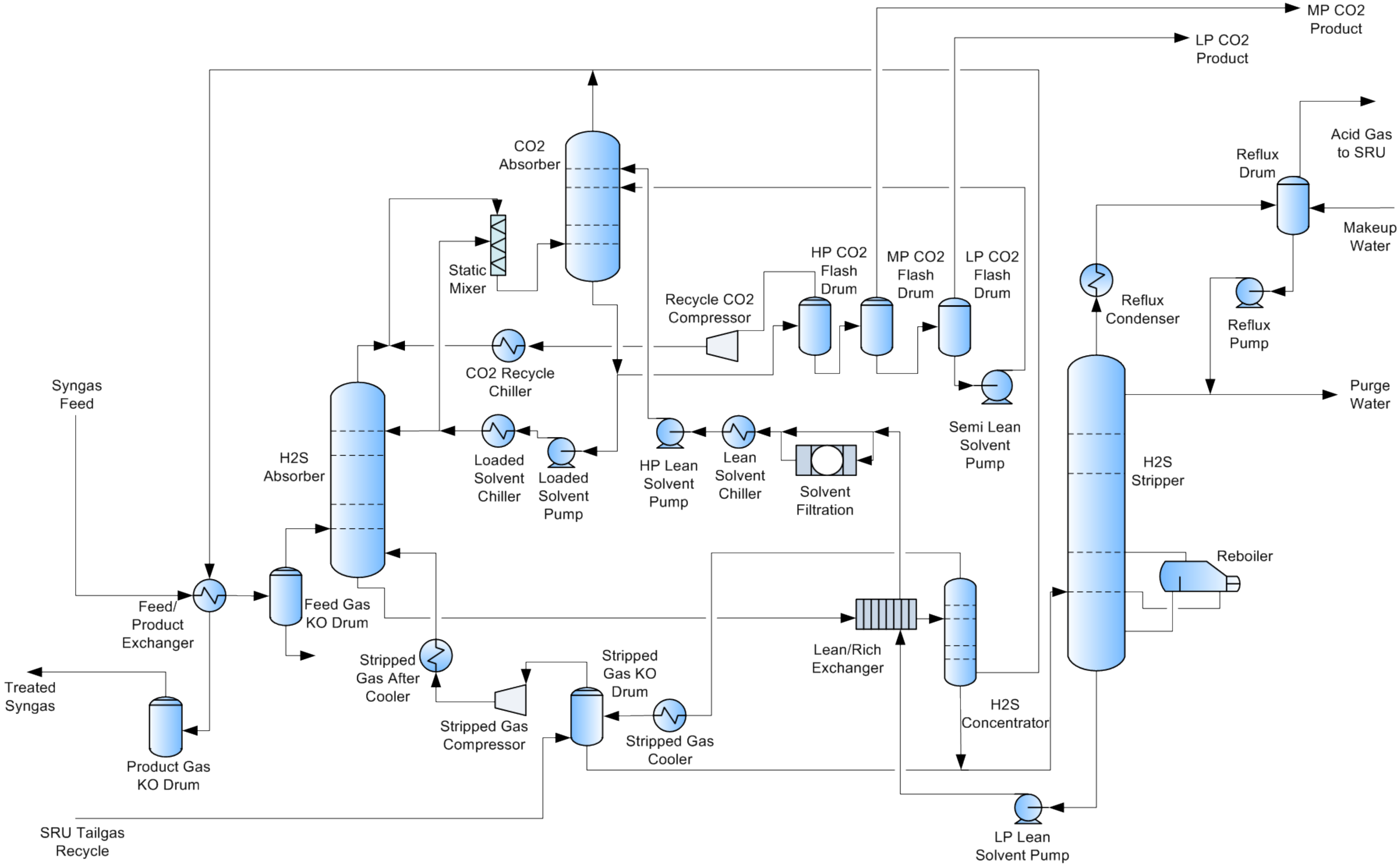


Figure 3-5
H_2S and CO_2 removal (retrofit case): AGR process flow diagram

- CO_2-laden solvent is flash regenerated in several stages:
  - The first HP flash stage is used to recover and recycle co-absorber H_2 back to the Selexol unit to increase H_2 yield into the product synthesis gas
  - The second MP flash stage is used to deliver some of the CO_2 at elevated pressure to an intermediate stage of the CO_2 compressor
  - The final LP flash stage is used for final flash-regeneration of the solvent, and the resulting CO_2 is fed to the suction of the CO_2 compressor
  - The individual flash tanks can be “stacked” in one segmented vertical vessel

3.3.5 Claus Unit

For the base case a generic oxygen-blown, two-stage Claus unit with tail gas hydrogenation and recycle has been specified. The plant is specified to process a range of acid gas specifications, depending on the coal fed to the gasification unit.

In terms of gas quality, it should be noted some cases may be marginal for effective Claus plant operation even with pure oxygen firing. It must be expected that under these conditions the design single-pass sulfur recovery rate may not be achieved. However since the tail gas is being recycled, this may be acceptable.

In terms of turndown, some cases may have gas flow rate to the burners near the minimum capability of a typical Claus plant burner (about 30%). If any case predicts a flowrate below 30% of the maximum design value, a special burner may be required to handle this turndown requirement or alternatively syngas may have to be added to the acid gas to increase the flow rate at the expense of the sulfur recovery rate. It will also be necessary to ensure that the relative oversizing of exchangers in the downstream processing does not lead to sub-dewpoint conditions and consequent corrosion.

3.3.6 Combined Cycle

The combined cycle being designed for the Edwardsport IGCC appears quite suitable for conversion to CO_2 capture in all of the 3 cases studied. The major issue is that after CO_2 removal, the flow and characteristics of the syngas fuel sent to the combustion turbine are significantly different. The resultant syngas will have higher H_2 levels and will require dilution (similar to syngas) with N_2 and/or steam to meet NOx requirements and maintain the combustion turbine output enhancement normally found when using low-Btu fuel gas. Both N_2 and moisture are available in sufficient quantities. Some minor modifications in the gas turbines, auxiliary equipment and steam cycle will be needed for each case and they are covered on a technology basis in the GT/CC technology section that follows with details in the individual case section.

The 7F Syngas incorporates many features for syngas and high hydrogen fuels. The conversion design issues for the various cases are mainly focused on the change in flows for various pieces of equipment and the increased hydrogen content in the syngas. The GT flows after conversion to capture tend to be smaller so it is this high H_2 content and combustion that has required a significant development effort. An estimated $150 million has been spent by GEE to develop the gas turbine and combustor system as a commercial product.

Model F syngas combustion systems were first developed in the early 90's and utilized 40.6 cm (16 in) diameter, can-annular designs to provide longer residence times allowing better CO burnout for low BTU fuels as opposed to NG designs with 36 cm (14 in) cans. These combustors were used successfully in IGCC plants such as TECO and Wabash.

In the late 90's combustion tests were completed with FA combustors using high hydrogen fuels and GEE offered the conversion capability as a commercial product. Combustion testing with these designs has now been done for the 7F Syngas model with regular syngas and with varying high hydrogen fuels. Initial offerings also include reductions in firing temperature from natural gas applications to further insure suitable RAM results.

In addition to the gas turbine modifications, this section also addresses the flow changes in the steam cycle and potential variations in operating modes.

3.3.6.1 GT/CC Technology and Description

The team has studied several ways to accommodate CO_2 capture in the Duke Edwardsport combined cycle based on various previous studies of GEE designs. The basic issues have to do with the GT/CC flow and composition variations when the CO is converted to CO_2 and removed from the syngas.

Generalized conclusions for the combined cycle with and without capture are as follows:

- Gas turbine power output is maintained except for high ambient temperature cases even with lower firing temperatures needed for higher H_2. Steam turbine power is expected to vary slightly as increasing levels of CO_2 capture are retrofitted onto the plant.
- GT mass fuel flow after CC is reduced due to the high H_2 and low CO_2 content of the resulting fuel. The GE specification on syngas feed to the gas turbine sets a maximum H_2 content. Due to elevated levels of H_2 in some cases, pre-dilution of syngas with N_2 is required.
- Syngas humidification is not utilized in CC Cases 2 and 3 due to elevated hydrogen levels in the syngas and sufficient availability of nitrogen for gas turbine NOx mitigation.
- Air extraction availability is reduced by as much as 57% due to lower diluted fuel mass flow for the high capture case. Other similar studies have shown a more drastic reduction suggesting that air extraction may not be economic except for a conversion application where the equipment is already in place.
- GT exhaust flows remain similar at lower temperature for the high capture cases indicating an overall balance with considerable variation in components but not a large issue for the steam cycle

- Steam turbine power outputs are maintained even with lower GT exhaust temperature due to use of the water-gas shift reactor intercooler heat for LP steam generation that feeds the steam turbine. This increase of LP flow is partially balanced by other operating conditions that dictated higher flows resulting in only a 2% increase. Over speed valve protection will be needed for this new flow.
- SOx emissions tend to be reduced as the sulfur ends up in the CO_2 stream. GT NOx emissions are controlled to similar levels with nitrogen and moisture. NOx emissions tend to increase on a unit of mass per MWhr basis as auxiliary losses reduce net IGCC power output.
- Modifications for the skimming case appear minor and even for the full capture case look feasible

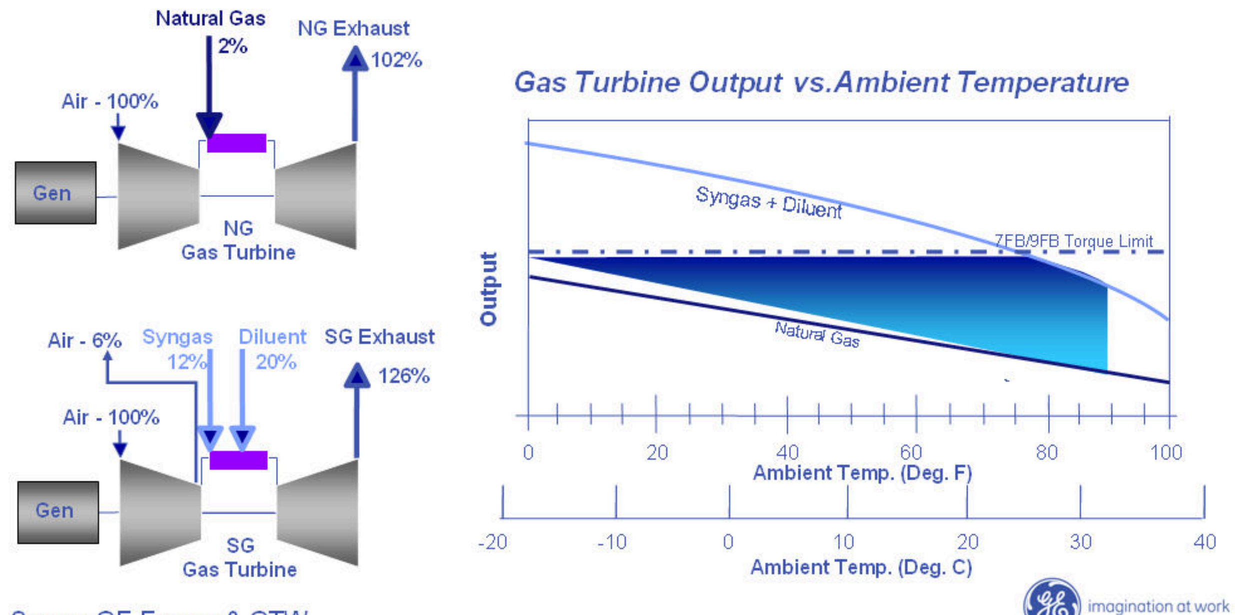
3.3.6.2 Gas Turbine Technology Discussion

An understanding of combustion turbine flows and output across the ambient range is very important for IGCC design and integration to obtain economic viability and operational flexibility. To obtain necessary syngas NOx reductions during combustion, diluents such as steam or nitrogen are used. A secondary effect of syngas dilution is increased flow through the turbine, which increases the power output of the machine. Diluents used for NOx control and power augmentation can increase GT ratings 15-25% above natural gas ratings.

Since combustion turbines are designed for high ratings at low ambient temperatures it is possible that the higher production capability will apply across the ambient range. Ratings are constrained by rotor torque limits and/or other turbine factors creating the so-called flat rating. The higher flows may require some combustion turbine modifications and possibly air extraction to balance flows for some models. The use of extraction of air to the ASU can add even more power augmentation to the IGCC by reducing ASU auxiliary losses. Figure 3-6 shows a typical curve for rating vs. ambient temperature for natural gas and syngas along with a rotor torque restriction.

Some notes for Figure 3-6:

- The top left diagram of a natural gas turbine shows that the mass flow of NG is about 2% of the flow through the turbine section
- The bottom left diagram shows that syngas (SG) with its lower heating value accounts for 12 % flow and the diluent about 20% of the flow through the turbine. In this example extraction air of 6% is taken leaving a net 24% more flow through the turbine than with NG fuel. The extraction helps to accommodate other production turbine limits and can mitigate some ASU auxiliary power requirements
- The principle of a gas turbine is that the turbine power is roughly double the net power and in this case the combustion turbine air compressor is still the same subtraction, so the turbine generator net would like to increase 48% if fired to the same temperature.
- On the right is the ambient vs. output curve. It shows the machine torque limit forcing the flat rating
- Providing the gasifier is sized for the combustion turbine torque limit the blue area is power that is available when on coal/syngas



Source GE Energy & GTW

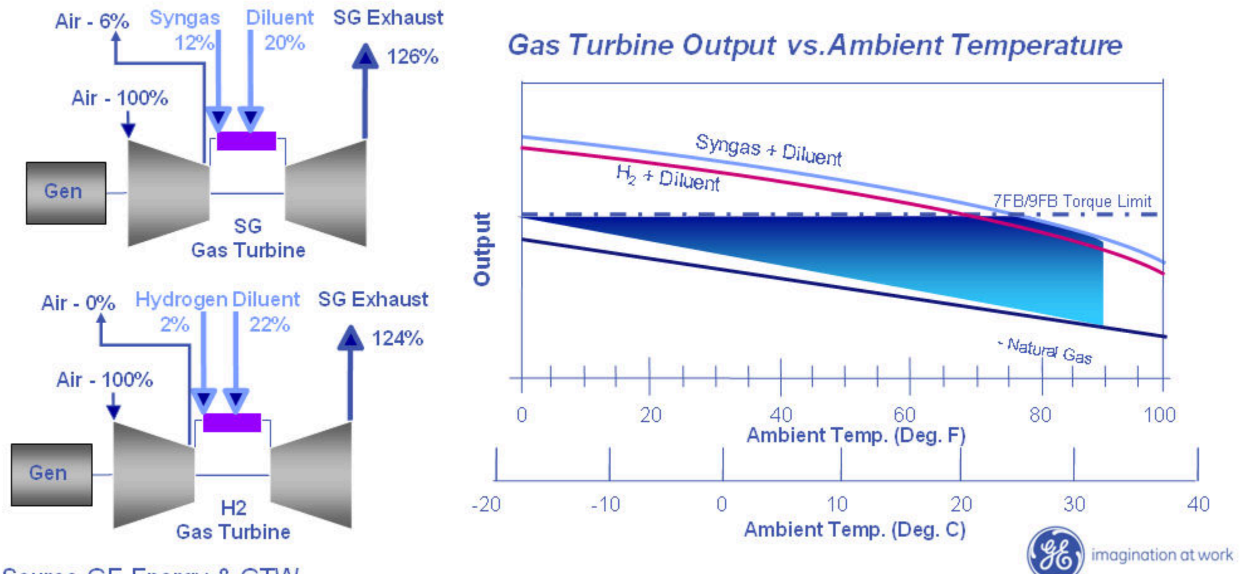
syngas firing case includes combustion turbine compressor air extraction to ASU

Note: *Chart is representative for typical conditions. The syngas rating ambient temperature crossover point with maximum torque varies with the specific syngas constituents.*

Figure 3-6
IGCC GT Output Enhancement w/o Capture vs. Ambient Temperature Compared to Natural Gas Operations

For CO_2 capture applications, the IGCC fuel for the combustion turbine will be a syngas with high H_2 content. Figure 3-7 is a similar rating vs. ambient temperature diagram for non-capture syngas versus high-hydrogen capture syngas.

- Some notes for Figure 3-7:
  - The bottom left diagram shows the CO_2 capture case where H_2 is only 2% of the mass flow and when added to the diluent is only 24% - not enough to allow much air extraction. The red line on the ambient curve shows the issue – dropping the break point to 70-80°F rather than 80-88ºF. Inlet evaporative coolers or chillers can enhance the ambient break point.



Source GE Energy & GTW

- *Syngas firing case includes combustion turbine compressor air extraction to ASU*
- *H_2 firing case may not include combustion turbine compressor air extraction to ASU*

Note: *Chart is representative for typical conditions. The syngas rating ambient temperature crossover point with maximum torque varies with the specific syngas constituents and turbine.*

Figure 3-7
IGCC GT Output with and w/o Capture vs. Ambient Temperature Compared to Natural Gas Operations

Gas Turbine Integration with ASU

The term “air-side integration” is intended to reflect extraction of air from the combustion turbine compressor section to partially supply the pressurized air used by the ASU producing oxygen and nitrogen for an oxygen-blown gasification system. This system usually implies an elevated pressure ASU so that N_2 is available for NOx control. Figure 3-8 illustrates a simplified process flow diagram of an IGCC where air is extracted from the GT for the ASU.

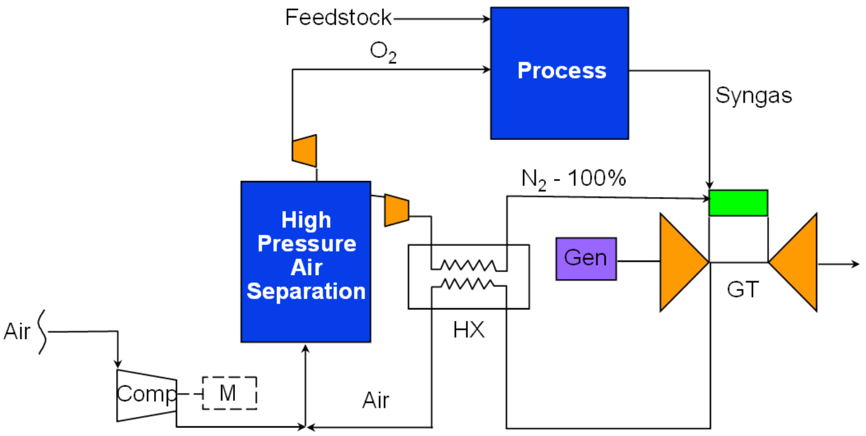


Figure 3-8
Integration of ASU and Gas Turbine

Some notes on Figure 3-8:

- For cases without capture it is possible to extract air from the gas turbine at certain ambient conditions to reduce ASU auxiliary loads. The air is extracted when the turbine capability is above its torque limit, allowing the gas turbine to produce maximum electrical power based on the torque limit, and increased partial air for the ASU which provides a net gain for the IGCC
- Partial extraction implies about 35% of the air for the ASU being extracted from the combustion turbine (maybe 6% of the gas turbine air flow). In this case the ASU can startup and operate independently. This is similar to the Duke Edwardsport base case.
- For CO_2 capture cases the turbine flow is reduced to a point where extraction of air without reducing GT output may be uneconomical. However, since Edwardsport will already have the investment in place a small amount of extraction may be justifiable.

3.3.6.3 Syngas and Hydrogen Combustion Technology

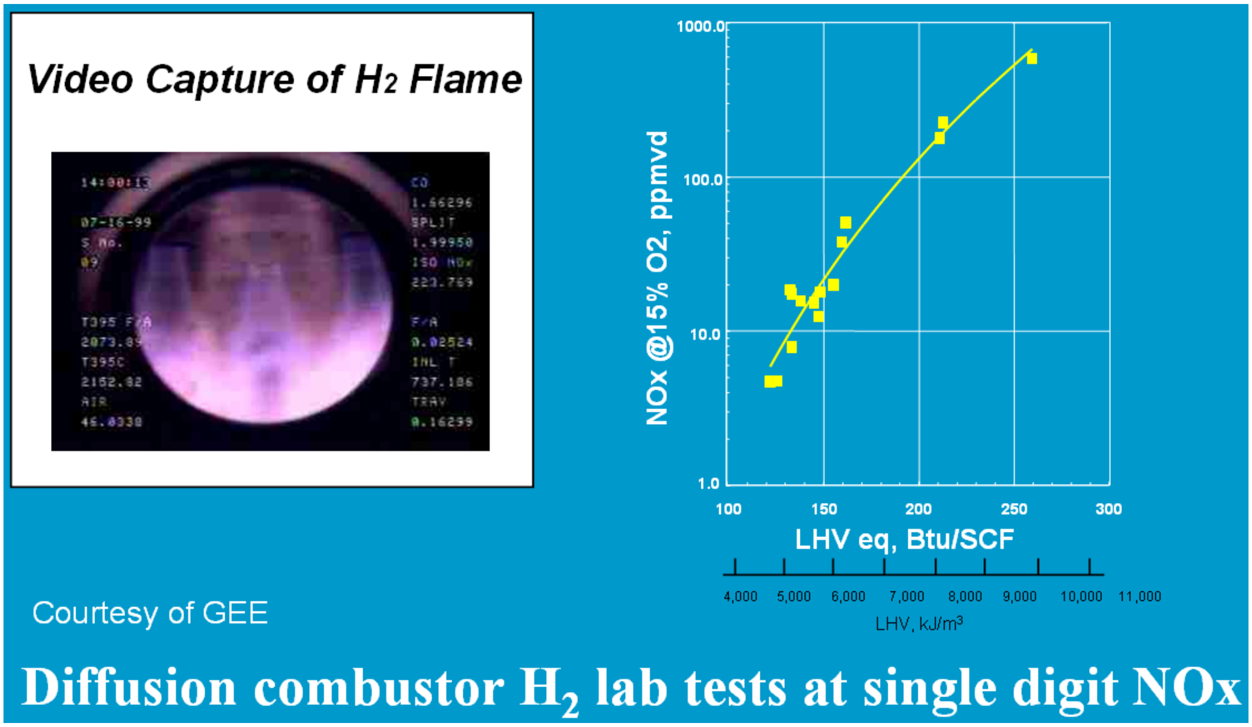
GE, MHI and Siemens currently use can annular diffusion combustors for most syngas and H_2 applications. There is considerable field experience for both syngas and high H_2 fuels in older gas turbines but it has been necessary to do full-scale, full-pressure tests on new production model combustors to ensure the transition to modern IGCC turbines. That effort has been successful in developing the commercial product planned for Edwardsport. It will be important to do a final test on the combustors to establish the production configuration based on detailed engineering completion.

CO_2 capture rates can vary up to 90% thereby increasing H_2 concentration in the GT fuel in a similar manner as evidenced by the three CC cases in this study. That means that the technology needs to be developed across a wide range of H_2 concentrations. It is very important to distinguish between H_2 concentration when the other combustible constituents are also present such as CO and H_2 concentration when the other constituents are only diluents.

In the late 1990s, GE ran production model syngas combustors modified for H_2 fuel at full pressure and temperature with various amounts of diluents from 45% H_2 to 85% H_2. The video photo in Figure 3-9 is one of the actual H_2 test runs in a GEE-6FA combustor can which is the syngas combustor model for use in F model turbines

Lab measurements show on the right as low as 4-5 ppmvd NOx with 45% H_2 and 55% N_2. This also happens to be near the point of maximum power augmentation. This work launched the commercial offerings of modern combustion turbines for CO_2 capture applications.

Extensive recent testing on the GEE-7F Syngas combustors shows similar results.

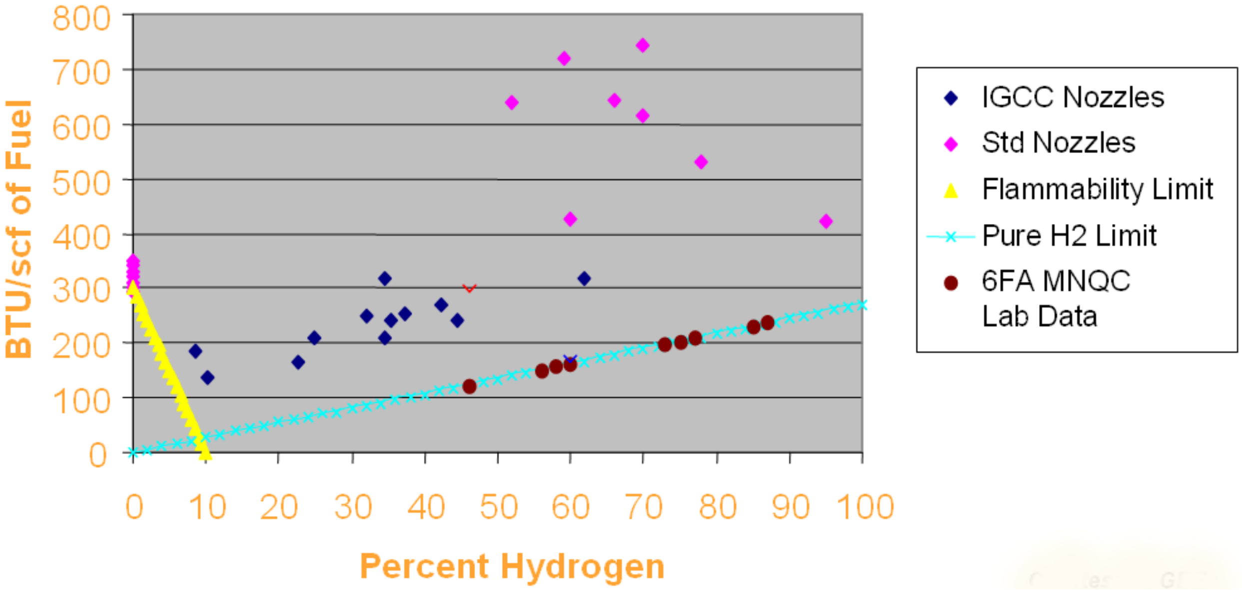

Video Capture of H_2 Flame

Courtesy of GEE

Diffusion combustor H_2 lab tests at single digit NOx

Figure 3-9
NOx Production for Hydrogen Fuel

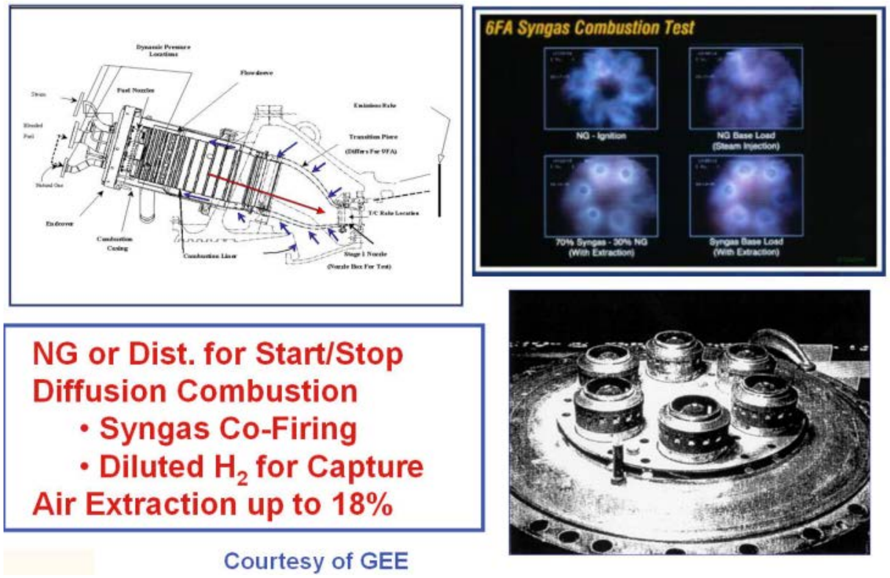
Figure 3-10 shows GEE combustion turbine experience across a wide range of hydrogen concentrations. The plot of Btu/scf vs. H_2 % shows actual field experience along with the above specific tests of diluted hydrogen-only fuel– red dots.



Courtesy of GEE

Figure 3-10
GEE Experience with Diffusion Combustors Firing High Hydrogen-Content Syngas

Typical can-annular combustors for syngas and hydrogen are shown in Figure 3-11.



NG or Dist. for Start/Stop
Diffusion Combustion
- Syngas Co-Firing
- Diluted H_2 for Capture

Air Extraction up to 18%

Courtesy of GEE

Figure 3-11
Syngas Combustion Development

3.3.6.4 Range of Combustor Fuel Flexibility

Fuel nozzle sizing in the multi nozzle combustors is critical to operational performance, parts life and other RAM criteria. It must take into account summer and winter as well as full speed full load and full speed no load operation. Typically, the Wobbe Index is used to establish fuel flexibility for a fixed fuel nozzle sizing. In the past this led to the need to change fuel nozzles for excessive changes in fuel heating value. Changes in CO_2 and H_2 content for CC cases above 60% capture would normally require such nozzle change outs. Below 60% capture rates the nozzles can be sized for mid-range and perform equally well for normal syngas and the higher H_2 fuel.

In other words, fuel switching may be accommodated even on-line. However, if a controllable syngas heater (performance heater) is installed it is possible to vary the fuel temperature to compensate for excessive Wobbe index changes and extend the flexibility, perhaps all the way to 90 % capture. This would require approximately 300°F syngas temperature variability. In any case it will be crucial to study this subject thoroughly for future IGCC CC designs.

3.3.6.5 Syngas Combustion Turbine RAM and Controls

Two separate factors involved in syngas and fuel can affect combustion turbine hot gas path (HGP) parts life. Excess flow due to LHV syngas can cause overheating of the gas turbine hot gas parts. High mass flow increases heat transfer and metal temperatures. Higher moisture content in the turbine flow due to H_2 in the syngas can cause overheating. Water vapor from H_2 combustion or steam injection has higher heat transfer properties. If not controlled, parts life could be decreased by up to 50%.

Fortunately both of these effects can be controlled and/or eliminated. Air extraction can partially or fully mitigate the excess flow effect. Controls should adjust the CT firing temperature to maintain metal temperatures equivalent to NG operation. Similar compensation is used with steam injection NOx control.

A review of expected exhaust gas moisture content reveals the following:

- The exhaust gas moisture content from typical syngas is similar to that from a natural gas fired machine, so there should be no effect on life due to moisture issues. This is similar to the Edwardsport base case.
- A full capture case will have a nearly equal mix of H_2 and N_2 case, producing an exhaust gas with double the moisture content of typical syngas.

In order to maintain HGP parts life, the GT firing temperature must be reduced for cases with higher exhaust gas moisture content (cases with higher levels of CO_2 capture). This has a negative effect on turbine efficiency, but will increase time between maintenance.

3.3.6.6 Materials qualification for hydrogen service

A third area of concern is materials of construction for hydrogen service. GE has specified a limitation on H_2 content at the fuel skid and elected not to use certain of the newer materials normally used for NG service. Additionally, there may be other material issues such as newer coatings that need to be qualified for hydrogen service. A discussion between the plant owner and gas turbine vendor would need to take place to determine whether a qualification program is required for an IGCC plant if CC is retrofitted onto the base plant design.

3.3.6.7 Emissions Technology Discussion

CC cases will typically have lower SO_2 emission rates than non-capture cases, on a unit of mass per thermal input unit basis, due to increased sulfur removal in the AGR due to inclusion of the CO_2 absorption towers in the clean syngas flow path.

Capture case design thinking should start with the CO_2 quality specs required for the pipeline or potential storage situation. H_2S and total sulfur should depend on the specific CO_2 pipeline requirements, which vary widely across the USA. This sets the criteria for the CO_2 removal and purification system which in turn sets the criteria for the syngas constituents going to the gas turbine.

Figure 3-12 shows emission levels for 4 different EPRI CoalFleet capture rate cases compared with a base case (EPRI UDBS Environmental Design 2) without capture. For comparison purposes, the Edwardsport base case design has about 20 ppm sulfur in the syngas to the GT whereas the UDBS Design 2 base case was for 15 ppm.

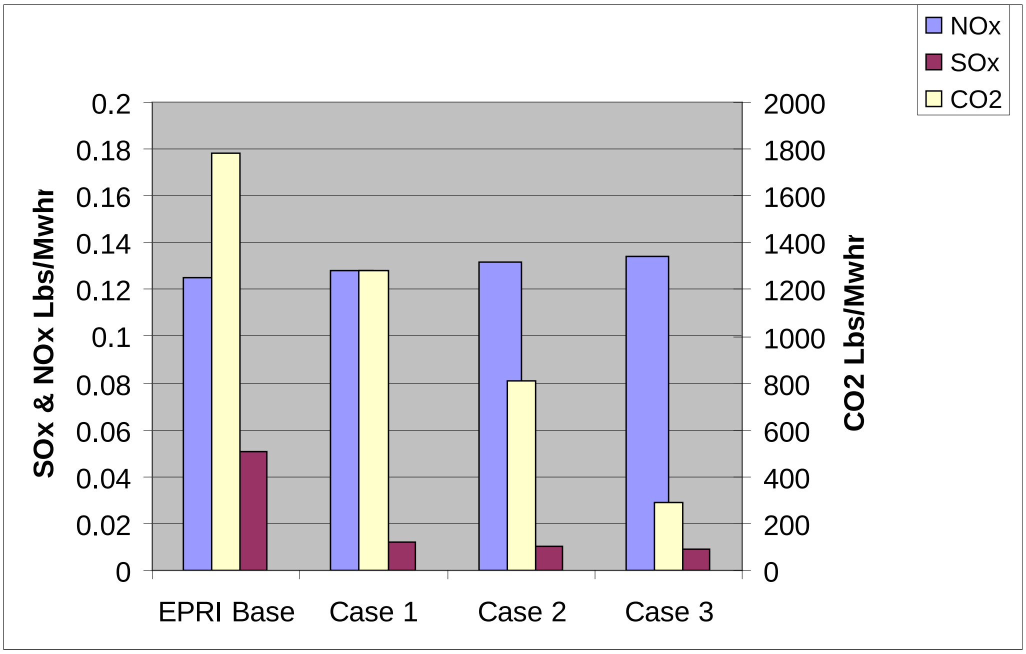
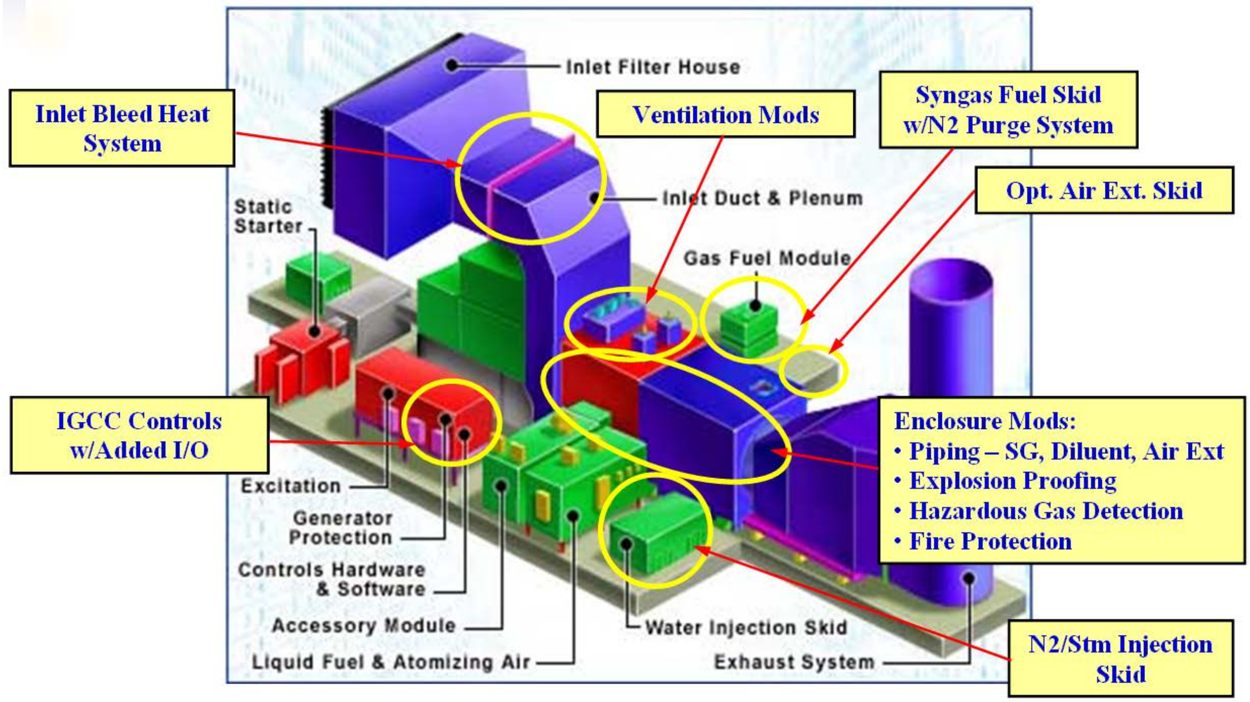


Figure 3-12
Estimated Emissions Results

3.3.6.8 General GT Modifications for IGCC

GEE has provided a sketch of the typical packaging for an IGCC machine. Figure 3-13 shows typical modifications for an IGCC machine such as those provided for Edwardsport. In addition modifications are made to the GT internal parts including the hot gas part and combustors. Each of the modifications should be studied to eliminate or mitigate limitations due to high H_2 fuel prior to the conversion to capture.



Courtesy of GEE

Figure 3-13
Combined Cycle Controls, Accessories and Packaging Changes Needed for Syngas Operations

4

CC CASE 1: UNSHIFTED CO_2 CAPTURE SCENARIO

CC case 1 studies removal of CO_2 from the syngas entering the existing AGR through addition of CO_2 absorption and flash recovery equipment to the Selexol plant. This approach will be employed on both operating trains in the plant and was expected to result in a net reduction of CO_2 emissions between 15-18%, or a nominal 20% reduction. There will be no attempt to alter the syngas characteristics prior to feed into the AGR system. Removing CO_2 from the syngas will result in changes to the clean syngas preparation area and possibly to the ASU-GT interface equipment. These changes will be assessed in this section.

4.1 Overview

In CC case 1, CO_2 absorption, flash recovery and compression equipment was retrofitted onto the Edwardsport IGCC plant model created by EPRI to facilitate removal of CO_2 native to the syngas entering the existing Selexol plant. It was assumed that, when carbon capture was added to the plant, the CO_2 recycle system that, in the plant design, returns CO_2 to the gasification island will be removed from service. CO_2 from high pressure flash recovery within the Selexol unit is recycled to the entrance of the Selexol H_2S absorber instead. It is recommended that this design assumption be discussed with the engineering team during FEED on this plant design.

The gross power output of the plant decreases with inclusion of CO_2 capture equipment. This is due, in part, to changes in the acid gas removal area that result in losses of some combustible gases and some nitrogen to the acid gas and captured CO_2 streams. Also, there are some additional steam demands on the plant, primarily in the Selexol stripper.

4.2 Plant Performance

Table 4-1
CC Case 1 Performance Summary

Plant Output		
Gas Turbine Power	468,400	kW_e
Steam Turbine Power	322,300	kW_e
Total	**790,700**	kW_e
Auxiliary Load		
Coal Handling	390	kW_e
Coal Milling	3,760	kW_e
Coal Slurry Pumps	1,800	kW_e
Slag Handling and Dewatering	1,800	kW_e
Miscellaneous Gasification Island Auxiliaries	2,670	kW_e
Syngas Treatment and Conditioning Line	160	kW_e
Selexol Unit Auxiliaries	10,000	kW_e
SRU Tail Gas Recycle Compressor	1,820	kW_e
Claus Plant/TGTU Auxiliaries	710	kW_e
CO_2 Compression and Dehydration	12,780	kW_e
Air Separation Unit Auxiliaries	6,300	kW_e
Air Separation Unit Main Air Compressor	50,100	kW_e
ASU Boost Compressor	11,140	kW_e
Oxygen Pump	510	kW_e
Nitrogen Compressor	42,230	kW_e
Gas Turbine Auxiliaries	1,000	kW_e
Steam Turbine Auxiliaries	1,500	kW_e
Boiler Feedwater Pump	5,650	kW_e
Condensate Pump	300	kW_e
Circulating Water Pump	9,000	kW_e
Cooling Tower Fans	4,000	kW_e
Miscellaneous Balance-of-Plant	1,000	kW_e
Deep Well Injection	1,600	kW_e
Transformer Losses	1,990	kW_e
Total Auxiliaries	172,210	kW_e
Net Plant Power	618,490	kW_e
Net Plant Efficiency (HHV)	36.9%	
Net Plant Heat Rate (HHV)	9,250	Btu/kWhr
Coal Feed Flowrate	513,111	lb/hr
Thermal Input, HHV[1]	1,676,714	kW_{th}
Condenser Duty	1,690.0	MMBtu/hr

1 - As-Fed Indiana #5 Coal HHV=11,150 Btu/lb

4.3 Evaluation of CC Retrofit Impacts on Edwardsport Plant Design

It should be noted that the analysis was performed only on a single operating condition using the performance coal and at near-ISO conditions. As such, the conclusions drawn in the analysis below may not apply to other operating conditions, fuels in the design envelope, or extreme ambient conditions.

4.3.1 Air Separation Unit

Coal flow for CC case 1 is expected to be approximately 1% higher than base case coal flow. Accordingly, oxygen and nitrogen requirements are each expected to be approximately 1% higher than for the base case. Because the design basis for the cryogenic section of the ASU at 15°C (59˚F) is 9.4% higher than the base, the 1% increase can be accommodated with no changes to the ASU cryogenic section, assuming a small potential constraint on the ASU's ability to support the requirements of the facility at high ambient temperatures and for poor coals can be tolerated. Input into this determination can be quantified by the ASU supplier; e.g., in terms of how high the ambient temperature can be for the plant to still be able to produce the 1% increases in oxygen and nitrogen and to provide the requirements of poor coal cases. Because of the very small additional production required of the ASU, it is expected that any such impacts on ASU performance would be minor.

Approximately 1% more total air flow will be required for the 1% increase in oxygen and nitrogen production. Because the amount of available extraction air is the same as for the base case, all the additional air (8,417 kg/hr or 18,555 lb/hr) will need to be supplied by the MAC. Based on an estimate of the MAC's maximum air flow capacity at 15°C (59˚F) of 621,886 kg/hr (1,371,000 lb/hr), the MAC will be able to provide the additional flow required, with an increase in MAC power consumption of approximately 1.6%. The ASU supplier can be asked to provide the capacity of the base case MAC at 15°C (59˚F) in order to confirm these conclusions.

4.3.2 Gasification Area

Due to the modest increase in coal flow (1%) for CC case 1, it is anticipated that no design changes will be required in the gasification area to accommodate the additional fuel requirements.

4.3.3 Low Temperature Gas Cooling

For Retrofit CC case 1 no change is planned for the LT gas cooling train. The gas side of the exchanger train is therefore unaltered and there is no impact on pressure drop. The utility heat exchange media (MP steam, LP steam, process condensate and cooling water) has for the purpose of this study been assumed to be largely unchanged. This assumption will need to be verified as part of the FEED process.

The reduced mass flow to the gas turbine may have an impact on the operation of the syngas saturator. A reduced amount of circulation water to syngas saturator circulation heater #2 would result in a maximum 10% increased production of LP steam which could probably be accommodated in the existing equipment.

A reduced amount of circulation water to syngas saturator circulation heater #1 would result in increased pre-heat of the condensate in the condensate heater and/or an increased cooling water demand to the trim cooler. If the increased condensate pre-heat leads to too high a temperature, a bypass control around the condensate side of the condensate heater may be required to avoid a condensate temperature, which is too high for a controlled operation of the deaerator.

Whether the trim cooler is adequate for such increased duty needs to be checked against any saturator duty changes.

4.3.4 Acid Gas Removal

For retrofit CC case 1, there is minimal change to the H_2S removal section of the Selexol unit. The feed gas flowrate, composition, and conditions are unchanged. This is because the upstream processing to deliver the synthesis gas to this point has not changed. This means that the quality of the product synthesis gas from the Selexol H_2S absorbers (<20 ppmv) and the quality of the acid gas from the Selexol stripper (>50%V H_2S) will be unaltered.

Referencing the discussion in Section 3.3.4, a new CO_2 removal section has to be added and integrated with the existing H_2S removal section of the Selexol unit:

- There would be two parallel Selexol CO_2 absorbers with feeds coming from the existing Selexol H_2S absorbers.
- A portion of the rich solvents from the CO_2 absorbers would be split off, further cooled by refrigeration, and then used as the pre-loaded lean solvents for the H_2S absorbers.
- The remainder of the rich solvents would be combined, and then processed in a common flash-regeneration section.
- The flashed solvent from the last LP flash drum would be pumped and then split to service the mid-point entries for the two parallel Selexol CO_2 absorbers.
- The lean solvent from the Selexol stripper base will continue to be pumped, equally divided, and cooled via heat exchange and refrigeration, but will proceed to the top entry of the new Selexol CO_2 absorbers
- The static mixer and pre-saturation drum in the Selexol H_2S section can be eliminated from service, as the pre-CO_2-loaded lean solvent is already available from the Selexol CO_2 absorber. The existing loaded solvent chiller can be used to cool this stream prior to entrance into the Selexol H_2S absorber.

In summary, the Selexol process configuration integration of the new CO_2 section with the existing H_2S section is fairly straightforward for CC case 1.

However, there are some *potential* concerns regarding CO_2 purity and CO_2 recovery for CC case 1, which will have to be further examined quantitatively with UOP:

- There is an inherent assumption that there is little flexibility for achieving significantly lower H_2S/COS concentrations in the product gas exiting the existing Selexol H_2S absorber. This is based upon the assumption that there has been limited additional margin designed into the H_2S absorber to permit this. As a result, almost all of the residual H_2S/COS will be captured with the CO_2 in the Selexol CO_2 section. Based upon syngas feeds containing 20-25% by volume CO_2, and H_2S/COS levels approaching 20 ppmv exiting the H_2S absorber, the captured CO_2 product from the Selexol CO_2 section will have total sulfur concentrations approaching or possibly exceeding 100 ppmv. Depending upon levels, this could challenge a 100 ppmv total sulfur specification for the CO_2 product.
- Without CO-shift, the CO_2 concentration in the syngas feed to the Selexol H_2S section will be in the 20-25% by volume range, case dependent. Capturing 90% of this contained CO_2 will require reducing the CO_2 level in the syngas exiting the CO_2 absorber to ~ 2.5% by volume or less. Assuming 90% CO_2 capture is the objective, this may be a difficult goal to achieve, merely because the CO_2 content of the feed gas is limited. The bulk CO_2 removal section of the Selexol CO_2 section can only capture so much CO_2 due to equilibrium limits, rendering the trim CO_2 removal section to achieve the final required levels. This may require an increase in lean solvent flow to the CO_2 absorber top trim section, which will require more capacity from the H_2S stripper section of the unit (lean solvent pumps, reboiler duty, heat exchanger duty, etc). If this capacity is not available, then achieving 90% CO_2 capture may be difficult.

By adding the CO_2 section, and capturing most of the residual H_2S/COS in that section, the final product synthesis gas proceeding to the gas turbines as fuel will have a much reduced total sulfur content (~1 ppmv total sulfur). This is not necessarily an issue, and in fact may be a positive with respect to HRSG performance and plant air emissions.

4.3.5 Claus Unit

For retrofit CC case 1, it is not expected that the acid gas quality to the Claus unit would be affected by the changes to the AGR. Therefore no change is considered necessary or planned for the Claus unit or other sulfur block equipment.

4.3.6 Combined Cycle

The following is a representative list of modifications generally required for IGCC with some comments in italics concerning the specific CC case conversion modifications:

4.3.6.1 Gas Turbine Modifications

- IGCC MNQC combustion system
  - *Likely requires no changes.*
- Combustion lab verification testing program
  - *Final production combustion tests will be required unless they have been performed as part of the base plant testing program.*

- Off-base syngas fuel control module with syngas stop and control valves
  - *Fuel flow mass is considerably reduced but probably within turndown range*
  - *Study required to determine necessary modifications*
- Syngas fuel piping on base and from skid to base.
  - *Piping from skid to base probably remains unchanged as the syngas will contain only 38* H_2 *%. However, much lower flow could affect controllability.*
- Fuel moisturization
  - *Continue to use*
- Nitrogen and steam diluent skid
  - *Unchanged, since* N_2 *flows are very similar*
- Fire and hazardous gas protection system
  - *Unchanged, since only 38%* H_2 *compared to Edwardsport base case of 32%*
- Air extraction and control skid
  - *Study indicates same air extraction*
- Steam injection for NOx control on backup fuel
  - *Assumed to be acceptable within design.*
- Proven HGP Materials instead of single crystal, etc
  - *Existing materials should be acceptable since* H_2 *content of syngas is similar to the base case.*
- Stage 1 turbine nozzle increased area for low-Btu syngas flow
  - *OK due to smaller flow*
- Redundant control and protection system additions for Mark Vle
  - *A holistic study of all operating and emergency modes*
  - *Determine plan for operation if* CO_2 *venting not allowed*
    - *Short term*
    - *Long term*
  - *Add site-specific software*

  - *Minor change in firing temperature reduction curves to maintain natural gas fuel metal temperatures for high hydrogen fuels due to the hydrogen combustion producing excess water in the combustion products.*
  - *Additional control for malfunction of CO_2 capture equipment*
  - *Means of restricting minimum syngas equivalent heating value (Btu/ft^3) into the fuel nozzles*
- Accessory system and enclosure design for syngas.
  - *Modify for above notes*

4.3.6.2 Heat Recovery Steam Generator

The gas turbine exhaust temperature is slightly higher (by 8°F) for this case, but flow is lower so it is likely that no modifications are necessary. The Edwardsport base case HRSG configuration will need to be modeled with each of the CC Cases to determine whether this is the case.

4.3.6.3 Steam Turbine and Condenser

Major steam flows and temperatures are very close to the Edwardsport base case, so no modifications are expected.

5

CC CASE 2: 50% CO_2 CAPTURE SCENARIO

CC case 2 for the Edwardsport IGCC plant studies reduction of the overall CO_2 emissions from the plant by approximately 50%. The targeted CO_2 removal rate is such that the emissions of the plant will be 363 kg/MWhr (800 lb/MWhr), net basis, in order to match that which would be expected from a current, state-of-the-art natural gas combined cycle. This will require installation of a single water gas shift reactor in the low temperature syngas cooling train to convert some carbon monoxide (CO) in the syngas to CO_2. This modification may have other impacts on the plant such as the need to redesign the low temperature syngas cooling train, addition of CO_2 absorption and flash recovery equipment sized larger than required for CC case 1 and further impacts to the ASU-GT interface equipment. These impacts will be evaluated in this section.

5.1 Overview

In CC case 2, water gas shift, CO_2 absorption, flash recovery and compression equipment was retrofitted onto the Edwardsport IGCC plant model created by EPRI to facilitate removal of CO_2 native to the syngas entering the existing Selexol plant. It was assumed that, when carbon capture was added to the plant, the CO_2 recycle system that, in the plant design, returns upstream, will be removed from service. CO_2 from high pressure flash recovery within the Selexol unit is recycled to the entrance of the Selexol H_2S absorber instead. This design assumption would need to be evaluated by the design team during ensuing engineering on this plant retrofit option.

The gross power output of this case is slightly higher than the base case design. This is primarily traced to additional steam turbine power output in the water gas shift area; however the result requires further investigation as some power loss had been expected. Further evaluation will be required in conjunction with the plant design team in order to more accurately estimate the power output effects of this level of CO_2 capture.

5.2 Plant Performance

Table 5-1
CC Case 2 Performance Summary

Plant Output		
Gas Turbine Power	468,400	kW_e
Steam Turbine Power	327,300	kW_e
Total	**795,700**	kW_e
Auxiliary Load		
Coal Handling	400	kW_e
Coal Milling	3,870	kW_e
Coal Slurry Pumps	1,850	kW_e
Slag Handling and Dewatering	1,850	kW_e
Miscellaneous Gasification Island Auxiliaries	2,760	kW_e
Syngas Treatment and Conditioning Line	200	kW_e
Selexol Unit Auxiliaries	17,000	kW_e
SRU Tail Gas Recycle Compressor	1,950	kW_e
Claus Plant/TGTU Auxiliaries	710	kW_e
CO_2 Compression and Dehydration	21,610	kW_e
Air Separation Unit Auxiliaries	6,400	kW_e
Air Separation Unit Main Air Compressor	63,040	kW_e
ASU Boost Compressor	11,470	kW_e
Oxygen Pump	500	kW_e
Nitrogen Compressor	43,040	kW_e
Gas Turbine Auxiliaries	1,000	kW_e
Steam Turbine Auxiliaries	1,500	kW_e
Boiler Feedwater Pump	5,630	kW_e
Condensate Pump	320	kW_e
Circulating Water Pump	9,480	kW_e
Cooling Tower Fans	4,210	kW_e
Miscellaneous Balance-of-Plant	1,000	kW_e
Deep Well Injection	1,600	kW_e
Transformer Losses	2,000	kW_e
Total Auxiliaries	203,390	kW_e
Net Plant Power	592,310	kW_e
Net Plant Efficiency (HHV)	34.3%	
Net Plant Heat Rate (HHV)	9,950	Btu/kWhr
Coal Feed Flowrate	528,554	lb/hr
Thermal Input, HHV[1]	1,727,179	kW_{th}
Condenser Duty	1,780.0	MMBtu/hr

1 - As-Fed Indiana #5 Coal HHV=11,150 Btu/lb

5.3 Evaluation of CC Retrofit Impacts on Edwardsport Plant Design

It should be noted that the analysis was performed only on a single operating condition using the performance coal and at near-ISO conditions. As such, the conclusions drawn in the analysis below may not apply to other operating conditions, fuels in the design envelope, or extreme ambient conditions.

5.3.1 Air Separation Unit

Coal flow for this case is expected to be 4% higher than base case coal flow. Accordingly, oxygen and nitrogen requirements are each expected to be approximately 4% higher than for the base case. Because the design basis for the cryogenic section of the ASU at 15°C (59˚F) is 9.4% higher than the base, the 4% increase can be accommodated with no changes to the ASU cryogenic section, assuming some potential constraint on the ASU's ability to supply the requirements of the facility at high ambient temperatures and for poor coals can be tolerated.

Inputs into this determination can be quantified by the ASU supplier; e.g., in terms of how high the ambient temperature can be for the plant to still be able to produce the 4% increases in oxygen and nitrogen. Approximately 9% of the diluent nitrogen stream will need to be delivered the gas turbine fuel line upstream of the gas turbine skid main fuel control valve, thus requiring a higher delivery pressure on this portion of the diluent flow.

The amount of extraction air available for this case is estimated to be 114,307 kg/hr (252,000 lb/hr) less than for the base case. In addition, 3.9% more total air flow will be required for the 3.9% increase in oxygen and nitrogen production. These two factors determine the total compressed air need. Based on an estimate of the MAC's maximum air flow capacity at 15°C (59˚F) of 621,432 kg/hr (1,370,000 lb/hr), the supplemental air compression (SMAC) capacity required is estimated to be 64,982 kg/hr (143,259 lb/hr). Either one SMAC of that capacity to supply both trains, or one half-sized SMAC for each train can be added to supply the additional compressed air. The ASU supplier can be asked to provide the capacity of the base case MAC at 15°C (59˚F) in order to generate a more accurate estimate of SMAC requirements.

5.3.2 Gasification Area

Due to the modest increase in coal flow (~4%) for CC case 2, it is anticipated that no design changes will be required in the gasification area to accommodate the additional fuel requirements.

5.3.3 Low Temperature Gas Cooling

The design basis for this case is that the equipment should be capable of being used for CC case 3. In particular, the shift reactor should be designed as a two stage reactor, but the filling omitted. For this reason, the reader is referred to CC case 3 for a full description of the effect of the shift on the plant configuration. At this point in the report only differences between the two Options are discussed.

The main process characteristics of the CC case 2 are tabulated in Table 5-2.

Table 5-2
Main Water-Gas Shift Process Characteristics of CC Case 2

	SOR	Intermediate	EOR
Feed gas CO, kg-mol/h (lb-mol/h)	4,017 (8,855)	4,017 (8,855)	4,017 (8,855)
Feed gas CO, mol% (dry basis)	38.91	38.91	38.91
Shift gas CO, kg-mol/h (lb-mol/h)	944 (2,085)	1,027 (2,264)	1,110 (2,446)
Shift gas CO, mol% (dry basis)	6.72	7.34	7.98
Overall conversion *	76.6%	74.6%	72.6%
LP Steam production, kg/hr (lb/hr)	17,408 (38,377)	17,389 (38,335)	17,372 (38,297)

* calculated ignoring CO contained in recycle gas.

5.3.3.1 Equipment

This case assumes that all the equipment from CC case 3 is in place, but only the catalyst on the second stage of the CO shift reactor is not loaded. Under these conditions, IP steam generator #2 does not generate any steam and could be omitted.

5.3.3.2 Other Issues

Depending on details of implementation, the additional pressure drop for adding a shift would be slightly less than for CC case 3. This has not been evaluated in detail, but is expected to be about 20 psi less.

5.3.4 Acid Gas Removal

For CC case 2, one operating stage of CO-Shift is applied to each of the two parallel synthesis gas streams. The gas compositions and flows exiting each CO-Shift Section for SOR and EOR are given below.

Table 5-3
Start-of-Run Water-Gas Shift Reactor Outlet Composition

Unit:	**CO Shift**				
	Stream	**KO Drum Outlet – Start-of-Run (SOR)**			
	Phase		vapour		
Component	**[kg/kmol]**	**lb-mol/hr**	**kg-mol/hr**	**mol% (dry)**	**mol% (wet)**
CO_2	44.010	12350.78	5601.26	**40.022**	**39.969**
CO	28.010	2084.26	945.24	**6.754**	**6.745**
H_2	2.016	14921.10	6766.94	48.351	48.287
CH_4	16.043	6.59	2.99	0.021	0.021
CH_3OH	32.041	0.00	0.00	0.000	0.000
N_2	28.013	594.99	269.84	1.928	1.925
O_2	32.000	0.00	0.00	0.000	0.000
Ar	39.948	254.73	115.52	0.825	0.824
H_2S	34.082	576.47	261.44	**1.868**	**1.866**
COS	60.076	4.43	2.01	**0.014**	**0.014**
SO_2	64.066	0.06	0.03	0.000	0.000
NH_3	17.031	66.24	30.04	0.215	0.214
HCN	27.026	0.00	0.00	0.000	0.000
H_2O	18.016	41.04	18.61	0.133	0.133
Carbon	[kg/h]				
Ash	[kg/h]				
Total (dry):		**30859.653**	13995.31	100.00	
Total (wet)		**30900.694**	14013.92	100.13	100.00
H_2O	[kg/h]		32		
Pressure	[psia/bar a]	516.50	35.62		
Temperature	[°F/°C]	90.61	33		
Molar Flow	[kmol/h]		14014		
Mass Flow	[kg/h]		308731		

Table 5-4
Start-of-Run Water-Gas Shift Reactor Outlet Composition

Component	[kg/kmol]	lb-mol/hr	kg-mol/hr	mol% (dry)	mol% (wet)
Norm. vol flow	[Nm^3/h dry]		313691		
Molar Mass	[kg/kmol]		22.03		
Vapor fract.	[%]		1.00		

Table 5-5
End-of-Run Water-Gas Shift Reactor Outlet Composition

Unit:	**CO Shift**				
	Stream	**KO Drum Outlet – End-of-Run (EOR)**			
	Phase		vapour		
Component	**[kg/kmol]**	**lb-mol/hr**	**kg-mol/hr**	**mol% (dry)**	**mol% (wet)**
CO_2	44.010	11987.59	5436.55	**39.318**	**39.266**
CO	28.010	2445.42	1109.03	8.021	8.010
H_2	2.016	14559.83	6603.10	47.755	47.692
CH_4	16.043	6.59	2.99	0.022	0.022
CH_3OH	32.041	0.00	0.00	0.000	0.000
N_2	28.013	594.99	269.83	1.952	1.949
O_2	32.000	0.00	0.00	0.000	0.000
Ar	39.948	254.72	115.52	0.835	0.834
H_2S	34.082	569.11	258.10	**1.867**	**1.864**
COS	60.076	5.02	2.27	**0.016**	**0.016**
SO_2	64.066	0.06	0.03	0.000	0.000
NH_3	17.031	65.18	29.56	0.214	0.214
HCN	27.026	0.00	0.00	0.000	0.000
H_2O	18.016	40.55	18.39	0.133	0.133
Carbon	[kg/h]				

Table 5-6
End-of-Run Water-Gas Shift Reactor Outlet Composition

Component	[kg/kmol]	lb-mol/hr	kg-mol/hr	mol% (dry)	mol% (wet)
Ash	[kg/h]				
Total (dry):		**30488.515**	13826.99	100.00	
Total (wet)		**30529.068**	13845.38	100.13	100.00
H_2O	[kg/h]		32		
Pressure	[psia/bara]	516.50	35.62		
Temperature	[°F/°C]	90.61	33		
Molar Flow	[kmol/h]		13845		
Mass Flow	[kg/h]		305630		
Norm. vol flow	[Nm³/h dry]		309918		
Molar Mass	[kg/kmol]		22.07		
Vapor fract.	[%]		1.00		

The changes in synthesis gas feed composition and conditions are significant, and will require modifications to the Selexol H_2S Section. These changes will have to be examined by UOP to define them quantitatively, but they involve the following areas:

- Assuming there is little margin in the two existing Selexol H_2S Absorbers, a third H_2S absorber plus associated equipment will have to be installed to accommodate the 30% increase in syngas flow. The three H_2S absorbers will have greater turndown restrictions, since they will no longer be operating at full load at 100% facility capacity.
- This will also bring about challenging control issues
  - The two syngas streams exiting the upstream CO-shift and cooling sections will have to be divided into three streams feeding the three H_2S absorbers (two existing and one new)
  - The three de-sulfurized syngas streams exiting the H_2S absorbers will have to be “combined” into two equal syngas streams feeding the two new CO_2 absorbers in the new Selexol CO_2 section
  - When a gasifier is taken offline, the syngas from one line of CO-shift will have to be split into two streams feeding two H_2S absorbers, and then recombined to feed a single CO_2 absorber

An alternate way to address this issue is to add a new Selexol H_2S absorber to each train, bringing up the total number of Selexol H_2S absorbers to four. This added investment may be compensated for by the reduction in control complexity that was just described. This issue of how best to add Selexol H_2S absorbers will have to be reviewed in detail during the FEED stage.

- The CO_2:H_2S ratio in the CO-shifted syngas stream has increased from ~ 9/1 to ~ 21/1 for start-of-run. The more-than-doubling of the CO_2:H_2S ratio in one of the syngas feeds will tax the Selexol H_2S section and its ability to maintain a H_2S concentration in the acid gas greater than 50% by volume. This may require the addition of a H_2S concentrator tower (see Section 3.3.4), which would replace the rich solvent flash drum in the base case configuration. A slipstream of either N_2 or product syngas from the CO_2 absorber would be used as the inert stripping agent to back-strip CO_2 from the rich solvent.

  In the base case configuration, however, the flash gas stream generated is a particular large flow. This large flow enables the existing rich solvent flash drum to act more like a H_2S concentrator, which may allow its continued use instead of adding a H_2S concentrator tower. This is viewed as a limited probability, but only a detailed evaluation by UOP can deny or confirm this. Even if that analysis shows the rich solvent flash drum can continue to perform, it is almost certain that there will be deterioration in acid gas quality. This can lead to reductions in capacity of the H_2S stripping section of the Selexol process and therefore the overall AGR, and/or can have an impact upon the performance and capacity of the sulfur removal unit.

- The addition of one stage of CO-shift on each of the two syngas streams presumes the elimination of the COS hydrolysis unit operation. However, reviewing the SOR and EOR feedgas compositions in the tables shows COS effluent concentrations of 140-160 ppmv, which is more than a tripling of the base case concentration (55 ppmv). This higher COS concentration will burden the Selexol H_2S section to achieve an effluent product gas concentration of less than 20 ppmv total sulfur. Consequently, there are three scenarios that may occur, which need to be considered in a future stage of engineering:

  - Retain the COS hydrolysis unit in operation <u>downstream</u> of the CO-Shift
  - If more than 20 ppmv total sulfur exits the H_2S absorber, it will be mostly captured in the Selexol CO_2 absorption section, and will exit with the CO_2 product. Depending upon levels, this could exceed the estimated CO_2 product purity specification limit on total sulfur (100 ppmv).
  - If less than 20 ppmv total sulfur exits the H_2S absorber at the expense of greater CO_2 capture in the H_2S absorber, this greater CO_2 load will tax the H_2S stripping section, leading to potential capacity limitations of that section and therefore of the overall AGRU, and/or poorer quality acid gas and/or can have an impact upon the performance and capacity of the sulfur recovery unit.

- The static mixer and pre-saturation drum in the Selexol H_2S section can be eliminated from service, as the pre-CO_2-loaded lean solvent will already be available from the Selexol CO_2 absorber. The existing loaded solvent chiller can be used to cool this stream prior to entrance into the Selexol H_2S absorber.

Referencing the discussions in Sections 3.3.4 & 4.3.4, a new CO_2 removal section has to be added and integrated with the modifications to the existing H_2S removal section of the Selexol unit. This CO_2 removal section will have to be oversized to accommodate the greater CO_2 load for CC case 3.

5.3.5 Claus Unit

The Claus unit in the base case is specified as part of the sulfur block package as discussed in Section 3.3.5.

Issues which may arise in connection with the introduction of carbon capture with CO shift are largely associated with any increased CO_2 content in the acid gas from the AGR compared with the base case, which in turn would depend on the extent of acid gas concentration in the modified AGR.

- For high sulfur fuels, it is unlikely to be a problem from the point of view of H_2S concentration. However added CO_2 would increase the gas volumes. This would require a review of the hydraulics of the unit as well as any possible limitations on the capacity of the tail gas recycle compressor. These issues must be addressed during the FEED process.
- For low sulfur fuels, the gas quality will be close to the lower limit for efficient sulfur recovery. Any additional dilution with additional CO_2 will only aggravate this situation further. In the turndown case, support gas may be required to maintain the furnace temperature, even if the added CO_2 volumes would ameliorate turndown issues associated with exit velocity at the burner.

A separate issue which may impinge on the Claus plant in the low sulfur is the minimum sulfur requirement of the CO shift catalyst. If this is an issue, then acid gas recycle (or partial H_2S recycle) may be required. This could be addressed by installing a bypass around the Claus unit to the tail gas recycle compressor.

5.3.6 Combined Cycle

The following is a representative list of modifications generally required for IGCC with some comments in italics concerning the specific CC case conversion modifications:

5.3.6.1 Gas Turbine Modifications

- IGCC MNQC combustion system
  - *Explore the potential need to change fuel nozzles and evaluate configurations including a performance heater to adjust fuel temperatures to meet the necessary Wobbe Index range.*
- Combustion lab verification testing program
  - *Final production combustion tests will be required unless they were performed during the base plant testing program.*
- Off-base syngas fuel control module with syngas stop and control valves
  - *Fuel and diluent flow mass is considerably reduced (61%) even though some* N_2 *has been injected into the fuel stream to meet the estimated fuel specification requirement of the gas turbine supplier*
  - *Determine method of mixing* N_2
  - *Provide a fuel blending skid*
  - *Perform a study of necessary modifications*

  - *Check for controllability*
- Syngas fuel piping on base and from skid to base.
  - *Addition or replacement for high H_2 fuel manifolds on base design*
  - *Piping from skid to base will likely remain unchanged*
- Fuel moisturization
  - *May not be needed after conversion but can be used to optimize performance and adjust flows for more air extraction as desired*
- Nitrogen and steam diluent skid
  - *Split diluent nitrogen supply into two streams*
  - *New piping and valves for diluent N_2 to fuel skid at approximately 0.2 MPa (29 psi) higher pressure to meet restriction on minimum syngas heating value. This may have an effect on the ASU design.*
  - *N_2 flow to combustor is only 4% less so controllability should remain*
  - *Study of N_2 supply adequacy for fuel switches and unplanned outages*
- Fire and hazardous gas protection system
  - *Determine adequacy of syngas protection for high hydrogen*
  - *Need hydrocarbon, H_2, CO, CO_2, and UV detection and ventilation*
- Air extraction and control skid
  - *Study base case skid and piping for operation and controllability at 0-50% of base case flow*
- Steam injection for NOx control on backup fuel
  - *Should be acceptable within current design limits*
- Proven HGP Materials instead of single crystal, etc
  - *Consider H_2 material qualification program, especially for coatings.*
- Stage 1 turbine nozzle increased area for low-Btu syngas flow
  - *Existing design should be acceptable due to smaller flow.*
  - *Check coatings for H_2 compatibility*
- Redundant control and protection system additions for Mark Vle
  - *A holistic study of all operating and emergency modes*
  - *Determine plan for operation if CO_2 venting not allowed*
    - *Short term*
    - *Long term*
  - *Add site-specific software*

  - *New firing temperature reduction curves to maintain natural gas fuel metal temperatures for high H_2 fuels due to hydrogen producing excess water in the combustion products. These would in theory be different for CC cases 1, 2 and 3 with a lower firing temperature as H_2 increases.*
  - *Additional control for malfunction of CO_2 capture equipment*
  - *Means of restricting minimum syngas equivalent heating value (Btu/ft^3) into the fuel nozzles*
- Accessory system and enclosure design for syngas.
  - *Modify for above notes*
- Heat Recovery Steam Generator
  - Gas turbine exhaust temperature and steam flows are lower; however, the Edwardsport base case HRSG configuration will need to be modeled with each of the CC cases 1, 2 & 3 data to determine whether any modifications are necessary
- Steam Turbine & Condenser
  - Main steam flows are lower for this case. A study was made concerning the use of the significant amount of steam coming from the shift reactors. One option considered doubling the flow of the 0.4 MPa (56 psi) LP injection and this was discarded due to potential restrictions of the existing piping and potential for changing the thrust balance as the injection appears to enter only one side of the LP units. A second option of admission at 0.6 MPa (84 psi) to the LP crossover flow appears to be a better approach.

5.3.7 Other Process Areas

The increased dry gas volume leaving the low temperature gas cooling section will increase the pressure drop across the mercury guard drum. If required, this could be ameliorated by installing an additional drum in parallel.

6

CC CASE 3: FULL (80-90%) CO_2 CAPTURE SCENARIO

CC case 3 studies the impact of retrofitting full CO_2 capture onto the Edwardsport IGCC plant. In this case, the amount of CO_2 in the syngas is increased in two stages of water gas shift reactors and the resulting syngas enters the AGR, where CO_2 removal may occur. The overall CO_2 emissions from the plant in this case may be reduced by 80-90%. This will require installation of a two water gas shift reactors, in series, in the low temperature syngas cooling train to convert carbon monoxide (CO) in the syngas to CO_2. This modification will have significant impacts on the plant such as the need to redesign the low temperature syngas cooling train, addition of CO_2 absorption and flash recovery equipment sized larger than required for CC case 1 and 2, further impacts to the ASU-GT interface equipment, and the need to redesign the gas turbine fuel inlet area. These and other hitherto unidentified impacts will be evaluated in this section.

6.1 Overview

In CC case 3, water gas shift, CO_2 adsorption, flash recovery and compression equipment was retrofitted onto the Edwardsport IGCC plant model created by EPRI to facilitate removal of CO_2 native to the syngas entering the existing Selexol plant. It was assumed that, when carbon capture was added to the plant, the CO_2 recycle system that, in the plant design, returns upstream, will be removed from service. CO_2 from high pressure flash recovery within the Selexol unit is recycled to the entrance of the Selexol H_2S absorber instead. It is recommended that this design assumption be discussed with the engineering during ensuing engineering on this plant design.

The gross power output of the plant decreases with inclusion of CO_2 capture equipment. This is due, in part, to changes in the acid gas removal area that result in losses of some combustible gases and some nitrogen to the acid gas and captured CO_2 streams. Also, there are some additional steam demands on the plant, primarily in the Selexol stripper. Further, there is some loss in steam turbine efficiency since the gas turbine exhaust temperature is quite lower than in the base case, resulting in reduction of throttle steam temperature for the steam turbine. This decreases the power output of the unit.

6.2 Plant Performance

Table 6-1
CC Retrofit Case 3 (80-90% CC) Case Performance Summary

Plant Output		
Gas Turbine Power	468,300	kW_e
Steam Turbine Power	319,800	kW_e
Total	**788,100**	kW_e
Auxiliary Load		
Coal Handling	410	kW_e
Coal Milling	3,930	kW_e
Coal Slurry Pumps	1,880	kW_e
Slag Handling and Dewatering	1,880	kW_e
Miscellaneous Gasification Island Auxiliaries	2,800	kW_e
Syngas Treatment and Conditioning Line	200	kW_e
Selexol Unit Auxiliaries	20,000	kW_e
SRU Tail Gas Recycle Compressor	1,720	kW_e
Claus Plant/TGTU Auxiliaries	710	kW_e
CO_2 Compression and Dehydration	35,120	kW_e
Air Separation Unit Auxiliaries	6,500	kW_e
Air Separation Unit Main Air Compressor	67,690	kW_e
ASU Boost Compressor	11,640	kW_e
Oxygen Pump	510	kW_e
Nitrogen Compressor	43,230	kW_e
Gas Turbine Auxiliaries	1,000	kW_e
Steam Turbine Auxiliaries	1,500	kW_e
Boiler Feedwater Pump	5,590	kW_e
Condensate Pump	310	kW_e
Circulating Water Pump	9,320	kW_e
Cooling Tower Fans	4,140	kW_e
Miscellaneous Balance-of-Plant	1,000	kW_e
Deep Well Injection	1,600	kW_e
Transformer Losses	1,980	kW_e
Total Auxiliaries	224,660	kW_e
Net Plant Power	563,440	kW_e
Net Plant Efficiency (HHV)	32.1%	
Net Plant Heat Rate (HHV)	10,614	Btu/kWhr
Coal Feed Flowrate	536,344	lb/hr
Thermal Input, HHV[1]	1,752,633	kW_{th}
Condenser Duty	1,750.0	MMBtu/hr

1 - As-Fed Indiana #5 Coal HHV=11,150 Btu/lb

6.3 Evaluation of CC Retrofit Impacts on Edwardsport Plant Design

It should be noted that the analysis was performed only on a single operating condition using the performance coal and at near-ISO conditions. As such, the conclusions drawn in the analysis described in this section may not apply to other operating conditions, fuels in the design envelope, or extreme ambient conditions.

6.3.1 Air Separation Unit

Coal flow for this case is expected to be 5.5% higher than base case coal flow. Accordingly, oxygen and nitrogen requirements are each expected to be approximately 5.5% higher than for the base case. Because the design basis for the cryogenic section of the ASU at 15°C (59˚F) is 9.4% higher than the base, the 5.5% increase can be accommodated with no changes to the ASU cryogenic section, assuming some potential constraint on the ability of the ASU to supply the requirements of the facility at high ambient temperatures and for poor coals can be tolerated.

Input into this determination can be quantified by the ASU supplier; e.g., in terms of how high the ambient temperature can be for the plant to still be able to produce the 5.5% increases in oxygen and nitrogen. Approximately 33% of the diluent nitrogen stream will need to be delivered to the gas turbine fuel line upstream of the main control valve, thus requiring a higher delivery pressure on this portion of the diluent flow.

The amount of extraction air available for this case is estimated to be 151,502 kg/hr (334,000 lb/hr) less than for the base case. In addition, 5.4% more total air flow will be required for the 5.4% increase in oxygen and nitrogen production. These two factors determine the total compressed air need. Based on an estimate of the MAC's maximum air flow capacity at 15°C (59˚F) of 621,432 kg/hr (1,370,000 lb/hr), the supplemental main air compression (SMAC) capacity required is estimated to be 115,614 kg/hr (254,880 lb/hr). Either one SMAC of that capacity to supply both trains, or one half-sized SMAC for each train can be added to supply the additional compressed air. The ASU supplier can be asked to provide the capacity of the base case MAC at 15°C (59˚F) in order to generate a more accurate estimate of SMAC requirements.

NOTE: To estimate the MAC maximum air flow capacity at 15°C (59˚F), it was assumed that the ASU design was based on 96% oxygen recovery (i.e., 96% of oxygen in the feed air ends up in the product oxygen stream). It was further assumed, with some guidance from the project ASU specification, that the maximum air case for the ASU set the HP limit and the resulting mass flow was used. As noted earlier, because the SMAC estimates depend directly on these estimates, the actual MAC capacity should be requested from the ASU supplier, in order to confirm or modify the SMAC estimates.

6.3.2 Gasification Area

Although the coal flow for CC case 3 is ~6% higher than the base case, it is anticipated that no design changes will be required in the gasification area to accommodate the additional fuel requirements. This is due to the coal flow of CC case 3 being lower than the design flow from the GE design basis document.

6.3.3 Low Temperature Gas Cooling

6.3.3.1 Process Description

A two stage CO-shift unit has been installed immediately downstream of the scrubber. Details of the pre-FEED design are shown in Figure 6-1. For the base case the gas leaving the scrubber is saturated with water vapor at 4.1 MPa (590 psia). The water loading is sufficient to operate the shift. Other gasifier operating cases do not have such a high water loading and therefore provision is made to add steam to bring up the total water loading to that required. During normal operation the saturated gas is preheated to the starting temperature of the catalytic reaction of about 282°C (540°F) in the shift feed-effluent exchanger. This also provides more than sufficient superheat to avoid capillary condensation on the catalyst. During start up, when the reaction is not yet providing the heat for this task, a start up heater operating on high pressure steam is required.

The preheated gas enters the CO shift reactor #1 where the carbon monoxide and steam react to form carbon dioxide and hydrogen according to the reaction

$$CO + H_2O \leftrightarrow H_2 + CO_2 \qquad \Delta h = -41 \text{ MJ/kmol.}$$

In the reactor the CO content of the gas is reduced over fresh catalyst from 39% to 7% (dry basis), equivalent to a 75% conversion. The endothermic reaction raises the gas temperature to 433°C (812°F) at the reactor outlet.

The gas is cooled to about 144°C (260°F) in the feed-effluent exchanger and subsequent IP steam generator, raising 3.8 MPa (555 psia) saturated steam. The gas can now be introduced into the CO shift reactor #2 where the CO content of the gas is reduced further to 2% (dry basis) for an overall conversion for both stages of 93%. The gas temperature at the reactor outlet is about 299°C (570°F), which is suitable for generating further quantities of IP steam. Further gas cooling to about 193°C (380°F) is used to generate LP (0.8 MPa [112 psia]) steam. Subsequently, the gas is cooled down to 128°C (262°F) in exchange with syngas saturator circulation water. This temperature is below the dew point and the process condensate is separated from the gas in knock out drum #1. Final cooling down to 33°C (91°F) takes place in the water cooled trim cooler.

6.3.3.2 Process Assumptions

The process calculations have be made with a number of assumptions that will need further verification in the FEED, but which nonetheless provide a first order degree of accuracy. In particular:

- The duty of the syngas saturator circulation water pre-heaters is as for the base case.
- The duty of the steam condensate preheat is as for the base case.
- IP steam is generated at 3.8 MPa (555 psia), which will allow it to be fed to the reheater of the HRSG together with the HP turbine exhaust steam. Alternatively all steam can be generated as LP steam to make a better match with the existing HRSG.
- Heat exchanger approach temperatures have been modeled generously to reduce capital cost. Opportunity may exist to increase steam make at increased plant capital cost.

Furthermore it must be recognized that the catalyst performance will deteriorate with age. The data in the process description above has been calculated with fresh catalyst (start-of-run or SOR). Calculation have been made by the EPRI team based on best current available knowledge for aged catalyst (end-of-run or EOR) and with an intermediate assumption. The results, including steam makes, are tabulated in Table 6-2.

Table 6-2
Water-Gas Shift Catalyst Performance Data for Start-of-Run, Intermediate, and End-of-Run Conditions

	SOR	Intermediate	EOR
Feed gas CO, kg-mol/h (lb-mol/h)	4,017 (8,855)	4,017 (8,855)	4,017 (8,855)
Feed gas CO, mol% (dry basis)	38.91	38.91	38.91
Shift gas CO, kg-mol/h (lb-mol/h)	290 (640)	320 (705)	418 (921)
Shift gas CO, mol% (dry basis)	1.97	2.18	2.86
Overall conversion *	92.8%	92.1%	89.79
LP Steam production, kg/hr (lb/hr)	17,558 (38,708)	17,551 (38,693)	17,528 (38,643)

* calculated ignoring CO contained in recycle gas.

6.3.3.3 Equipment

The following review of equipment has been made at a high level without benefit of detailed engineering. For this reason the comments are based solely on a comparison of the process flowsheet data for various cases in the study. The indications of reusability listed here must be investigated in more detail during the FEED.

Additionally the comments below take no account of the layout of the base plant or the location of the new CO-shift unit. The potential for reuse of equipment may be constrained by excessively long pipe runs or the time taken to relocate them from their present location to the new unit. This aspect must also be investigated during the FEED.

Figure 6-1
CC Retrofit Case 3 Low Temperature Syngas Cooling Area Process Flow Diagram

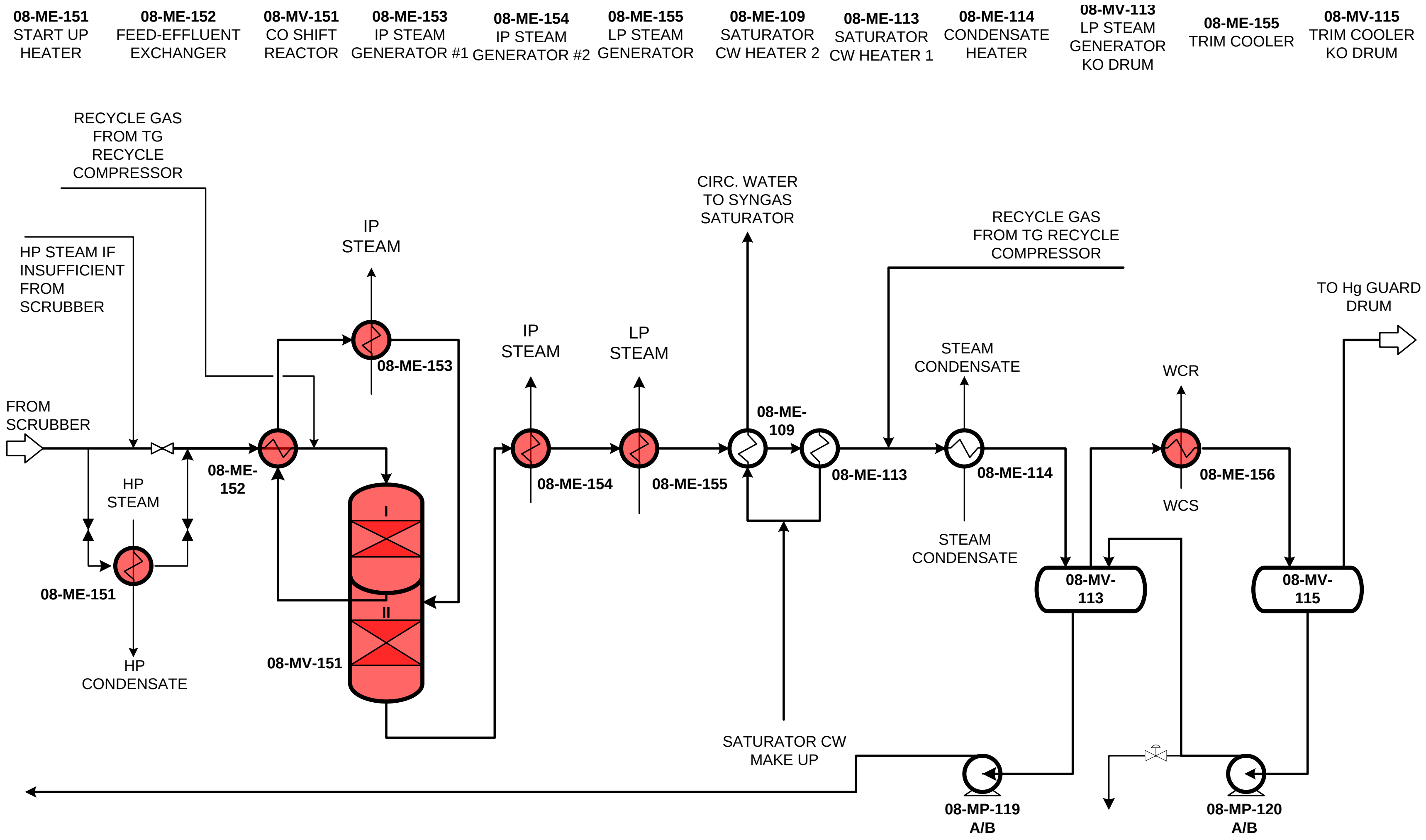


Table 6-3
Existing Equipment

08-ME-109	**Saturator Circ. Water Heater #2**. It is expected that this exchanger can be reused, albeit at a different location in the LT cooling train.
08-ME-110	**LP Steam Generator**. While the gas side hydraulics of this exchanger might be satisfactory for reuse, the surface area will be insufficient for the retrofitted plant. It is expected that a total replacement will be the more economical solution in this case, but this will need confirmation during the FEED.
08-ME-111	**COS Hydrolysis Preheater**. There is no use for this exchanger in the retrofitted plant. Even if the steam side were designed for high pressure steam, it is unlikely that there will be sufficient surface to perform the duty required of the CO Shift start-up heater.
08-ME-113	**Saturator Circ. Water Heater #1**. It is expected that this exchanger can be reused.
08-ME-114	**Steam Condensate Preheater**. It is expected that this exchanger can be reused. There may be a limitation on the gas side hydraulics, in which case a small unit in parallel to the existing exchanger is likely to be the most economic solution.
08-ME-115	**Trim Cooler**. The trim cooler is too small for reuse, both from the point of view of gas hydraulics and of surface area. Though no shown on the process flowsheet, there may be benefit to inserting an air cooler upstream of the trim cooler to reduce the gas temperature to about 54°C (130°F). In this case a small trim cooler in parallel to the existing one may be a good solution.
08-MV-113	**KO Drum #1**. Can be reused
08-MV-114	**COS Hydrolysis Reactor.** Cannot be reused. The materials of construction are probably inappropriate to the higher temperatures and higher H_2 partial pressures of the CO-shift reaction. The volume has not been checked.
08-MV-115	**KO Drum #2**. Can be reused
08-MP-119 A/B	**KO Drum #1 Pumps**. Can be reused
08-MP-120 A/B	**KO Drum #1 Pumps**. Can be reused

Table 6-4
New equipment

08-ME-151	**CO Shift Start up Heater**.
08-ME-152	**CO Shift Feed Effluent Exchanger**.
08-ME-153	**CO Shift LP Steam Generator #1**. Rate for SOR conditions
08-MV-151	**CO Shift Reactor**. Divided into sections 1 and 2.
08-ME-154	**CO Shift LP Steam Generator #2**. Rate for EOR conditions
08-ME-155	**CO Shift LP Steam Generator**.
08-ME-156	**Trim Cooler.** Review splitting into air cooler and water cooler if no alternative useful heat sink available. Cooling water temperature to be checked.

6.3.3.4 Other Issues

- Introducing a two-stage shift increases the pressure drop over the LT cooling system by about 0.2 MPa (30 psi) from 0.3 MPa (42 psi) to 0.5 MPa (~73 psi). The introduction of a CO_2 removal stage to the AGR will introduce even further pressure drop into the system. It will be necessary to address this matter during the FEED stage. Possible approaches could include the following:
  - Review the extent to which the increased pressure drop can be accommodated with the existing or a modified gas turbine control system.
  - Operate the gasifier at a higher pressure. Margins could be reduced by exchanging spring loaded relief valves for pilot operated valves.
  - Further optimization of the retrofit equipment to reduce pressure drop at a possible increase of capital cost.
  - In practice, one must assume that a combination of all three approaches will be required.
- Other operating cases will need to be checked during the FEED to determine whether more steam is required to meet the operating requirements of the CO-shift reactors (e.g. a high CO case from the gasifier or a low temperature case from the scrubber). This will involve reviewing impacts on the steam cycle. The necessary control valve for such steam supply should be included in any case as precaution.
- Note that the CO shift reaction is a consumer of steam even if no additional steam is added to the scrubber exit gas. The result is that the process condensate return from the LT cooling to the gray water system is reduced substantially. In the case considered the process condensate flow decreases from about 7,938 to 4,082 kg-mol/hr (17,500 to 9,000 lb-mol/hr). Note that in this calculation it is assumed that much less process condensate is routed to the ammonia stripper, since much of the ammonia in the scrubber overhead gas will be converted on the shift catalyst. In cases where steam addition is required, the return flow may be somewhat higher, but this is not

expected to be beyond the capability of the existing process condensate pumps.
The overall water balance will in any case need to be reviewed as part of the FEED exercise. As a first order estimation, one can assume that the steam consumption of the shift reaction (about 3,720 or 68,040 kg-mol/hr [8,200 or 150,000 lb/hr]) additional fresh water will have to be supplied to the system.

- Although the shift catalyst is capable of a degree of COS conversion, the resulting conversion rate is less than with COS hydrolysis, which operates at a lower temperature. In the base case we can see that of 8.8 kg-mol/hr (19.4 lb-mol/hr) entering the system, the slip from the hydrolysis unit is only 0.6 kg-mol/hr (1.33 lb-mol/hr); whereas the preliminary calculations for the CO-shift show a slip of 1.0 kg-mol/hr (2.28 lb-mol/hr), about 70% more. This issue will need to be addressed during the FEED, in particular:
  - The final determination of COS slip must be reviewed with the catalyst vendor.
  - The effect of the increased COS slip on the Selexol performance will need to be reviewed, especially whether it is compatible with the CO_2 specification.
  - If the COS slip from the CO shift is unacceptable, then the COS hydrolysis bed may have to be retained, but due consideration will have to be given to the additional pressure drop.
- The data available does not allow a detailed review of the low sulfur case, which would be necessary for a complete assessment of CC case 3. The sour shift catalyst only operates in the sulfided state and this requires a minimum amount of sulfur in the feed gas to ensure that it remains sulfided. The minimum sulfur content is of the order of magnitude of 550-850 ppmv, depending on operating temperature. An exact limit would need to be confirmed by the catalyst vendor. The sulfur content of the base case feed gas is about 1% (wet basis) (10,000 ppmv). The sulfur contained in the acid gas for the low sulfur case is about 12% of that of the high sulfur case. It is therefore assumed that the sulfur content of the scrubber overhead gas will be about 1,200 ppmv, which will be sufficient for the catalyst. This issue needs review during the FEED process.

  Should the sulfur content prove to be insufficient, then there is a simple solution at hand, namely to install a bypass around the Claus unit and recycle part of the acid gas. Since this case is based on a low acid gas flow, it is not expected that the capacity of the tail gas recycle compressor would present any limitation, but his would need to be checked during the FEED.
- Consideration needs to be given to the initial start up procedure for a fresh loading of shift catalyst. This can be achieved using the gasifier gas. Typically the gas would need to be cooled to condense out some of the water before routing it to the catalyst. This can be achieved with the existing equipment, provided suitable bypasses are installed. Typically, however, this procedure is associated with additional sulfur emissions. The whole procedure needs to be reviewed during the FEED, to determine how these emissions can be minimized (e.g. by using the Selexol LP Absorber) and he result incorporated into the permit for the retrofitted plant.

6.3.4 Acid Gas Removal

For CC case 3, the SOR and EOR synthesis gas compositions for a single train discharging to two stages of CO-shift and cooling are presented in the tables below. The following significant macro-changes have occurred from the base case:

- The flowrate of synthesis gas per train has increased from 10,843 kg-mol/hr (23,905 lb-mol/hr) (dry basis) for the performance coal (Case 1) to 14,630 kg-mol/hr (32,253 lb-mol/hr) (dry) for SOR, an increase of 35%.
- The CO_2/H_2S ratio has increased from ~ 9/1 for the performance coal (Case 1) to ~ 23/1 for SOR

Table 6-5
Start-of-Run CO-Shift Performance Characteristics

Unit:	**CO Shift**				
	Stream	KO Drum Outlet – **Start-of-Run (SOR)**			
	Phase		Vapor		
Component	**[kg/kmol]**	**lb-mol/hr**	**kg-mol/hr**	**mole% (dry)**	**mole% (wet)**
CO_2	44.010	13803.44	6260.07	**42.665**	**42.609**
CO	28.010	639.94	290.22	**1.978**	**1.975**
H_2	2.016	16365.82	7422.14	50.586	50.518
CH_4	16.043	6.59	2.99	0.020	0.020
CH_3OH	32.041	0.00	0.00	0.000	0.000
N_2	28.013	595.00	269.84	1.839	1.837
O_2	32.000	0.00	0.00	0.000	0.000
Ar	39.948	254.75	115.53	0.787	0.786
H_2S	34.082	604.98	274.37	**1.870**	**1.867**
COS	60.076	1.78	0.81	**0.006**	**0.006**
SO_2	64.066	0.07	0.03	0.000	0.000
NH_3	17.031	80.38	36.45	0.248	0.248
HCN	27.026	0.00	0.00	0.000	0.000
H_2O	18.016	42.97	19.49	0.133	0.133
Carbon	[kg/h]				
Ash	[kg/h]				
Total (dry):		**32352.758**	14672.45	100.00	
Total (wet)		32395.729	14691.94	100.13	100.00
H_2O	[kg/h]		32		
Pressure	[psia/bar a]	516.50	35.62		

Table 6-5 (continued)
Start-of-Run CO-Shift Performance Characteristics

Component	**[kg/kmol]**	**lb-mol/hr**	**kg-mol/hr**	**mole% (dry)**	**mole% (wet)**
Temperature	[°F/°C]	90.61	33		
Molar Flow	[kmol/h]		14692		
Mass Flow	[kg/h]		321194		
Norm. vol flow	[Nm³/h dry]		328868		
Molar Mass	[kg/kmol]		21.86		
Vapor fract.	[%]		1.00		

Table 6-6
End-of-Run CO-Shift Performance Characteristics

Unit:	**CO Shift**				
	Stream	KO Drum Outlet -- End-of-Run (EOR)			
	Phase		Vapor		
Component	**[kg/kmol]**	**lb-mol/hr**	**kg-mol/hr**	**Mol% (dry)**	**Mol% (wet)**
CO_2	44.010	13520.82	6131.89	**42.168**	**42.112**
CO	28.010	920.89	417.64	**2.872**	**2.868**
H_2	2.016	16084.79	7294.69	50.164	50.098
CH_4	16.043	6.59	2.99	0.021	0.021
CH_3OH	32.041	0.00	0.00	0.000	0.000
N_2	28.013	595.00	269.84	1.856	1.853
O_2	32.000	0.00	0.00	0.000	0.000
Ar	39.948	254.75	115.53	0.794	0.793
H_2S	34.082	599.53	271.89	**1.870**	**1.867**
COS	60.076	2.36	1.07	**0.007**	**0.007**
SO_2	64.066	0.07	0.03	0.000	0.000
NH_3	17.031	79.35	35.99	0.247	0.247
HCN	27.026	0.00	0.00	0.000	0.000
H_2O	18.016	42.59	19.32	0.133	0.133
Carbon	[kg/h]				
Ash	[kg/h]				

Table 6-6 (continued)
End-of-Run CO-Shift Performance Characteristics

Component	[kg/kmol]	lb-mol/hr	kg-mol/hr	Mol% (dry)	Mol% (wet)
Total (dry):		**32064.145**	14541.56	100.00	
Total (wet)		**32106.736**	14560.88	100.13	100.00
H_2O	[kg/h]		32		
Pressure	[psia/bar a]	516.50	35.62		
Temperature	[°F/°C]	90.61	33		
Molar Flow	[kmol/h]		14561		
Mass Flow	[kg/h]		318785		
Norm. vol flow	[Nm³/h dry]		325935		
Molar Mass	[kg/kmol]		21.89		
Vapor fract.	[%]		1.00		

The changes in synthesis gas feed composition and conditions are significant, and will require modifications to the Selexol H_2S section. These changes will have to be examined by UOP to define them quantitatively, but they involve the following areas:

- Assuming there is little margin in the two existing Selexol H_2S absorbers, a third H_2S absorber will have to be installed to accommodate the 35% increase in gas flow. The three H_2S absorbers will have greater turndown restrictions, since they will no longer be operating at full load at 100% facility capacity. This will also bring about challenging control issues
  - The two syngas streams exiting the upstream CO-shift and cooling sections will have to be divided into three streams feeding the three H_2S absorbers
  - The three syngas streams exiting the H_2S absorbers will have to be "combined" into two equal syngas streams feeding the CO_2 absorbers
  - When a gasifier is taken offline, the syngas from one line of CO-shift will have to be split into two streams feeding two H_2S absorbers, and then recombined to feed a single CO_2 absorber

An alternate way to address this issue is to add a new Selexol H_2S absorber to each train, bringing up the total number of Selexol H_2S absorbers to four. This added investment may be compensated for by the reduction in control complexity that was just described. This issue of how best to add Selexol H_2S absorbers will have to be reviewed in detail during the FEED stage.

- The more-than-doubling of the CO_2:H_2S ratio in the syngas feed will tax the Selexol H_2S section and its ability to maintain a H_2S concentration in the acid gas at greater than 50% by volume. This may require the addition of a H_2S concentrator tower (see Section 3.3.4), which would replace the rich solvent flash drum in the base case configuration. A slipstream of either N_2 or product syngas from the CO_2 absorber would be used as the inert stripping agent to back-strip CO_2 from the rich solvent.

In the base case configuration, however, the flash gas stream generated is a particular large flow, as mandated by GEE to service their upstream gasifier requirements. This large flow enables the existing rich solvent flash drum to act more like a H_2S concentrator, which may allow its continued use instead of adding a H_2S concentrator tower. This is viewed as a limited probability, but only support by UOP can deny or confirm this. Even if that analysis shows the rich solvent flash drum can continue to perform, it is almost certain that there will be deterioration in acid gas quality. This can lead to reductions in capacity of the H_2S stripping section of the Selexol process, and can have impact upon the performance and capacity of the sulfur removal unit.

- The static mixer and pre-saturation drum in the Selexol H_2S section can be eliminated from service, as the pre-CO_2-loaded lean solvent will already be available from the Selexol CO_2 absorber. The existing loaded solvent chiller can be used to cool this stream prior to entrance into the Selexol H_2S absorber.

Referencing the discussions in Sections 3.3.4 & 4.3.4, a new CO_2 removal section has to be added and integrated with the modifications to the existing H_2S removal section of the Selexol unit.

6.3.5 Claus Unit

The issues which impact the sulfur block in CC case 3 are essentially the same as those described in Section 5.3.5 for Case 2. Only small changes to any numerical values could show up when comparing the two options.

6.3.6 Combined Cycle

The following is a representative list of modifications required for IGCC with some thoughts in italics concerning the specifics of the three capture case modifications:

6.3.6.1 Gas Turbine Modifications

- IGCC MNQC combustion system
  - *Explore the potential need to change fuel nozzles and evaluate configurations including a performance heater to adjust fuel temperatures to meet the necessary Wobbe Index range.*
- Combustion lab verification testing program
  - *Final production combustion tests will be required unless they were performed during the base plant testing program.*

- Off-base syngas fuel control module with syngas stop and control valves
  - *Fuel and diluent mass flow is considerably reduced (73%) even though a large amount of N_2 has been injected into the fuel stream*
  - *Need study to determine necessary modifications*
  - *Determine method of mixing N_2 and fuel*
  - *Provide fuel blending skids*
  - *Check for controllability*
- Syngas fuel piping on base and from skid to base.
  - *Addition or replacement for high H_2 fuel manifolds on base*
  - *Piping from skid to base is likely within design limits*
- Fuel moisturization
  - *May not be needed after conversion but can be used to optimize performance and adjust flows for more air extraction as desired*
- Nitrogen and steam diluent skid
  - *Split diluent N_2 supply into two streams*
  - *New piping and valves for diluent N_2 to fuel skid at approximately 2 bar higher pressure to meet restriction on minimum syngas heating value. This will have an effect on the ASU.*
  - *Check controllability of N_2 to combustors at reduced flow*
  - *Study of N_2 supply adequacy for fuel switches and unplanned outages*
- Fire and hazardous gas protection system
  - *Determine adequacy of syngas protection for high H_2*
  - *Need unburned hydrocarbon, H_2, CO, CO_2, UV detection and ventilation*
- Air extraction and control skid
  - *Study indicates that there is adequate flow to allow partial air extraction (57% of base case). Other studies do not agree but since the equipment is already in place, it will probably be feasible to continue use.*
  - *Need to study base case skid and piping for operation and controllability at 0-50% of base case flow*
- Steam injection for NOx control on backup fuel
  - *Likely acceptable within design limits*

- Proven HGP Materials instead of single crystal, etc
  - *Consider H_2 material qualification program, particularly for coatings.*
- Stage 1 turbine nozzle increased area for low-Btu syngas flow
  - *Reduction in flow should be acceptable within design limits*
  - *Check coatings for H_2 compatibility*
- Redundant control and protection system additions for Mark Vle
  - *A holistic study of all operating and emergency modes*
  - *Determine plan for operation if CO_2 venting not allowed*
    - *Short term*
    - *Long term*
  - *Add site-specific software*
  - *New firing temperature reduction curves to maintain natural gas fuel metal temperatures for high H_2 fuels due to hydrogen producing excess water in the combustion products.*
  - *Additional control for malfunction of CO_2 capture or sequestration equipment*
  - *Means of restricting minimum syngas equivalent heating value (Btu/ft^3) into the fuel nozzles*
- Accessory system and enclosure design for syngas.
  - *Modify for above notes*

6.3.6.2 Steam Cycle Modifications

HRSG

- GT exhaust temperature and steam flows are lower; however, the Edwardsport base case HRSG configuration will need to be modeled with each of the CC Case 1, 2 and 3 data to determine whether any modifications are necessary.

Steam Turbine and Condenser

- Main steam flows are lower for CC case 3. A study was made concerning the use of the significant amount of heat coming from the shift reactors. One option considered doubling the flow of the 0.4 MPa (56 psi) LP injection and this was discarded due to potential restrictions of the existing piping and potential for changing the thrust balance as the injection appears to enter only one side of the LP units. A second option of admission at 0.6 MPa (84 psi) to the LP crossover flow appears to be a better approach.

6.3.6.3 Operational Considerations

A complete study of operating modes should be made for periods where CO_2 cannot be captured or sequestered to determine fuel characteristics, necessary GT hardware modifications and change in emissions. These modes may include:

- Venting of CO_2 - no change
- Continue shift but shutdown CO_2 removal
- Shutdown shift and CO_2 removal but re-start COS hydrolysis
- Re-injection of CO_2 into GT fuel - will require a fuel gas compressor and change in emissions
- Continue shift but shutdown CO_2 removal
- Shutdown shift and CO_2 removal but re-start COS hydrolysis

Export Control Restrictions

Access to and use of EPRI Intellectual Property is granted with the specific understanding and requirement that responsibility for ensuring full compliance with all applicable U.S. and foreign export laws and regulations is being undertaken by you and your company. This includes an obligation to ensure that any individual receiving access hereunder who is not a U.S. citizen or permanent U.S. resident is permitted access under applicable U.S. and foreign export laws and regulations. In the event you are uncertain whether you or your company may lawfully obtain access to this EPRI Intellectual Property, you acknowledge that it is your obligation to consult with your company's legal counsel to determine whether this access is lawful. Although EPRI may make available on a case-by-case basis an informal assessment of the applicable U.S. export classification for specific EPRI Intellectual Property, you and your company acknowledge that this assessment is solely for informational purposes and not for reliance purposes. You and your company acknowledge that it is still the obligation of you and your company to make your own assessment of the applicable U.S. export classification and ensure compliance accordingly. You and your company understand and acknowledge your obligations to make a prompt report to EPRI and the appropriate authorities regarding any access to or use of EPRI Intellectual Property hereunder that may be in violation of applicable U.S. or foreign export laws or regulations.

The Electric Power Research Institute Inc., (EPRI, www.epri.com) conducts research and development relating to the generation, delivery and use of electricity for the benefit of the public. An independent, nonprofit organization, EPRI brings together its scientists and engineers as well as experts from academia and industry to help address challenges in electricity, including reliability, efficiency, health, safety and the environment. EPRI also provides technology, policy and economic analyses to drive long-range research and development planning, and supports research in emerging technologies. EPRI's members represent more than 90 percent of the electricity generated and delivered in the United States, and international participation extends to 40 countries. EPRI's principal offices and laboratories are located in Palo Alto, Calif.; Charlotte, N.C.; Knoxville, Tenn.; and Lenox, Mass.

Together…Shaping the Future of Electricity

1021656

Electric Power Research Institute
3420 Hillview Avenue, Palo Alto, California 94304-1338 • PO Box 10412, Palo Alto, California 94303-0813 • USA
800.313.3774 • 650.855.2121 • askepri@epri.com • www.epri.com