Carnegie Mellon University

A Technical and Economic Assessment of CO₂ Capture Technology for IGCC Power Plants

A Dissertation Submitted in Partial Fulfillment of the Requirements for the Degree of

Doctor of Philosophy in Engineering and Public Policy

by

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Pittsburgh, Pennsylvania December 2005

ABSTRACT

As an emerging technology for electric power generation, Integrated Gasification Combined Cycle (IGCC) power plants are of increasing interest because of their potential advantage for CO_2 capture in addition to conventional pollution control. To further explore this technology, this thesis develops a general modeling framework to provide tools for assessing gasification-based energy conversion systems with various CO_2 capture options on a systematic and consistent basis.

Many factors influence the performance and cost of an IGCC power plant. Simulation studies of an oxygen-blown Texaco quench gasifier system with a water gas shift (WGS) reactor and Selexol CO₂ capture unit indicated that the CO₂ avoidance cost is lowest when the CO₂ removal efficiency is in the range of 85%-90%. The overall cost of IGCC systems with and without CO₂ and storage varied significantly with coal quality and plant size (among other factors). For low rank coals (sub-bituminous and lignite) costs increased significantly relative to the nominal case with bituminous coal. It was also found that larger IGCC plants have slightly higher thermal efficiency and lower capital cost. Without incentive financing, however, an IGCC power plant without CO₂ capture was found to be less competitive (more costly) than PC and NGCC power plants in terms of both the total capital requirement and cost of electricity production. However, IGCC plants with CO₂ capture were competitive with PC and NGCC capture plants without incentive financing.

This thesis also provides an overview of available options and decisions factors for using IGCC technology to repower aging PC power plants. Studies in this thesis show

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that IGCC repowering is less capital intensive than greenfield plants, but the feasibility of repowering is very site-specific. Under suitable conditions, IGCC repowering may be an economically attractive option for existing PC plants.

This thesis also attempts to characterize key uncertainties affecting the performance and cost of IGCC systems with CO_2 capture through data mining and Monte Carlo simulation. Most of the capital cost uncertainty in an IGCC capture plant comes from the IGCC process, rather than the CO_2 capture process. Considering the historical variability of capacity factor and coal price for large U.S. coal plants, the COE of an IGCC capture plant may be higher than the expected value based on typical deterministic assumptions.

This thesis also presents preliminary evaluations of IGCC systems using two advanced technologies, the Ion Transport Membrane (ITM) system for oxygen production and the GE H-class gas turbine system for power generation. Study results show that these two technologies can significantly improve the competitiveness of IGCC systems and will influence the application of IGCC technologies in the near future.

ACKNOWLEDGEMENTS

This thesis was written with the generous contributions of time, advice, patient guidance and support of many people.

First, I would like to express my gratitude to my thesis advisor, Prof. Edward S. Rubin, for his unwavering support. He met with me for countless hours to lead me carefully towards the creation of a final piece of work. His expertise, vast knowledge, understanding, patience and kindness added considerably to my graduate experience, and truly made a difference in my life. I doubt that I would ever be able to convey my appreciation fully, but I owe him my eternal gratitude.

I would like to thank the members of my committee, Professors Jay Apt, Howard J. Herzog, and Allen L. Robinson, for their constant encouragement and feedback at all levels of this research project. Their inputs were very helpful in fulfilling the objectives of this research.

I also want to acknowledge the extensive contributions of the entire IECM team. My special thanks go to Mike Berkenpas for his never-ending assistance in computer modeling, and technical writing, and to Connie Zaremsky for her tremendous input and help in the manuscript editing during the stressful writing-up period. I would also like to thank Anand Rao and Sean McCoy, two PhD students in our team, for our exchanges of knowledge, skills, and venting of frustration during my graduate program, which enriched my experience. I would like to thank the entire Department of Engineering and Public Policy which provided a friendly and warm environment. I am especially thankful to Gloria Blake, Elizabeth Ganley and Victoria Finney for their help. Thank you all, without you all, life would have been very difficult.

I would also like to thank my whole family for the support they provided me through my entire life and in particular, I must acknowledge my wife, Li Li, without whose love, encouragement and support, I would not have finished this thesis.

This research was supported by the Carnegie Mellon Electricity Industry Center (CEIC) and the U.S. Department of Energy National Energy Technology Laboratory (DOE/NETL).

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NOMENCLATURE

α	CO ₂ removed from the syngas (percent)
ω	Selexol flow rate (lb-mole/hr)
ψ_{CO_2}	Solution heat of CO ₂ in selexol (Btu/lb solute)
η	Percentage of the theoretical recovery
$\eta_{\scriptscriptstyle HS}$	H ₂ S removal efficiency (%)
$\eta_{\scriptscriptstyle tur}$	Power recovery turbine efficiency (%)
$\eta_{\scriptscriptstyle comp}$	Compressor efficiency (%)
$\eta_{_{pump}}$	Pump efficiency (%)
γ	Ratio of actual to required flow rate of selexol (fraction)
${\mathcal X}_i$	Solubility of gas component i in selexol (SCF/gallon-psia)
ΔT	Temperature increase of solvent in the absorber (F)
ΔT_1	Selexol temperature change due to heat transfer from syngas (F)
ΔT_2	Selexol temperature change due to solution heat (F)
ΔT_{Sel}	Selexol temperature difference between the inlet and outlet of the refrigeration unit (C)
ξ	CO conversion
ξ _h	CO conversion in the high temperature shift reactor
ξı	CO conversion in the low temperature shift reactor
ξ _{tot}	Total CO conversion in the shift reactors
В	The amount of debt
C _{cat.}	Capital cost of catalysts (k\$ in 2000)
$C_{p,S}$	Specific heat (constant pressure) of selexol (Btu/lb-F)

$C_{p,i}$	Specific heat (constant pressure) of gas component i (Btu/lb-F)
$C_{v,i}$	Specific heat (constant volume) of gas component i (Btu/lb-F)
dP_{pump}	Pressure increase in the pump (psia)
dP_{tur}	Decreased pressure of Selexol in the power recovery turbine (F)
dT_{HE}	Log mean temperature difference in the heat exchanger (C)
dT_{tur}	Decreased temperature of Selexol in the power recovery turbine (F)
dT_{pump}	Increased temperature of CO ₂ lean Selexol due to pumping (F)
dT_h	Approach temperature in the high temperature reactor (F)
dT ₁	Approach temperature in the low temperature reactor (F)
$(dT)_{HE1}$	Log temperature difference in heat exchanger 1 (F)
$(dT)_{HE2}$	Log temperature difference in heat exchanger 2 (F)
$(dT)_{HE3}$	Log temperature difference in heat exchanger 3 (F)
F	Molar flow rate of air fed into the ITM unit (mole/hr)
$F_{\it perm}$	Molar flow rate of permeated oxygen (mole/hr)
\mathbf{f}_0	Total molar flow rate of syngas at the inlet of high temperature reactor (lb-mole/hr)
$\boldsymbol{f}_{\text{fuel}}$	Total molar flow rate of syngas (fuel) from Selexol process (lb-mole/hr)
f_{gas}	Flow rate of gas captured in Selexol (lb-mole/hr)
$f_{\it Sel}$	Flow rate of Selexol (gal/min or lb-mole/hr)
$f_{{\rm SG},i}$	Total flow rate of syngas entering the absorber (lb-mole/hr)
$\boldsymbol{f}_{SG,HE2,o}$	Total molar flow rate of syngas from the exit of heat exchanger 2 (lb-mole/hr)
\mathbf{f}_{water}	Water molar flow rate entering heat exchanger 1 (lb-mole/hr)

H_s	Total dynamic pressure head of a turbine or pump (psia)
hp_{comp}	Power consumption of compressor (hp)
hp pump	Power consumption of pump (hp)
hp _{tur}	Power recovered through the power turbine (hp)
hp _{RC}	Power consumption of the recycle compressor (hp)
hp _{sp}	Power consumption of the Selexol pump (hp)
$h_{T_{satt.}}$	Enthalpy of water at the saturation temperature (Btu/lb-mole)
h_{T_0}	Enthalpy of water at the inlet of heat exchanger 1 (Btu/lb-mole)
Κ	Reaction equilibrium constant
k _{gas}	Specific heat capacity ratio of gas (C_p/C_v)
K _h	Shift reaction equilibrium constant in the high temperature reactor with taking accounting of the approach temperature
$\mathbf{K}_{k,real}$	Shift reaction equilibrium constant in the high temperature reactor without taking accounting of the approach temperature
K ₁	Shift reaction equilibrium constant in the low temperature reactor with taking accounting of the approach temperature
K _{l,real}	Shift reaction equilibrium constant in the low temperature reactor without taking of the approach temperature
k	Reaction rate constant
$M_{syn,S,i}$	Total molar flow rate of syngas through the sulfur removal system (lb-moles/hr)
MW_{S}	Selexol molecular weight (lb/lb-mole)
$N_{O,slump}$	Operating train number of the sump tanks
$N_{O, reft}$	Operating train number of the refrigeration unit

$N_{O \tan k}$	Operating	train number	of flash tank
U. tan t	1 0		

- $N_{O,HE}$ Operating train number of the heat exchanger
- N_{0.8} Number of operating sulfur removal absorber vessels
- N_{0.R} Operating train number of reactor
- N_{O,HE1} Operating train number of heat exchanger 1
- N_{0.HE2} Operating train number of heat exchanger 2
- N_{O.HE3} Operating train number of heat exchanger 3
- $N_{T,reft}$ Total train number of the refrigeration unit
- $N_{T,\tan k}$ Total train number of flash tank
- N_{T sump} Total train number of sump tanks
- $N_{T,HE}$ Total train number of the heat exchanger
- N_{TS} Total number of sulfur removal absorber vessels (operating plus spares)
- $N_{T.abso}$ Train number of absorbers
- N_{T.R} Total train number of reactor
- N_{T,HE1} Total train number of heat exchanger 1
- $N_{T,HE2}$ Total train number of heat exchanger 2
- N_{T.HE3} Total train number of heat exchanger 3
- P₀ Syngas pressure at the inlet of high temperature reactor (psia)
- $P_{abso.,i}$ Inlet pressure of absorber (atm)
- P_{perm} Pressure of permeated oxygen (psia)
- $P_{o,1}$ Total pressure at the outlet of recovery turbine #1 (psia)

$P_{o,2}$	Total pressure at the outlet of recovery turbine #2 (psia)
$P_{comp,i}$	Inlet pressure of compressor (psia)
$P_{comp,o}$	Outlet pressure of compressor (psia)
P_i	Partial pressure of gas component i (psia)
$P_{tur,o}$	Outlet pressure of the turbines (atm)
$P_{\scriptscriptstyle HE}$	Pressure of syngas entering the heat exchanger (atm)
$P_{_{HE2}}$	Pressure at heat exchanger 2 (atm)
P_{R}	Pressure in the shift reactor (atm)
P_{sc}	Steam cycle pressure (psia)
PFC _{avso}	Process facility cost of the absorber (US k\$ in 2000)
PFC_{tur}	Process facility cost of power recovery turbine (US k\$ in 2000)
PFC _{sump}	Process facility cost of sump tank (US k\$ in 2000)
PFC_{RC}	Process facility cost of the recycle compressor (US k\$ in 2000)
PFC _{SP}	Process facility cost of the Selexol pump (US k\$ in 2000)
PFC_{comp1}	Process facility cost of the CO_2 compressor (US k in 2000)
PFC _{comp2}	Process facility cost of the compressor (US k\$ in 2000)
PFC_{refr}	Process facility cost of the refrigeration unit (US k\$ in 2000)
$PFC_{\tan k}$	Process facility cost of flash tanks (US k\$ in 2000)
PFC_{HE}	Process facility cost of the heat exchanger (US k\$ in 2000)
PFC_{R}	Process facility cost of shift reactor (k\$ in 2000)
PFC_{HE1}	Process facility cost of heat exchanger 1

PFC_{HE2}	Process facility cost of heat exchanger 2
PFC _{HE3}	Process facility cost of heat exchanger 3
Q_1	Heat released from the syngas to the solvent (Btu/hr)
$Q_{\scriptscriptstyle HE}$	Heat load of the exchangers (kW), 1200~96000 /train
$\mathbf{Q}_{\mathrm{HE1}}$	Heat exchanged in heat exchanger 1 (Btu/hr or kW)
Q_{HE2}	Heat exchanged in heat exchanger 2 (Btu/hr or kW)
$\mathbf{Q}_{\mathrm{HE3}}$	Heat exchanged in heat exchanger 3 (Btu/hr or kW)
$\boldsymbol{q}_{\mathrm{HE1}}$	Heat released in heat exchanger 1 by per lb-mole syngas (Btu/lb-mole)
$\boldsymbol{q}_{\mathrm{HE2}}$	Heat released in heat exchanger 2 by per lb-mole syngas (Btu/lb-mole)
$\boldsymbol{q}_{\mathrm{HE3}}$	Heat absorbed in heat exchanger 3 by per lb-mole feed water (Btu/lb-mole)
R	Overall recovery of the ITM unit
R_T	Theoretical overall recovery of the ITM unit
r _B	Cost of debt (borrowing rate)
r _s	Cost of equity
<i>r_{wACC}</i>	Average cost of capital after tax for the project
S	Amount of equity
SV	Space velocity in a reactor (1/hr)
T_0	Temperature of syngas at the inlet of high temperature reactor (F)
T_{C}	Tax rate
T _{evap}	Evaporation temperature of refrigerant (F)
$T_{SG,i}$	Syngas temperature at the inlet of the absorber (F)
$T_{SG,o}$	Syngas temperature at the outlet of the absorber (F)

$T_{\rm HE2,o}$	Temperature of syngas exiting the heat exchanger 2 (F)
T_h	Equilibrium temperature in the high temperature reactor (F)
T_l	Equilibrium temperature in the low temperature reactor (F)
T _{sat.}	Temperature of saturation water (F)
$T_{\text{SG},\text{HE3},i}$	Temperature of syngas at the inlet of heat exchanger 3 (F)
$T_{wat.,HE3,o}$	Temperature of water at the exit of heat exchanger 3 (F)
V _{cat.}	Catalyst volume (ft ³)
V _{cat.,h}	High temperature catalyst volume (ft ³)
V _{cat.,l}	Low temperature catalyst volume (ft ³)
V _{CO2} ,res	Volume flow rate of residual CO ₂ in the lean solvent (lb-mole/hr)
$V_{CO_2,abs}$	Volume flow rate of CO ₂ captured in the absorber (lb-mole/hr)
V_{i}	Volume flow rate of gas component i captured by Selexol (SCF/hr)
v_i	Specific volume of CO ₂ (SCF/lb-mole)
VF	Volumetric flow rate of syngas (ft ³ /hr)
VF _{gas}	Volume flow rate of gas (ft ³ /min)
$W_{e,S}$	Sulfur removal equipment power consumption (kW)
W _{ref.}	Power consumption of solvent refrigeration process (kW)
x_{feed}	Molar concentration of oxygen in the air
[y]	Molar concentration of species \mathcal{Y} in syngas
[y] ₀	Molar concentration of species \mathcal{Y} in syngas entering the high temperature reactor

- $[y]_{h,o}$ Molar concentration of species y in syngas exiting the high temperature reactor
- $[y]_{l,o}$ Molar concentration of species y in syngas exiting the low temperature reactor

Chapter 1. INTRODUCTION

1.1. Climate change and CO₂ emissions

Global climate change, a widely discussed topic in environmental studies, represents a potentially serious threat to natural ecosystems, and to the quality of human life on earth. Studies predict that the adverse consequences of climate change induced by human activities will include some of the following in the future [World Energy Assessment, 1999]:

- The average temperature of the global surface air will increase by 1.0~3.5°C during this century.
- The global mean sea level is likely to rise by about 6cm per decade during this century, mainly due to the thermal expansion of the ocean and the melting of some land ice.
- Even though food production may increase in some areas, the high likelihood of its decrease in other areas, especially in the tropics and subtropics, will bring hardship to large segments of population.
- Fast climatic changes may result in the instability of ecosystems, causing natural disasters such as floods and droughts.
- Some diseases currently contained within certain areas may spread further to threaten new populations.

To minimize the impacts of climate change, it is important to pinpoint and eliminate factors responsible for this phenomenon. It is well-known that a number of gases, such as

carbon dioxide, ozone, methane, nitrous oxide and CFCs in the atmosphere, induce the greenhouse effect that drives global climate change. The contribution of carbon dioxide is, however, dominant because of two reasons. First, the concentration of this gas is already higher than other greenhouse gases in the atmosphere, and, second, human activity on earth today is adding carbon dioxide to the atmosphere at historically unprecedented rates.

The recent increase in the atmospheric concentration of carbon dioxide has resulted from the large scale utilization of fossil fuels in modern times. Since the onset of the industrial revolution, for instance, 296 gigatonnes of carbon from fossil fuels have been released to the atmosphere, raising carbon dioxide concentration from 280ppm to 360ppm [World Energy Assessment, 1999].

At present, fossil fuels fulfill about 84.8% of the world's primary energy consumption needs [International Energy Annual 1999], and in the foreseeable future, fossil fuels will still be the major energy source. Estimates indicate that the world's fossil fuel reserves contain approximately 6600 gigatonnes of carbon, with 5200 gigatonnes of carbon in coal alone. In the absence of adequate measures to control the emissions of carbon dioxide from these sources, the atmospheric concentration of carbon dioxide could more than double by the end of this century [World Energy Assessment, 1999]. Concerns about the greenhouse effect call for new strategies regarding the use of coal to reduce the emission of carbon dioxide into the atmosphere. The task of reducing carbon dioxide emissions without abruptly cutting off the utilization of fossil fuels, however, presents a serious challenge. The following section discusses some methods for facing this challenge.

1.2. IGCC—a promising technology for CO₂ emission control

As the most carbon-intensive and most abundant fossil fuel, coal is traditionally utilized through combustion. Coal combustion plants produce flue gas streams consisting mostly of nitrogen (from combustion air), with diluted concentrations of CO₂. Technologies have been developed to capture CO₂ with low partial pressure from flue gas stream, but the energy and economic performance of coal combustion power plants would be degraded substantially. Post-combustion CO₂ capture from coal combustion plants would nearly double the cost of electricity (COE), and reduce their net output by about 25-30% [Parson, 1998; Doctor, 1994; McCarthy, 1985]. Although oxyfuel can make CO₂ capture from coal combustion power plant much cheaper, the expensive oxygen production required adds significantly to the overall plant cost. For these reasons, there is growing interest in Integrated Gasification Combustion Cycle (IGCC) systems as an alternative.

Gasification offers a way of converting coal to a gaseous state where it can be cleaned and burnt in a gas turbine. CO_2 emissions can be prevented in a gasification power plant by transferring almost all carbon compounds to CO_2 through the water gas shift reaction, and then removing the CO_2 before it is diluted in the combustion stage. Hence, CO_2 removal from IGCC requires considerably smaller and simpler process equipment than the post-combustion CO_2 removal [Herzog, 1999; Herzog, 1997]. Therefore, compared to coal combustion plants, IGCC power plants provide an option for CO_2 capture with relatively low cost and small energy losses.

In addition, IGCC systems are of interests to governments and utility companies in many countries for other reasons. For example, the United Stated government may be interested in developing reliable and cost effective IGCC systems for generating power using its abundant coal reserves in order to reduce US dependence on foreign energy sources.

1.3. Research motivation and objectives

As a new emerging coal-based technology, IGCC systems are becoming an increasingly attractive option to limit CO₂ emissions and other pollutants relative to conventional coal power plants.

A number of previous studies have reported cost and performance results for IGCC systems with CO₂ capture [Michael, 1997; OLeefe, 2000; Doctor, 1997]. However, there are no generally available process models that can be used or modified for studying options of CO₂ removal from IGCC systems in detail. Currently reported cost data also are relatively limited and often incomplete, and uncertainties are seldom considered.

This research, therefore, is motivated by a desire to have a better understanding of the technological options for CO_2 capture from IGCC systems and their effects on the performance and cost of IGCC systems. Some key research questions which need to be addressed include: What kind of technologies may be used for CO_2 capture? What are the key parameters that affect the performance and the cost of IGCC systems with and without CO_2 capture? What are the uncertainties associated with IGCC system? How will CO_2 capture influence the future development and application of IGCC systems? How will IGCC systems and CO_2 capture benefit from developing technologies?

With these objectives in mind, this thesis develops a general framework to assess the range of options for CO_2 capture from IGCC power systems. In this general framework, energy and economic models are developed to simulate the performance and costs of IGCC systems with and without CO_2 capture under different scenarios. Both new and retrofit (repowering) applications of IGCC systems with and without CO_2 capture are studied. The thesis also characterizes key uncertainties affecting performance and costs. It also assesses process design improvements and technology development trends that offer the potential to reduce the cost of IGCC power generation with CO_2 capture. Through evaluating and comparing various IGCC power plant configurations in terms of the cost, performance, and uncertainty, this thesis provides a method for systemic comparison of IGCC options with and without CO_2 capture.

REFERENCES (CHAPTER 1)

- 1. Doctor R.D., Molburg J.C., Thimmapuram P.R., 1997: Oxygen-Blown Gasification Combined Cycle: Carbon Dioxide Recovery, Transport, and Disposal. Energy Convers. Mgmt. Vol. 38, Suppl.
- 2. Doctor R.D., etc., 1994: Gasification combined cycle: carbon dioxide recovery, transport, and disposal, ANL/ESD-24
- Herzog H, 1999: An Introduction to CO₂ Separation and Capture Technologies. Cambridge, U.S.A: MIT Energy Laboratory Working Paper; 1999
- Herzog H and Drake E., 1997: CO₂ capture, reuse, and storage technology for mitigating global climate change: A White Paper. DOE order #DE-AF22-96PC01257, 1997
- 5. International Energy Annual 1999, Energy information Administration, U.S.
- 6. Keith DW and Morgan MG, 2000: Industrial carbon management: a review of the technology and its implications for climate policy, Elements of Change 2000, Aspen Global Change Institute
- McCarthy, C.B., and Clark W.N., 1985: Integrated gasification/combined cycle (IGCC) electric power production-A rapidly emerging energy alternative. Presented at symposium on coal gasification and synthetic fuels for power generation, San Francisco, CA
- 8. OLeefe. L.F., Griffiths J., 2000: A single IGCC design for variable CO₂ capture, 2000 Gasification technologies conference, San Francisco, California, Oct. 2000
- 9. Parson EA, Keith DW, 1998: Climate change Fossil fuels without CO₂ emissions, SCIENCE, 282 (5391) NOV 6, 1998
- 10. World Energy Assessment, 1999: Energy and the challenge of sustainability, http://www.undp.org/seed/eap/activities/wea/drafts-header.html

Chapter 2. THE ROLE OF IGCC POWER PLANTS FOR ABATING CO₂ EMISSIONS

This section discusses the potential role of IGCC power plants for abating CO_2 emissions. As the first step, it presents a brief introduction to gasification and IGCC technology, starting with a generalized overview of IGCC systems. This is followed with detailed descriptions of major IGCC components, including the air separation unit, coal preparation facility, gasifier, syngas cooling unit, basic syngas cleanup options, and combined cycle power block. Then the expected advantages of IGCC for abating CO_2 emissions are discussed, along with technologies that can be incorporated into IGCC systems for CO_2 capture, and the process designs for IGCC systems with CO_2 capture.

2.1. Overview of IGCC system

IGCC is an innovative electric power generation system that combines modern coal gasification technologies with both gas turbine (Brayton cycle) and steam turbine (Rankine cycle) technologies, and offers an exceptionally clean, flexible and cost-efficient way to generate electricity. The gasification system converts coal or other solid or liquid feed stocks such as petroleum coke or heavy oils into a gaseous syngas, which is mainly composed of hydrogen (H₂) and carbon monoxide (CO). The combustible syngas is used to fuel a combined cycle generation power block to produce electricity.

The first commercial IGCC plants were put into service in 1980 in the U.S. through DOE's cooperative Clean Coal Technology (CCT) program. Through the rapid development in recent 20 years, IGCC is now considered as a mature technology and a viable coal power plant option. So far, four IGCC power plants have been commercially running, and other IGCC projects are being planned. All these IGCC projects have

achieved, or are expected to achieve the lowest levels of criteria pollutant air emissions (NOx, SOx, CO, PM10) among any coal-fueled power plants in the world [Brown, 2003]. Table 2-1 represents the major milestones during the course of IGCC development history.

Time	Events
1887	The first patent for a gasifier was granted to Lurgi GmbH in Germany.
1940	Commercial coal gasification to provide cities with gas for streetlights and domestic consumption became common in Europe and the United States.
1950	The chemical industry began using gasification to make chemicals such as ammonia and fertilizers. However, the feedstock was mostly crude oils rather than coal.
1970	The U.S. Department of Energy funded various studies to evaluate the feasibility of gasifying coal and using syngas as a gas turbine fuel. These studies showed good economics.
1980	Coolwater was commissioned in 1984, which demonstrated the feasibility of IGCC.
1996	Polk Tampa Electric plant was built, which successfully used nitrogen injection for NOx control and demonstrated the commercial feasibility of IGCC technology, Wabash River IGCC repowering plant began operation.
2000	Exxon Singapore plant was built, which employs the widest variety of gas turbine fuels and operability range.
Present	IGCC is now considered a mature technology and a viable coal power plant option.

Table 2 - 1Major milestones of the history of IGCC development (source: GE
webpage)

A general figure of the major processes of an IGCC power plant is given by Figure 2-1. The first part of the IGCC process involves the chemical conversion of coal into syngas, a mixture of mostly hydrogen and carbon monoxide. This conversion is carried out in a gasifier, using very high temperature and only a limited amount of oxygen. When the syngas leaves the gasifier, it must be cleaned of any particulates and other contaminants such as sulfur, so that it can be used as a fuel for gas turbines for power generation. After the syngas is cleaned, it is fed into a gas turbine, which turns an electric generator to produce electric power. In addition, the hot exhaust gas from the gas turbine

flows into a heat recovery steam generator (HRSG) for steam production, which turns a steam turbine that drives another electric generator to generate power.





2.2. Major components of an IGCC system

The major components of IGCC power plants, as shown in Figure 2-1, include the coal handling facility, gasifier, air separation unit, syngas cooling process, syngas cleanup processes, and combined cycle power block. Most of the components of IGCC power plants are associated with processes that have been already widely used in the power, petroleum refining, and chemicals industries. The following sections describe each of these components.

2.2.1. Coal handling equipment

The coal handling facility is employed to unload, convey, prepare, store and feed coal delivered to an IGCC power plant. Generally, the coal handling facility used for an IGCC plant can be divided into five sections: unloading unit, feeding unit, crushing and screening unit, stacking and reclaiming unit, and bunker [Joshi, 2000], which is largely the same as that used at PC power plants.

2.2.2. Gasification technology and gasifier

The gasification process is the heart of an IGCC plant. The process is a partial oxidation process which converts many carbon-based fuels, including most grades of coal, into a synthesis gas (syngas). The fuel is fed into a pressurized vessel, which contains controlled and limited amounts of oxygen or air and steam or water. The chemistry of coal gasification reactions is quite complex. The basic conversion procedures are as in the following. Rising temperature in the gasifier initiates devolatilization and breaking of weaker chemical bonds to yield tars, oils, phenols and hydrocarbon gases. The fixed carbon that remains after devolatilization is gasified through reactions with O_2 , steam, CO_2 , and H_2 . The heat produced by the partial oxidation provides most of the energy required to break chemical bonds in the feedstock, and increases the products to the reaction temperature, and drives endothermic gasification reactions [Rubin, 1989]. These reactions further produce the final syngas. Syngas is a mixture of mainly hydrogen and carbon monoxide, with a small fraction of CO_2 . A small amount of methane may also be present. Methane formation is a highly exothermic reaction and does not consume oxygen and, therefore methane amount is relatively higher in lower-temperature systems, and methane formation increases the

efficiency of gasification and the final heating value of the syngas [O'Brien, 2004]. Overall, about 70% of the feed fuel's heating value is associated with the CO and H_2 components of the gas [O'Brien, 2004], but this value can be higher or lower depending upon the gasifier type and feed stock quality. Most gasification processes being demonstrated use oxygen, instead of air, as the oxidant.

Coal gasification technology has been developed for over one hundred years and by now more than one hundred processes of gasification have been developed [O'Brien, 2004]. According to the coal movement and coal/gas contact pattern in the gasifier, gasification technologies can be classified into three types as the moving bed, fluidized bed and entrained flow bed. Different types of gasifiers have advantages and disadvantages of their own, and IGCC systems can incorporate any one of a number of gasifier designs, but all are based on one of these three types [O'Brien, 2004, Rubin, 1989, Brown, 2003, Cargill, 2001]. The next section will briefly discuss these three types of gasifiers.

Moving-bed gasifier

In moving-bed gasifiers, gas and solid contact in the pattern of counter flow, where large particles of coal move slowly down through the gasifier and react with gases moving up through it. Several different reaction zones are formulated that implement the gasification process. In the drying zone at the top of the gasifier, the entering coal is heated and dried, and the product gas is cooled before it leaves the reactor. The coal is further heated and devolatized by higher temperature gas as it descends through the carbonization zone. In the next zone, the gasification zone, the devolatized coal gasifies by reaction with steam and carbon dioxide. Near the bottom of the gasifier is the combustion zone, which operates at the highest temperature, where oxygen reacts with the remaining char. For the moving bed gasifiers, the discharge gas temperature is principally controlled by the feed coal moisture content. High-moisture lignite coal produces a raw gas temperature of about 600 °F, and low-moisture bituminous coal produces a raw gas temperature of over 1000 °F [Brown, 2003].

The cold gas efficiency (chemical energy in cold gas/chemical energy in fuel) of moving bed gasifiers is higher than that of fluidized bed and entrained flow gasifiers, and the oxidant requirement for moving bed gasifiers is also relatively lower. Because the moving-bed gasifier has higher cold gas efficiency, a larger portion of the original heating value of the coal turns into the chemical energy in the gas, instead of the thermal energy in the gas. Hence, the moving bed gasifier typically does not require the high temperature heat exchangers that are required by entrained-flow and fluidized-bed systems. Thus, more of the total output is generated by the gas turbine and less by the steam turbine in an IGCC system using a moving-bed gasifier. Because of lower gas temperature, the volatile material in coal is difficult to decompose, and there is greater concentration of methane and tar in gas.

The Lurgi dry-ash gasifier, shown in Figure 2-2, is a pressurized, dry ash, movingbed gasifier. It uses steam and O_2 as the oxidants. It uses lump coal rather than pulverized coal and, it produces tars. For the Lurgi gasifier, lump coal enters the top of the gasifier through a lock hopper and moves down through the bed. A rotating coal distributor ensures even distribution of coal around the reactor. Steam and oxygen enter at the bottom and react with the coal as the gases move up the bed. Ash is removed at the bottom of the gasifier by a rotating grate and lock hopper. The coal moves slowly down the gasifier, and it is warmed by the syngas flowing upwards through the bed; thus the coal is sequentially dried and devolatilized, then gasified. The countercurrent operation results in a temperature drop in the reactor. Gas temperatures in the drying and devolatization zone near the top are approximately 260 to 538 °C. The very bottom of the bed is the hottest part of the gasifier (~1000 °C), where almost any remaining coal is oxidized. The CO₂ produced at the bottom reacts with carbon higher in the bed to form CO [DOT, 1998].



Figure 2 - 2 Simplified process of the Lurgi dry-ash gasifier [Brown, 2003]

The Lurgi dry-ash gasifier uses about a 4-5:1 ratio of steam to O_2 as oxidant. The result of this is that the temperature in the dry-ash system is kept sufficiently low at all

points. Thus, the ash does not melt, and can be removed as a dry ash. The low temperature of the dry-ash system means that it is suited more to reactive coals, such as lignite, than to bituminous coals [DOT, 1998].

Fluidized-Bed gasifier

For a fluidized-bed gasifier, coal is typically supplied through one side of the reactor, and oxidant and steam are supplied near the bottom. Fluidized-bed reactors can efficiently mix coal particles in the reactor vessel. Thus, a constant temperature is sustained that is below the ash fusion temperature, which avoids clinker formation and possible de-fluidization of the bed. This means that fluidized bed gasifiers are best suited to relatively reactive fuels, such as biomass. Some char particles are entrained in the raw gas as it leaves the top of the gasifier, but are recovered and recycled back to the reactor via a cyclone. Ash particles which are removed below the bed give up heat to the incoming steam and recycle gas. Fluidized bed gasifiers may differ in ash conditions and in design configurations for improving char use [Worldbank, 2000].

The fluidized bed gasifier has the advantages of simpler reactor structure, uniform and moderate operating temperature, easy operating, accepting a wide range of solid feedstock, free of tar and phenol, and moderate oxygen and steam requirements. In conventional fluidized bed coal gasifiers, like the Winkler gasifier, absence of selected ash discharge design results in low temperature operation and higher carbon content in bottom ash, which causes low carbon conversion, limited coal feedstock resources and relatively small gasification capacity [Worldbank, 2000].
There are relatively few large fluidized bed gasifiers in operation. Commercial versions of this type of gasifier include the high temperature Winkler (HTW) and KRW designs. The latter gasifier was incorporated into the Piñon Pine Coal Gasification Plant [Cargill, 2001].

The KRW gasification process, originally developed by M.W. Kellogg Company, is a pressurized, dry feed, fluidized bed slagging process. The gasifier design is shown in Figure 2-3. The KRW technology is capable of gasifying all types of coals, including high sulfur, high-ash, low rank, and high-swelling coals, and it is also capable of gasifying bio-derived and refuse-derived waste. The only solid waste from the plant is a mixture of ash and calcium sulfate, which is identified as a non-hazardous waste [NETL, 2000]. Coal and limestone, crushed to below 1/4", are transferred from feed storage to the KRW fluidized-bed gasifier via a lock hopper system. Gasification takes place by mixing steam and air (or oxygen) with the coal at a high temperature. The fuel and oxidant enter the bottom of the gasifier through concentric high velocity jets, which assure complete mixing of the fuel with oxidant and char and limestone that collects in the gasifier. Upon entering the gasifier, the coal immediately releases its volatile matters, which are oxidized rapidly to supply the endothermic heat of reaction for gasification. The oxidized volatiles form a series of large bubbles that rise up in the center of the gasifier, which cause the char and sorbent in the bed to move down the sides of the reactor and back into the central jet. The recycling of solids cools the jet and efficiently transfers heat to the bed material. Steam, which enters with the oxidant and through a multiplicity of jets in the conical section of the reactor, reacts with the char in the bed, converting it to syngas.

At the same time, the limestone sorbent, which has been calcined to CaO, reacts with H_2S released from the coal during gasification to from CaS [NETL, 2000].

As the char reacts, the particles become enriched in ash. Repeated recycling of the ash-rich particles through the hot gas of the jet melts the low-melting components of the ash, which causes the ash particles to stick together. These particles are cool when they return to the bed, and this agglomeration permits the efficient conversion of even small particles of coal in the feed. The velocity of gases in the reactor is selected to maintain most of the particles within the bed. The smaller particles that are carried out of the gasifier are recaptured in a high efficiency cyclone and returned to the conical section of the gasifier. Eventually, most of the smaller particles agglomerate when they become richer in ash and gravitate to the bottom of the gasifier. Since the ash and spent sorbent particles are substantially denser than the coal feed, they settle to the bottom of the gasifier, where they are cooled by a counter-current stream of recycled gas [Cargill, 2001].

The char, ash, and spent sorbent from the bottom of the gasifier flow to the fluidbed sulfator, where both char and calcium sulfide are oxidized. The CaS forms CaSO₄, which is chemically inert and can be disposed of in a landfill. Sulfur released from burning residual char in the sulfator is also converted to CaSO₄ [Cargill, 2001].



Figure 2 - 3 Simplified process of the KRW gasifier [Brown, 2003]

Entrained-flow gasifier

In an entrained-flow gasifier, fine coal particles react with steam and oxygen at high temperatures. Entrained-flow gasifiers have the ability to gasify all coals regardless of rank. Depending on designs, entrained-flow systems may use different coal feed systems (dry or water slurry) and heat recovery systems.

In an entrained flow bed, the contact time of gas and solid is very short, which is only about several seconds, but the reaction rate and gasification capacity is greater because of higher gasification temperature (1200-1500 °C) and smaller diameter of pulverized coal (<100um). On the other hand, because of higher operating temperature, part of the coal energy is converted to heat and its cold gasification efficiency is lower. High gas temperature also makes gas cleaning and waste heat recovery system more expensive [Worldbank, 2000]. Entrained-flow gasifiers have the following characteristics:

- Ability to gasify all coals regardless of coal rank, caking characteristics, or amount of coal fines (although feed stocks with lower ash content are favored)
- Uniform temperatures
- Very short fuel residence time in gasifier
- Solid fuel must be very finely divided and homogeneous
- Relatively large oxidant requirements
- Large amount of sensible heat in the raw gas
- High-temperature slagging operation
- Entrainment of some molten slag in the raw gas.

Nearly all commercial IGCC systems in operation or under construction are based on entrained-flow gasifiers. Commercial entrained-flow gasifier systems are available from GE Energy Gasification Technology (formerly ChevronTexaco), ConocoPhillips, Shell, Prenflo, and Noell [Rosenberg, 2004]. The commercial gasification processes believed most suited for near-term IGCC applications using coal or petroleum coke feed stocks are the GE Energy, ConocoPhillips, and Shell entrained-flow gasifiers [SFA Pacific, 2003].

ChevronTexaco gasification technology uses a single-stage, downward-feed, entrained-flow gasifier, shown as Figure 2-4. Fuel/water slurry (e.g., 60-70% coal) and 95% pure oxygen (from an air separation unit) are fed to at the top of a hot, pressurized gasifier. The fuel and oxygen react exothermally to produce raw fuel gas and molten ash at a temperature ranging from 2200 to 2700 °F, and a pressure greater than 20 atmospheres. Operation at the elevated temperatures eliminates the production of hydrocarbon gases and liquids in the syngas. In the syngas cooler design-type, the hot gas flows downward into a radiant syngas cooler where high-pressure steam is produced. The syngas cooler is specifically designed to meet the conditions of high thermal gradients and the ability to handle soot. The syngas passes over the surface of a pool of water at the bottom of the radiant syngas cooler and exits the vessel. The slag drops into the water pool and is fed from the radiant syngas cooler sump to a lock hopper. The black water flowing out with the slag is separated and recycled after processing in a dewatering system. The slag is eventually removed through a lock hopper. This design configuration maximizes heat recovery for steam production, as well as CO production, which is appropriate for an IGCC application. After exiting the gasifier, the syngas is further cooled and cleaned by a water scrubber, and the fine particulate matter and char may be recycled to the gasifier. The cooled, water-scrubbed syngas consists mainly of hydrogen and carbon monoxide, and contains no hydrocarbons heavier than methane. Metals and other ash constituents become part of the glassy slag [Cargill, 2001].



Figure 2 - 4 Texaco gasifier with radiant syngas coolers [Brown, 2003]

An alternate design to the use of a radiant syngas cooler is the use of an exit gas quench. In this design mode, hot gas exiting the reaction chamber is cooled by direct contact with water, and then enters a scrubber for particulate and soot removal. This design provides an effective mechanism to add water to the syngas to promote the water-gas shift reaction and maximize hydrogen production. The quench design mode is often used to accommodate heavy hydrocarbon feedstock [Brown, 2003].

Gasifier type	Moving bed		Fluidized bed		Entrained flow
Commercial Manufacturer	Lurgi	_	KRW		Texaco, Shell
Ash conditions	Dry ash	slagging	Dry ash Agglomerating		Slagging
Fuel size limits	6-50 mm	6-50 mm	<6 mm	< 6 mm	<0.1 mm
Acceptability of caking coal	Yes	Yes	Possibly	No, non-caking	Yes
Preferred feedstock	Lignite, reactive bituminous, anthracite wastes	Bituminous, anthracite, pet coke, waste	Lignite, reactive bituminous, anthracite, waste	Lignite, bituminous, anthracite, cokes, biomass, wastes	Lignite, reactive bituminous, anthracite, pet cokes
Ash content limits	No limitation	< 25% preferred	No limitation	No limitations	<25% preferred
Preferred ash melting temperature, F	>2200	<2370	>2000	>2000	<2372
Exit gas temperature, °F	Low (800- 1200)	Low (800- 1200)	Moderate (1700- 1900)	Moderate (1700-1900)	High (>2300)
Gasification pressure, psia	435+	435+	15	15-435	<725
Oxidant requirement	Low	Low	Moderate	Moderate	High
Steam requirement	High	Low	Moderate	Moderate	Low
Unit capacity, MWh	10-350	10-350	100-700	20-150	Up to 700
Key distinguishing characteristics	Hydrocarbon gas	liquids in raw	Large char recycle		Large amount of sensible of heat energy in the hot raw gas
Key technical issue	Utilization of fines & hydrocarbon liquids		Carbon conversion		Raw gas cooling

Table 2 - 2Important characteristics of three types of gasifiers [Brown, 2003]

Type of gasifier	Ash agglomerating	Fixed bed		Fluidized bed	Entrained flow bed	
Gasification technology	fluidized bed	Oxygen blow (atmosphere)	Lurgi pressurized	HTW	К-Т	Texaco
Coal feeder tyep	Dry, crushed	Lump	Lump	Dry, crushed	Pulverized coal	Slurry
Gasification temperature	~1080 C	800~1000 C	800~1000 C	800~1000 C	~1800 C	~1400 C
Gasification pressure	~30 kPa	~20 kPa	2.24 MPa	1.0~2.5 MPa	34~48 kPa	3.4 MPa
Ash removed	agglomerating	solid	solid	solid	slag	slag
Gasification medium	92% O2+steam	95.2% O2+steam	O2+steam	O2+steam	O2+steam	O2+steam
Oxygen/coal Nm3/kg	0.454	0.64	0.41	0.37	0.7	1.17
Steam/coal kg/kg	0.94	1.37	1.65	0.37	0.27	0.92
Carbon conversion %	~90	>95	>95	~95	99	>95
Cold gas efficiency %	~73	~85	~85	76	76	~76
Carbon content in ash %	7.7			11		9

Table 2 - 3Comparison of parameters of gasification technologies [U.K.
Department of Trade and Industry, 1998]

A general comparison of these three types of gasifiers, and a specific comparison of some commercial gasifiers are given in Table 2-2 and Table 2-3, respectively.

2.2.3. Air Separation Unit (ASU)

This section discusses the function of the air separation units (ASU) in IGCC power plants, followed by descriptions of current commercial technology and developing technology for oxygen production.

Is ASU necessary for IGCC systems?

All coal gasification processes require an oxidant for the gasification reactions. Air, oxygen, or oxygen-enriched air can be used as oxidant for gasification processes. The choice of oxidant affects the amount of nitrogen the gasification system has to handle, and depends on the application, types of gasifiers, and the degree of the system integration. Air-blown gasification eliminates the need for the ASU. Oxygen-blown IGCC systems, however, have several advantages over air-blown IGCC systems.

Syngas from an oxygen-blown gasifier has a heating value ranging from 250 to 400 Btu/scf, compared to an air-blown gasifier with 90 to 170 Btu/scf fuel gas and high nitrogen content [Rubin, 1989]. Syngas with a medium heating value can potentially be used as a replacement for natural gas as gas turbine fuel. In addition, the moderate heating value of the gas helps minimize the size of the gasifier and auxiliary systems. The cold-gas efficiency is 7-10 percentage points higher for oxygen-blown gasification than air-blown gasification due to the avoidance of nitrogen dilution. Gasifier operability and carbon conversion also improves with the use of oxygen [Rubin, 1989]. Comparing to oxygen-blown gasification, air-blown gasification creates additional technical challenges for the gas clean up and combustion turbine operation. Air-blown gasification also is less suited for cost effective separation and capture of CO₂ due to the diluted CO₂ by nitrogen. For these reasons, the next generation of IGCC facilities are expected to be based on entrained-flow, oxygen-blown gasification technologies [Rosenberg, 2004]. To date, all of the gasification processes demonstrated for commercial IGCC plants is oxygen-blown systems.

Cryogenic oxygen production and novel air separation methods

Currently, air separation in large scale is achieved by using a cryogenic process in which air is cooled to a liquid state and then subjected to distillation. The basic elements of an air separation are [Air Products, 2004].

- Filtering and compressing air
- Removing contaminants, including water vapor and carbon dioxide (which would freeze in the process)
- Cooling the air to very low temperature through heat exchange and refrigeration processes
- Distilling the partially-condensed air (at about -300°F / -185°C) to produce oxygen.
- Warming oxygen and waste streams by heat exchange with incoming air

Cryogenic oxygen production, commercialized early in the 20th century, is an established process that is used extensively worldwide. Currently cryogenic processes remain the most economically efficient separation method of making high purity oxygen for high production rate plants. However, an ASU based on the cryogenic process requires a large amount of power and accounts for the largest parasitic load on an

oxygen-blown IGCC plants. In addition, cryogenic processes in general have large capital cost, due mostly to the cost of compressors, turbines, and numerous heat exchangers for high pressure requirements and the recovery of refrigeration energy [Holt, 2001].

For these reasons, lowering the cost of air separation will significantly improve the economics and efficiency of IGCC power plants and lower their capital costs. The Department of Energy is sponsoring a research project to develop a novel air separation technology--the Ion Transport Membrane Oxygen (ITM). ITM is based on ceramic membranes that selectively transport oxygen ions when operated at high temperature. A commercial-scale ITM oxygen module with a capacity of producing 0.5 ton/day oxygen has been run by Air Products. It is predicted that ITM oxygen module capable of producing 1000 ton/day oxygen will be available in 2010. According to Air Products, this technology has the potential to lower the cost of producing oxygen by more than 30% [Air Products, 2004]. Hence, it is expected that the upgrade in air separation technology will significantly improve the economics of IGCC systems [O'Brien, 2004].

2.2.4. Syngas cooling

Coal gasification systems operate at high temperatures and produce raw, hot syngas at temperatures from 800 to 1800 °C. The syngas from the gasifier has to be cooled down for cleanup. Heat recovery is typically utilized to cool down the syngas and increase overall system efficiency. Heat recovered can represent about 15% of the energy in the feed fuel, but this varies with the gasification technology employed (5% for moving bed to 25% for entrained flow processes) [Bruijn, 2003]. Depending on the design of a gasifier, the raw syngas leaving the gasification reactor can be cooled by radiant and/or convective heat exchange and/or by a direct quench system, which injects

either water or cool recycle gas into the hot raw syngas. In most IGCC plant design configurations, saturated steam raised from cooling the raw gasifier syngas is sent to the heat recovery steam generator (HRSG) for superheat and reheat. The steam and water systems are integrated between the gasification island and the power conversion block, but the superheated steam is generally better generated in the HRSG than in the raw syngas coolers [Bruijn, 2003].

2.2.5. Syngas clean-up

The raw syngas produced by the gasification contains various impurities. However, the concentrations of these various components depend on the feedstock composition and the specific gasification process employed. The primary feedstock impurities of concern are the sulfur and ash constituents. In gasification, the sulfur is converted mainly to H₂S and COS, a portion of the ash and unburned carbon is entrained as particulates. Small amounts of HCN and NH₃, and traces of metal carbonyl compounds, are also produced [McCarthy, 1985].

Particulate materials have to be removed from raw syngas before it can be used as a fuel of gas turbine to avoid damaging the turbine. This is generally accomplished by cooling the syngas to much lower temperatures, and then using conventional cleaning methods including cyclones or water scrubbers. The particulate material, including char and fly ash, is then typically recycled back to the gasifier [Bruijn, 2003]. Another option to water scrubbing for particulate removal is the use of ceramic candle filters or sintered metal filters [Korens, 2002].

Next the syngas is treated in "cold-gas" clean up processes, also known as the Acid Gas Recovery (AGR) process, to remove most of the H₂S, carbonyl sulfide (COS) and nitrogen compounds. The primary processes are chemical solvent-based processes or physical solvent-based processes. Sulfur recovery processes recover sulfur either as sulfuric acid or as elemental sulfur. The most common removal system for sulfur recovery is the Claus process, which produces elemental sulfur from the H₂S in the syngas [O'Brien, 2004].

Carbonyl sulfide (COS), which is usually present at a several hundred ppmv level in syngas from coal and petroleum residues, is difficult to remove in AGR units. Therefore, further sulfur removal may be accomplished by the addition of a COS hydrolysis unit (before the AGR), which catalytically converts COS to H₂S. For high sulfur coal, IGCC plants that use COS hydrolysis, together with conventional AGR and sulfur recovery units, have been able to achieve nearly 99% sulfur recovery [Wabash Energy Ltd, 2000].

DOE is currently working on new syngas cleanup systems in which the syngas will need to be cooled only moderately [Simbeck, 2002]. Such a system would have higher process efficiency achievable without syngas cooling and removal of water from the syngas. Potential capital and operating cost savings of these new processes are related to their reduced complexity compared to current cold gas cleanup processes. Hence, once these technologies are commercialized, the economics and environmental friendliness of IGCC power plants is expected to improve.

The focus of most new syngas cleanup programs is the removal of the sulfur, chloride, alkali, and particulates from syngas at temperatures close to the highest inlet

temperature at which gas turbine fuel control and delivery systems could be designed. This level was set at about 1,000 °F by the requirement for very low alkali (potassium and sodium) content of the fuel gas to prevent alkali corrosion of hot gas turbine components and the desire to avoid expensive materials and unreliable refractory-lined piping [Todd, 1994; Holt, 2001]. However, both industry interest and government interest in such processes have declined for several reasons [Stiegel, 2001], including the technical challenge of the process and equipment development, the trend toward more stringent air emissions and the success of the demonstration and commercial O₂-blown gasification projects.

2.2.6. Combined cycle power unit

The clean syngas is sent to the combined cycle power unit. In a combined cycle system, the first generation cycle involves the combustion of syngas in a combustion turbine. The gas turbine powers an electric generator, and the hot exhaust gases from the gas turbine is directed to a heat recovery steam generator to produce steam for a steam turbine to complete the combined power cycle.

For any gas turbine manufacturer, the fuels that will be used will have a profound effect upon both the machine design and the materials of construction. It is most meaningful from the standpoint of turbine application to classify gaseous fuels by their calorific values, which cover a very wide range: from a low of about 100 Btu/ft³ to a high of 5,000 Btu/ft³. Table 2-4 shows such a classification of gaseous fuels.

Classification by calorific value	Calorific value kcal/nm ³ (Btu/scf)	Typical specific fuels	Primary gas components
Very high	10700-44500 (1200-5000)	Liquefied Petroleum Natural gas liquids	Propane Butane
High	7100-10700 (800-1200)	Natural gas Synthetic natural gas Sour gas	Methane
Medium	2700-7100 (300-800)	Coal gas (O ₂ blown syngas) Coke oven gas Refinery gas	Hydrogen Carbon monoxide Methane
Low	900-2700 (100-300)	Coal gas (air blown syngas)	Carbon Monoxide Hydrogen Nitrogen
Very low	Under 900 (under 100)	Blast furnace gas	Carbon Monoxide Nitrogen

Table 2 - 4Classification of fuel gases [Foster, 2003]

Historically, natural gas has been the primary fuel for gas turbines. According to Table 2-4, comparing to natural gas, the volumetric heating value of cleaned syngas is about 40-50 percent that of natural gas, so a much larger volume of fuel is required with syngas firing to provide the necessary energy input to the gas turbine. Hence, when syngas is used as fuel for modern combustion turbines, there are some process differences. Recently GE initiated a program of extensive analysis to investigate the combustion characteristics of a number of lower-heating-value fuels. Based upon the results of this study, full scale single-burner and sector tests were conducted to confirm expected performance of their MS5000 and LM2500 engine. In general, the only change required to the standard combustion system is modification of the gas fuel nozzle to handle the increased volume of fuel. A variation in heating value of more than 20 percent could be tolerated while still maintaining adequate combustor performance. Many improvements that have maintained flexibility for lower grade fuels have been made in

the modern, higher temperature machines such as the MS6001, MS7001, and MS9001 units [Foster, 2003].

The exhaust temperature from the combustion turbine is generally about 1100°F, which can make additional power through a steam cycle. A HRSG can produce steam by cooling the combustion turbine flue gas. This steam is supplied to a steam turbine to generate additional electric power. In addition, the HRSG is always used to superheat the high-pressure steam generated in the syngas cooler since satisfactory superheater materials have not been demonstrated in the reducing atmosphere of a syngas cooler [Rubin, 1989].

2.3. Literature Review of CO₂ capture from IGCC systems

As one of the most promising technologies for CO_2 capture from coal-fueled power plants, IGCC power plants with CO_2 capture have been studied in some previous studies over the past 15 years [Holt 2003]. These studies covered conceptual technology descriptions, flowsheet modeling and simulation. This section provides a review of the literatures associated with CO_2 capture from IGCC power plants.

Doctor et al. [Doctor 1994, 1996] developed engineering evaluations of CO_2 capture technologies combined with IGCC power plants. The base case for this study was a 458 MW IGCC system that used an air-blown KRW agglomerating fluidized-bed gasifier, Illinois No.6 bituminous coal feed, and in-bed sulfur removal. This study investigated several commercial available chemical and physical solvents for CO_2 capture from IGCC plants, which included amine, glycol , chilled methanol and hot potassium carbonate, and two emerging technologies for CO_2 capture were also considered, which

were high-temperature CO_2 separation with calcium-based sorbents and ambient temperature facilitated transport polymer membranes for acid gas removal. The CO_2 capture efficiency was set to be 90%. From the IGCC plant, a 500-km pipeline took the CO_2 to geologic sequestering. This group also did case studies of Shell gasifier-based multi-product system with CO_2 capture. Life Cycle Analysis (LCA) was adopted in their studies. For these cases, the net electric power production was reduced by 73.6~185.1 MW, with a CO_2 release rate of 0.29~0.53 kg/kWh. The life cycle CO_2 sequestering costs ranged from \$113 to \$201/ton of CO_2 .

Chiesa et al. [Chiesa, 1999] evaluated the energy balances, performance and cost of electricity for two IGCC plants based on oxygen-blown, Texaco gasifiers and large, heavy-duty gas turbines. In one plant, the raw syngas exiting the gasifier was cooled in a high-temperature, radiation cooler; in the other it is quenched by the injection of liquid water. Selexol systems were employed to recovery 90% CO₂ in the syngas after shift reaction. Comparing to the reference plants, the thermal efficiencies of the capture plants were reduced by 5 to 7 percentage points and the cost of electricity were increased by about 40 percent.

Haslbeck et al. [Haslbeck, 2002] investigated CO_2 capture from oxygen-blown, Destec- and Shell-based IGCC power plants. The reference plants fed with Illinois #6 coal, using W501G gas turbine and, three pressure level sub-critical reheat steam cycle, had a net output of 400 MW. Selexol process was used for CO_2 capture with an overall capture efficiency of 87%, and the CO_2 final product was compressed to 2100 psia. For the Destec case with CO_2 capture, the net output was reduced by 42 MW, and the thermal efficiency was decreased by 6.6 percentage points. The COE also showed a corresponding increase to 54.5 from 40.9 MWh. For the Shell case with CO₂ capture, the net output was reduced by 61 MW, and the thermal efficiency was decreased by 7.3 percentage points. The COE showed a corresponding increase to 62.9 from 40.6 MWh.

The study from the report of Parsons [Parsons, 2002] investigated oxygen-blown Egas and GE 7H turbine based IGCC power plants with and without CO₂ capture. For the reference plant, particulate was removed by the hot side filter, MDEA for sulfur removal were employed. For the capture plant, two-stage Selexol are used for H₂S and then CO₂ removal at capture efficiency of 90%, and then CO₂ is compressed to 2200 psig. The net output of the reference plant was 424.5 MW, and the net output of capture plant was reduced by 21 MW. Comparing with the capture plant, the thermal efficiency was decreased by 6.1 percentage points. The COE also showed a corresponding increase to 65.7 from 52.4 \$/MWh at a capacity factor of 65%.

O'Keefe et al. [O'Keefe, 2002] studied a 900 MW IGCC power plant configured to remove 75% of the feed carbon as CO₂. The authors' aim was to present a concept using currently available commercial technology to provide an IGCC plant with the option to capture CO₂. The plant used Texaco Quench gasifiers followed by a sour shift system, a physical absorption acid gas removal, a sulfur recovery system, and a combined cycle unit consisting of two GE 9FA gas turbines and a single steam turbine. The coal feedstock was Pittsburgh 8. Selexol was used for acid gas removal. 75% of the carbon in the coal was removed, but this could be higher. The capture of 75% of the carbon in the coal results in a loss of efficiency of only two percentage points and a decrease in net output of 3%, or 26MW.

A research group of the Foster Wheeler [Foster Wheeler, 2003] assessed the current state of the art of coal-based 750 MWe nominal IGCC, with and without CO₂ capture, and the potential for improvements, between now and 2020. Two types of gasifier were selected, one was the slurry feed gasifier, with product gas cooling by Water Quench (Texaco quench gasifier); the other one was dry feed gasifier, with product gas cooling in a waste heat recovery boiler (Shell gasifier). Several chemical solvents and physical solvents for H₂S removal and CO₂ capture are investigated. An open-cut coal from eastern Australia was used for these plants. All the IGCC plant configurations were based on two 9FA frame gas turbines. The ASU based on the cryogenic process was integrated with gas turbines. Nitrogen produced by the ASU and exceeding the process consumption was injected into the gas turbine for NO_X reduction and power augmentation. Sensitivities to a variety of potentially significant parameters, such as gasification pressure, separate removal of CO₂ and H₂S vs. production of a combined CO₂/H₂S stream, are assessed to help to determine the way forward for IGCC with CO₂ capture. For each alternative plant configuration, overall performances and investment cost were estimated and used to evaluate the electric power production cost. For some alternatives specific optimization studies had been made in order to select the most convenient acid gas removal process and the best arrangement of the shift reactors. This study showed that dry feed gasifier-based IGCC displayed a higher thermal efficiency, however slurry feed gasifier-based IGCC required a lower investment, and in term of cost of electricity and cost of CO_2 recovery, slurry feed gasifier based IGCC was marginally better than the dry feed gasifier based IGCC. The authors also pointed out that the pressure at which gasification was operated was an important design parameter for IGCC optimization.

Increasing the gasifier operating pressure, the heat recovery on the syngas stream was enhanced, the driving force for physical solvent scrubbing of CO_2 was increased and the equipment size was reduced.

According to the above review, the costs of CO₂ capture and sequestration from new IGCC plants added 25-50% to the COE and reduced the thermal efficiency by 10-20% percent. Most studies concluded that the costs of pre-combustion CO₂ capture from syngas in an IGCC plant was much lower than the post combustion removal from Pulverized Coal (PC) or Natural Gas Combined Cycle (NGCC) plants. Most studies focused on the use of bituminous coals. In addition, for bituminous coals the costs of CO₂ removal vary significantly between the various coal gasification technologies and, the advantage in capture costs over PC plants heavily depends on the gasification technology selected.

The IGCC studies surveyed in this section covered the main gasification technologies offered by ChevronTexaco, Shell and ConocoPhillips (E Gas). Among these studies, Texaco quench gasifier was likely to provide the lowest cost option, and the Selexol process was usually used for CO_2 capture from IGCC systems.

REFERENCES (CHAPTER 2)

- 1. Air products, 2004: ITM Oxygen for Gasification, presented at Gasification Technologies Conference, Washington, D.C.
- Brown J.R., Manfredo L., Hoffmann J., and Ramezan M., 2002: Major Environmental Aspects of Gasification-based Power Generation Technologies, Final Report, Project Prepared for and Supported by: Gasification Technologies Program, National Energy Technology Laboratory, DOE
- 3. Bruijn et al, 2003: Treating Options for Syngas, presented at the Gasification Technologies Conference, San Francisco
- 4. Buchanan T.,DeLallo M., Schoff R., and White J., 2002, Evaluation of Innovative Fossil Fuel Power Plants with CO₂ Removal, Technical report prepared for EPRI, Palo Alto, CA
- Cargill P., 2001: Pinon Pine IGCC Project Final Technical Report to the Department of Energy, Sierra Pacific Resources, Reporting Period August 1, 1992 to January 1, 2001
- Chiesa P. and Consonni S., 1999: Shift Reactors and Physical Absorption for Low-CO₂ Emission IGCCs, Journal of Engineering for Gas Turbines and Power, Vol. 121
- Clayton S.J., Stiegal G.J., Wiemer J.G., 2002: Gasification Markets and Technologies – Present and Future: An Industry Perspective, National Energy Technology Laboratory (NETL) U.S. Department of Energy (Report No. DOE/FE-0447)
- 8. Doctor R.D., etc., 1994: Gasification combined cycle: carbon dioxide recovery, transport, and disposal, Report prepared by ANL/ESD
- 9. Doctor R.D., etc., 1996: KRW oxygen-blown gasification combined cycle carbon dioxide recovery, transport, and disposal, Report prepared by ANL/ESD
- Edward L. P., Shelton W.W., and Lyons J.L., 2002: Advanced Fossil Power Systems Comparison Study, Final report prepared for National Energy Technology Laboratory
- 11. Foster, A.D., Doering H.E., and Hilt M.B., 2003, Fuels Flexibility in Heavy Duty Gas Turbines, GE Company Schenectady, New York
- 12. Foster Wheel, 2003: Potential for improvement in gasification combined cycle power generation with CO₂ capture. Report # PH4/19, May 2003

- O'Keefe L.F., Griffiths J., Weissman R.C., De Puy R.A., and Wainwright J.M., 2002: A Single IGCC Design for Variable CO₂ Capture, Fifth European Gasification Conference
- 14. Haslbeck J.L., 2002, Evaluation of Fossil Power Plants with CO₂ Recovery, Parsons Infrastructure & Technology Group Inc., February 2002
- 15. Holt N., Booras G., and Todd D., 2003: A Summary of Recent IGCC Studies of CO₂ Capture for Sequestration By Presented at The Gasification Technologies Conference San Francisco, CA
- 16. Holt N., 2001: Coal Gasification Research, Development and Demonstration Needs and Opportunities, 2001 Gasification Technologies Conference, sponsored by the gasification Technologies Council and EPRI, San Francisco, CA
- Korens N., Simbeck D.R., Wilhelm D.J., 2002: Process Screening Analysis of Alternative Gas Treating and Sulfur Removal for Gasification, Revised Final Report prepared for National Energy Technology Laboratory, U.S. Department of Energy
- Joshi M.M., 2000, Quality Assured maintenance Management For Coal Handling Plant. <u>http://www.plant-</u> maintenance.com/articles/Maintenance_Management_QA.pdf
- McCarthy, C.B., and Clark W.N., 1985: Integrated gasification/combined cycle (IGCC) electric power production- A rapidly emerging energy alternative. Presented at symposium on coal gasification and synthetic fuels for power generation, EPRI, Palo alto, CA
- 20. O'Brien J.N., Blau J., Rose M., 2004: An Analysis of the Institutional Challenges to Commercialization and Deployment and Deployment of IGCC Technology in the U.S. Electricity Industry: Recommended Policy, Regulatory, Executive and Legislative Initiatives, Final Report, Project Prepared for and Supported by: DOE, NETL, Gasification Technologies Program and National Association of Regulatory Utility Commissioners
- 21. Parsons E., NETL, "Advanced fossil power systems comparisons study", NETL report, 2002
- 22. Rosenberg W.G., Alpern D.C., Walker M.R., 2004: Deploying IGCC in This Decade With 3 Party Covenant Financing, Volume I
- 23. Rubin, E., 1989: Implications of Future Environmental Regulation of Coal-Based Electric Power, Annual Rev. Energy, 14:19-45
- 24. SFA Pacific, Inc., 2003: Evaluation of IGCC to Supplement BACT Analysis of Planned Prairie State Generating Station

- 25. Simbeck D.R., 2002: Industrial Perspective on Hot Gas Cleanup, Presentation at the 5th International Symposium on Gas Cleaning at High Temperatures, U. S. DOE National Energy Technology Laboratory
- 26. Stiegel G.J., Clayton S.J., and Wimer J.G., 2001: DOE's Gasification Industry Interviews: Survey of Market Trends, Issues and R&D Needs, 2001 Gasification Technologies Conference, sponsored by the Gasification Technologies Council and EPRI, San Francisco, CA
- 27. Todd D.M., 1994: Clean Coal and Heavy Oil Technologies for Gas Turbines, Paper No. GER-3650D, 38th GE Turbine State-of-the-Art Technology Seminar, GEZ-7970, General Electric Company, Schenectady, NY
- 28. U.K. Department of Trade and Industry, 1998: Gasification of Solid And Liquid Fuels For Power Generation - Status Report, <u>http://www.dti.gov.uk/cct/pub/tsr008.pdf</u>
- 29. Wabash Energy Ltd, 2000: The Wabash River Coal Gasification Repowering Project, U.S. Department of Energy Topical Report Number 20, prepared for National Energy Technology Laboratory, U.S. Department of Energy
- 30. World Bank, 2000: <u>http://www.worldbank.org/html/fpd/em/power/EA/mitigatn/igccsubs.stm#igcctec</u> <u>h</u>
- 31. NETL, 2000: http://www.netl.doe.gov/coalpower/gasification/description/gasifiers.html#Lurgi

Chapter 3. PERFORMANCE AND ECONOMIC SIMULATION MODEL OF IGCC SYSTEMS

The purpose of this chapter is to summarize the information of IGCC modeling process. Plants with and without CO_2 capture are modeled with Aspen Plus. The details include the design basis, the mass and energy balances of major units of the IGCC systems, design specifics which is required by commercial available technologies, and the convergence sequence, which specifics the calculation sequence of the simulation model.

3.1. Model design basis

Figure 3-1 provides a simplified overview of the reference plant configuration, which does not incorporate CO_2 recovery. This layout is a typical oxygen-blown IGCC with cold-gas cleanup in which H₂S is captured by an acid gas removal system. The cleaned gas is then saturated and reheated before it is fed into gas turbine combustion chamber. The hot exhaust from gas turbine is used to generate steam for a steam cycle through the HRSG. Oxygen is supplied by an air separation unit.

Figure 3-2 is an overview of the capture plant configuration. Comparing to the reference plant, a water gas shift reactor is added to increase CO_2 partial pressure through converting CO into CO_2 . CO_2 is captured in a Selexol process, a commercial glycol-based process for acid gas removal.



Figure 3 - 1 An IGCC system without CO₂ capture





The objective of this modeling study is to assess coal-based IGCC plants with and without CO_2 capture based on current commercial available technology, hence equipment

selection and system design of this IGCC system focuses on mature technologies and methodologies

3.1.1. Gasification technology selection

Currently, many different types of gasifier are commercially available, such as Texaco, Shell, E-gas. This study is based on the oxygen-blown, slurry feed Texaco quench gasifier with product gas cooling by water quench, which is the most widely used gasifier type in IGCC plants. In addition, the Texaco quench gasifier is known for its low-capital cost requirement.

Currently available Texaco gasifiers can be operated in a wide pressure range, from 15 bar to 70 bar. For a given capacity an increase of the gasification pressure will reduce the size of the equipment but increase the operating costs. In this study, a medium pressure (42 bar) gasifier is adopted. Considering that the final destination of the syngas in an IGCC is the combustion chamber of the gas turbine, which available today with a rang of 20~30 bar, a gas expander is installed between the gasification and the gas turbine, which can offset the extra operation cost due to the high gasification pressure.

3.1.2. Air separation unit

The state-of-the-art air separation plants are based on cryogenic mechanism. Various ASU configurations can be used in IGCC systems, ranging from complete integration, in which all of the air for the ASU is provided by the gas turbine compressor, to zero integration in which the ASU is a completely stand-alone unit providing only oxygen to gasifiers. Considering operation stability and flexibility, the "stand alone" option is employed. For the "stand alone" option, low pressure air separation plant is chosen, with its own air compressors delivering air to the cryogenic process at the minimum pressure requirement to meet the energy demand of the process. In this case, syngas humidification is generally preferred to nitrogen addition for NO_x control to avoid the large nitrogen compression energy consumption.

3.1.3. Syngas clean up system

Raw syngas from gasification unit is hot, humid and contaminated with H_2S , and COS. Before used as gas turbine fuel, this raw syngas has to be cleaned by removing all the contaminants and prepared at the proper conditions of temperature, pressure and water content to meet the requirement of the gas turbine combustion under conditions of desired environmental performance and operation stability.

The key factor in achieving the environmental performance of IGCC systems is sulfur removal from the syngas. Sulfur is contained in two types of acid gases, H_2S and COS. The first step in the sour gases removal process is to remove the carbonyl sulfide (COS) from the gas stream. For an IGCC system without CO₂ capture, the conventional method is to pass the syngas through a fixed bed, catalytic hydrolysis reactor, which will hydrolyze the COS to CO₂, H_2S and CO. Hence for the plant without CO₂ capture, a particle scrubber is employed to remove solids entrained in the syngas, then COS hydrolysis reactors are used to converted COS into H_2S . The Selexol/Claus/SCOT process is used for sulfur removal and recovery.

For the capture plant, after particle removal, the water gas shift reactor is used to convert CO into CO_2 . The CO shift catalyst also hydrolyses COS to H_2S . Hence there is

no need of a separate COS hydrolysis system. Two types of catalyst are usually used for the water gas shift reaction [IEA, 2003]:

- Sour shift catalysts based on Co-Mo, which operate at medium/high temperature and requiring a steam/dry gas volume ratio in the range of 1.1-1.6. This type of catalysts can withstand high concentration of sulfur in syngas.
- Clean shift catalysts based on Fe-Cr or Cu, which operate at high temperature or low temperature and require a steam/dry gas volume ratio equal to 1. For this type of catalyst, the total sulfur content of syngas should be less than 10 ppm.

For IGCC systems with Texaco quench design, preliminary thermodynamic analysis shows that sour shift dominates the clean shift option because syngas at particle scrubber outlet has all the characteristics required by the sour shift reaction (temperature and steam to carbon ratio). In the capture plant, the acid gases, H₂S and CO₂, are removed through two Selexol processes, separately.

3.1.4. Gas turbine selection and steam cycle design

Syngas produced by gasification process is a type of Low Calorific Value (LCV) fuel. GE gas turbines applied to LCV applications have accumulated rich experience and hold a leading position in this field. In this study, GE 7FA is selected, which has been designed to operate at base load conditions on syngas at Tampa Electric [Brdar, 2003].

Selection of a single- or multiple-pressure steam cycle for a specific application is determined by economic evaluation, which considers the plant-installed cost, fuel cost and quality, plant-duty cycle, and operating and maintenance cost. According to the recommendation of GE, single- and multiple-pressure non-reheat steam cycles are applied to systems equipped with GE gas turbines that have rating point exhaust gas temperatures of approximately 1000°F / 538°C or less. Multiple-pressure reheat steam cycles are applied to systems with GE gas turbines that have rating point exhaust gas temperatures of approximately 1100°F / 593°C or greater [Chase, 2003]. Table 3-1 gives such recommendations in detail. Since the exhaust gas temperature of GE 7FA is approximately 1104 °F/596 °C, a three-pressure reheat steam cycle is employed in this study.

Heat Recovery Steam Cycle	Non-Reheat Three-Pressure			Reheat Three-Pressure
Steam Turbine Size (MW)	\leq 40	> 40 < 60	≥ 60	> 60
Throttle Pressure (psig)	820	960	1200	1400-1800
Throttle Temperature (°F)	40 approach to Gas Turbine Exhaust Gas Temperature			1000-1050
Reheat Pressure (psig)				300-400
Reheat Temperature (°F)				1000-1050
IP Admission Pressure (psig)	100	120	155	300-400
IP Admission Temperature (°F)	20 Approach to Exhaust Gas Ter Superheater			mperature upstream of
LP Admission Pressure (psig)	25	25	25	40
LP Admission Temperature (°F)	20 Approach to Exhaust Gas Temperature upstream of superheater			mperature upstream of

Table 3 - 1Steam and gas product line steam turbine throttle and admission steam
conditions

3.2. Major process sections of the IGCC model

The present model consists of slurry preparation units, gasification units with quench, particle removal, low temperature gas cooling and clean up units, fuel expender, fuel gas saturator and reheater, by product sulfur production, gas turbine system, steam cycle system, and heat integration among different units. In addition to these units and heat integration, the model also incorporates auxiliary power consumption for ASU and summary files of the whole IGCC system.

Each major process section mentioned above is referred to as a flowsheet section in Aspen models. Within each flowsheet, unit operation models represent specific components of that process. There are user-specified inputs regarding key design assumptions for each unit model. The numerical values of these design assumptions are shown in this report. However, users can change these values to simulate their specific design alternatives. The major flowsheet sections in the IGCC system are presented as in the following.

3.2.1. Coal slurry preparation and gasification flowsheet

Coal from the coal grinding system is continuously fed to the grinding mill. Grey water from waste water treatment facility is used for slurrying the coal feed. The coal slurry with a desired slurry concentration is pumped into the gasifier. In this section, the methodology used to model coal preparation is presented.

Coal is a type of non-conventional solid, and its composition has to be input by the user in forms which Aspen accepts. In Aspen, the component attributes of coal are specified in three forms: PROXANAL for proximate analysis, ULTANAL for ultimate analysis, and SULFANAL for sulfur analysis. Table 3-2, as an example, gives the typical compositions of Pittsburgh #8 coal and its input values for Aspen model. Aspen Plus estimates the heat of coal combustion based on its PROXANAL, ULTANAL, and SULFANAL. Users can also enter the heat of combustion directly.

Coal composition (wet basis)		PROXANAL		ULTANAL		SULFANAL	
Element	Value	Element	Value	Element	Value	Element	Value
ASH	7.24	MOISTURE	5.05	ASH	7.63	PYRITIC	1.23
CARBON	73.81	FC	49.855	CARBON	77.74	SULFATE	0
HYDROGEN	4.88	VM	42.515	HYDROGEN	5.14	ORGANIC	1
NITROGEN	1.42	ASH	7.63	NITROGEN	1.5		
CHLORINE	0.06			CHLORINE	0.06		
SULFUR	2.13			SULFUR	2.23		
OXYGEN	5.41			OXYGEN	5.7		

 Table 3 - 2
 Coal composition and its corresponding input in Aspen Plus



Figure 3 - 3 Slurry preparation and gasification flowsheet

Figure 3-3 illustrates the mass and heat flows in the coal slurry preparation process and gasification units, and Table 3-3 shows the corresponding unit operations that are simulated in Aspen Plus. The coal slurry is compressed up to 710 psia through a slurry pump, which is simulated by a unit named as "SlurryPump". Gasification simulation calculates the Gibbs free energy of the coal. However, the Gibbs free energy of coal cannot be calculated because it is a non-conventional component. Hence, a RYield unit, which simulate a reactor with a known yield, and does not require reaction stoichiometry and kinetics, named as "DeCoal" is used to decompose the coal into its constituent elements based on the ultimate composition analysis of coal.

The gasification unit converts coal slurry into syngas. The coal slurry and oxygen from the air separation unit react in the gasifier at high temperature (approximately 2450 °F), high pressure (approximately 620 psia in this study) and under the condition of insufficient oxygen to produce syngas. Syngas consists primarily of hydrogen and carbon monoxide with lesser amounts of water vapor, carbon dioxide, hydrogen sulfide, methane, and nitrogen. Traces of carbonyl sulfide and ammonia are also formed. Ash presenting in the coal melts into slag. Hot syngas and molten slag from the gasifier flow downward into a quench chamber, which is filled with water, and is cooled into medium temperature (approximately 450 °F). The slag solidifies and flows to the bottom of the quench chamber.

In addition to CO, H_2 and CO₂, small amounts of CH₄, HCl, COS and NH₃ are also formed. Various amounts of H_2S depending on the sulfur content of the feed coal.

The Texaco process uses an entrained flow gasifier. Slagging is an important problem with this type of gasifier. The slagging formation is modeled as follows.

In this study the gasification process is modeled on the fixed carbon conversion model [Altafini, 2003; Zaimal, 2002], and simulated in three steps. At first, slag formation is simulated in the Slag block based on the following stoichiometric reaction. The stoichiometric coefficients of carbon and ash are determined by the ash percentage in the coal and carbon loss in the gasification process.

mC+nAsh \rightarrow Slag

Second, coal slurry from the Slag block mixed with oxygen from the ASU unit enter the gasifier reactor, reactions occur in the reactor is simulated in GasifRx unit, which is based on RGibbs model. RGibbs models chemical equilibrium by minimizing Gibbs free energy. Chemical reactions and their approach temperatures¹ modeled in this equilibrium gasifier reactor are as follows [Altafini, 2003; Zaimal, 2002, Zhu, 2003]:

C+2H₂ \rightarrow CH₄ (approach temperature: -300°F)

 $C+H_2O\rightarrow CO+H_2$

 $C+O_2 \rightarrow CO$

 $2CO+O_2 \rightarrow 2CO_2$ (approach temperature: -550°F)

 $CH_4+2O_2 \rightarrow CO_2+2H_2O$ (approach temperature: -500°F)

S+H₂ \rightarrow H₂S (approach temperature: -500°F)

 $N_2+3H_2 \rightarrow 2NH_3$ (approach temperature: -500°F)

 $CO+H_2S \rightarrow COS+H_2$ (approach temperature: -500°F)

¹ The approach temperature is a pseudo-temperature used in Aspen to adjust calculated equilibrium concentrations to actual (observed) values under non-equilibrium conditions.

 $Cl_2+H_2 \rightarrow 2HCl$ (approach temperature: -300°F)

The reaction temperature and heat loss, which is assumed to be 1% of the total low heating value of the inlet coal flow, in the gasification reactor is maintained by adjusting the inlet flow rate of oxygen.

Third, raw syngas and molten slag discharge from the reactor into the quench chamber, which is simulated by the Quench unit. The Quench unit performs rigorous vapor-liquid equilibrium calculations to determine the thermal and phase conditions of syngas saturation process. In this quench unit, molten slag is cooled down and separated from the syngas.

No	Aspen unit ID	Unit parameters	Unit function
1	CoalMult (Mult)	Multiplication factor: 0.5~5	Manipulate slurry flow rates through Design Spefic
2	SlurryPMP (Pump)	Discharge P=710 psi Efficiency=Default	This unit simulate coal slurry pump
3	DeCoal(RStoic)	Pressure=620 psi Temperature=59 F	This block decomposes coal into its elements using the Calcularor
4	MkSlag (RStoic)	Pressure=620 psi Temperature=59 F	Simulate the stoichiometric reaction which produces slag based on the coal's ultimate analysis and carbon loss percentage in gasification process
5	GasifRX (RGibbs)	Pressure=620 psi Temperature=2450 F Products: O2, N2, H2, CO, CO ₂ , H ₂ O, CH ₄ , H ₂ S, NH ₃ , COS, HCL	Simulate the stoichiometric reactions occurring in the gasifier. Heating loss in the gasifier is maintained as 1% of the total LHV of coal through Design Specific
6	Quench (Flash 2)	Pressure drop=15 psi Heat duty=0 But/hr	Simulate the quench process of syngas, slag cooling and separation

Table 3 - 3Coal slurry preparation and gasification process unit description.

3.2.2. Low temperature gas cooling and clean up

Raw syngas from the quench chamber enters the scrubber to remove solid particles. Figure 3-4 illustrates the mass flows in the syngas cooling process, and Table 3-4 shows the corresponding unit operations that are simulated in Aspen Plus. The scrubber is simulated by the block PartRemv, which separates solids from the syngas. As the first step of sour gases removal, the syngas passes through a fixed bed, catalytic hydrolysis reactor, which hydrolyzes the COS to CO₂ and H₂S, and the HCN to NH₃ and CO. Activated alumina type catalysts are generally employed for this application, and COS concentrations approaching equilibrium levels can be achieved. This reactor is modeled by a block named COSHydro, which is a rigorous equilibrium reactor based on stoichiometric approach for the following hydrolysis reaction:

$COS+H_2O\rightarrow H_2S+CO_2$

Syngas after COS hydrolysis is at a temperature of approximately 460 °F, which has to be cooled down to approximately 100 °F for H₂S removal. Blocks named as SgasCol1~5 simulate this cooling process. Condensate water from this cooling process is collected for the syngas quench and scrubber processes. Heat released from syngas cooling is recovered to produce low pressure steam (390°F/48 psia) for steam cycle, and intermediate pressure hot water (408°F/325 psia) for syngas saturation and reheating. The flow rate of feed water for the low pressure steam and intermediate pressure hot water is manipulated by the Design Specification SGTEMP, which adjusts the feed water flow rate to satisfy the syngas temperature at the exit of the last syngas cooler is 100°F. The flow rate ratio of the low pressure steam to the intermediate pressure hot water is controlled by another Design Specification SGIPFLOW, which adjusts the flow rate of the intermediate pressure hot water to meet the need of syngas saturating and reheating.

No	Aspen unit ID	Unit parameters	Unit function
1	Scrubber (Flash 2)	Pressure drop=10 psia Heat duty=0 But/hr	Simulate the scrubber process of particle removal from raw syngas
2	COSHydro (REquil)	Pressure drop=5 psia Heat duty=0 But/hr	Simulate the COS hydrolysis process converting COS into H ₂ S
3	SgasCol1 (Heater)	Pressure drop=3 psia	Simulate syngas cooler
4	SgasCol2 (Heater)	Pressure drop=3 psia	Simulate syngas cooler
5	SgasCol3 (Heater)	Pressure drop=5 psia	Simulate syngas cooler
6	SgasCol4 (Heater)	Pressure drop=3 psia	Simulate syngas cooler
7	SgasCol5 (Heater)	Pressure drop=5 psia	Simulate syngas cooler
8	FWPMP2 (Pump)	Discharge pressure=18 psia Efficiency=default value	Simulate feed water pump
9	FWPMP3 (Pump)	Discharge pressure=18 psia Efficiency=default value	Simulate feed water pump
10	FWSPLIT (SPLIT)		Indicate that feed water is divided into two streams, the flow rate to the LP steam evaporator is manipulated by the Design Specification SGIPFLOW

Table 3 - 4Syngas cooling process unit description of the reference plant


Figure 3 - 4 Syngas cooling section flowsheet of the reference plant

Figure 3-5 illustrates the mass flows in the syngas cooling process of the capture plant, and Table 3-5 shows the corresponding unit operations that are simulated in Aspen Plus. For the capture plant, syngas from the scrubber is at a temperature of approximately 420 °F. The water gas shift reaction occurs at two rectors, the high temperature reactor and the low temperature reactor, which are simulated by block HTShift and LTShift. In the shift reactors, the following reactions occur in the presence of catalysts:

 $CO+H_2O\rightarrow CO_2+H_2$

$$COS+H_2O\rightarrow H_2S+CO_2$$

Because the shift reaction is exothermic, there is high quality energy available for generating high pressure and intermediate pressure steam during the syngas cooling process. Blocks named as HTCol1~4 and LTCOl1~5 simulate this cooling process. The condensate water from this cooling process is collected for syngas quench and scrubber processes. The high pressure steam is sent to the high pressure steam turbine. Part of the intermediate pressure is used for syngas reheating, and the rest is sent to the steam cycle. The flow rate of the feed water for the high pressure steam and the intermediate pressure steam is manipulated by the Design Specification SGTEMP, which adjusts the feed water flow rate ratio of intermediate pressure steam to the high pressure steam is controlled by another Design Specification SGSHIFT, which adjusts the flow rate of the high pressure feed water for the high pressure steam is controlled by

No	Aspen unit ID	Unit parameters	Unit function	
1	Scrubber (Flash 2)	Pressure drop=10 psia Heat duty=0 But/hr	Simulate the scrubber process of particle removal from raw syngas	
2	SgasHet (Heater)	Pressure drop=4 psia Temperature=469.4 F	Simulate the syngas heater	
3	HTShift (Requil)	Pressure drop=4 psia Heat duty=0 Btu/hr	Simulate the high temperature shift reactor	
4	HTCol1 (Heater)	Pressure drop=4 psia	Simulate syngas cooler	
5	HTCol2 (Heater)	Pressure drop=3 psia	Simulate syngas cooler	
6	HTCol3 (Heater)	Pressure drop=4 psia	Simulate syngas cooler	
7	HTCol4 (Heater)	Pressure drop=5 psia	Simulate syngas cooler	
8	LTShift (Requil)	Pressure drop=5 psia Heat duty=0 Btu/hr	Simulate the low temperature shift reactor	
9	LTCol1 (Heater)	Pressure drop=4 psia	Simulate syngas cooler	
10	LTCol2 (Heater)	Pressure drop=4 psia	Simulate syngas cooler	
11	LTCol3 (Heater)	Pressure drop=4 psia	Simulate syngas cooler	
12	LTCol4 (Heater)	Pressure drop=4 psia	Simulate syngas cooler	
13	LTCol5 (Heater)	Pressure drop=5 psia	Simulate syngas cooler	
14	FWPMP1 (Pump)	Discharge pressure=365 psia Efficiency=default value	Simulate feed water pump	
15	FWPMP3 (Pump)	Discharge pressure=1734 psia Efficiency=default value	Simulate feed water pump	
16	FWSPLIT (SPLIT)		This unit is used to indicate that feed water is divided into two streams, the flow rate to the intermediate pressure steam evaporator is manipulated by the Design Specification SGSHIFT	

Table 3 - 5Syngas cooling process unit description of the capture plant



Figure 3 - 5 Syngas cooling section flowsheet of the capture plant

3.2.3. H₂S capture and sulfur recovery section

After COS hydrolysis, almost all of the sulfur in the gasifier feedstock is converted into H_2S . Figure 3-6 illustrates the mass and energy flows in the sulfur removal and recovery section, and Table 3-6 shows the corresponding unit operations that are simulated in Aspen Plus. In this modeling study, Selexol process, a physical solvent system, is employed to capture H_2S . A block, named as SulfSep, is used to simulate this Selexol process. In this block, approximately 99% of H_2S is removed from the syngas. The H_2S rich gas and the flash gas from the Selexol process are sent to the Claus/Scot unit for sulfur recovery.

The Claus process has been the sulfur recovery workhorse for applications with large amounts of sulfur. The Clause process is carried out in two stages. In the first stage, about one third of the gases from the Selexol unit, which exits at about 120 °F, are burned in the first furnace. This first furnace is simulated by the block named Furnace. In this block, the following reaction is modeled based on stoichiometric mechanism which is close to the real situation in the reactor. Low pressure air from an air compressor is used as the oxidant of Claus reaction.

 $H_2S+O_2 \rightarrow SO_2+H_2O$ with 33% of H_2S converted

The remaining acid gases enter the second stage furnace, where the H_2S and SO_2 react in the presence of a catalyst to form elemental sulfur:

$$2H_2S+SO_2 \rightarrow 3S+2H_2O$$

The gas is cooled in a waste heat boiler and then sent through a series of reactors where more sulfur is formed. The sulfur is condensed and removed between each reactor. An Aspen block, ClausRxr, is used to simulate the reactors, where 98% of H_2S is recovered because the Claus process is limited by chemical equilibrium to removal efficiencies of approximately 98% if three catalytic reactor stages are employed. To achieve higher removal efficiencies, a tail gas treating unit is required.

SCOT process is a conventional tail gas treating process. In the process, the tail gas from the Claus unit and the flash gas from the Selexol unit are heated to approximately 570 °F in an in-line burner, which serves the dual purpose of heating the gas stream and producing a reducing gas, which is needed in the downstream reactor. The effluent from the burner is then passed over a cobalt-molybdenum catalyst. In the reactor, all of the SO₂ and CS₂ are converted to H₂S by a combination of hydrogenation and hydrolysis reactions. This process is modeled by the block named BsComb. In this block, the following combustion and hydrogenation reactions occur:

 $CO+0.5 \rightarrow CO_2$ $CH_4+2O_2 \rightarrow CO_2+2H_2O$ $SO_2+3H_2 \rightarrow H_2S+2H_2O$ $COS+H_2O \rightarrow H_2S+CO_2$

The reactor effluent gas is then cooled and processed through a Stretford unit, where H_2S is converted to elemental sulfur, and remaining gases exhaust to the atmosphere. The block named StretFrd is used to model this process, where the following reactions occur:

$$2H_2S+O_2 \rightarrow 2S+2H_2O$$

 $2H_2+O_2 \rightarrow 2H_2O$

The block QMix, which simulates the heat recovery process of the waste heat boiler, collects heat from the above reactors to preheat the feed water from the steam cycle.

No	Aspen unit ID	Unit parameters	Unit function
1	SulfSep (Sep)	Flue gas: T=85 F, P=32 atm Acid gas: T=120 F, P=22 psia Flash gas: T=58 F, P=115 psia	This unit simulates the Selexol process for H ₂ S removal.
2	CAirComp (Compr)	Type: Isentropic Discharge pressure=23 psia Isentropic efficiency=0.9	Model the air compressor for Claus process
3	CAirMix1 (Mixer)		Simulate the mixer of air and acid gas from Selexol unit
4	Furnace (RStoic)	Pressure drop=0 psia Temperature=589 F	Simulate the first stage of Clause process, where about one third of acid gas from Selexol process is burned.
5	ClausRxR (RStoic)	Pressure drop=0 psia Temperature=589 F	Simulate the second stage of Clause process, where the H_2S and SO_2 react in the presence of a catalyst to form elemental sulfur.
6	ClausSep (Sep)		Simulate the sulfur removal process between each reactor, water condensate, and tail gas separation.
7	BsComp1 (Compr)	Type: Isentropic Discharge pressure=30 psia Isentropic efficiency=0.9	Model the air compressor for Scot process
8	BsComp2 (Compr)	Type: Isentropic Discharge pressure=30 psia Isentropic efficiency=0.9	Model the tail gas compressor for Scot process
9	BsMix (Mixer)		Simulate the mixer of tail gas and air
10	BsComb (RStoic)	Pressure drop=0 psia Temperature=400 F	Simulate the tail gas treatment process, which converts SO_2 into H_2S with the aid of a cobalt- molybdate catalyst
11	Stretfrd (RStoic)	Pressure drop=0 psia Temperature=100 F	Simulate sulfur recovery process, where H ₂ S reacts with O ₂ to generate Sulfur
12	QMix (Mixer)		Simulate the waste heat boilers which recover heat generated in sulfur recovery process for feed water heating in steam cycle

Table 3 - 6Sulfur removal and recovery unit description



Figure 3 - 6 Sulfur removal and recovery section flowsheet

3.2.4. Clean syngas saturation, expend, and reheat section

Clean syngas from the Selexol unit for sulfur removal could be used as the fuel of the gas turbine. In order to meet the emission and pressure requirements of the gas turbine combustion, the fuel is saturated, expended, and preheated before entering the combustion chamber. For the reference plant, fuel from the Selexol unit at a temperature is heated up by the condensate water from the syngas cooling process in the heat, which is simulated by the block FuelHet1. The heated fuel is expended in a turbine expender to generate electricity, and its pressure is reduced to match the pressure at the gas turbine combustor. Fuel from the expender enters the saturator to mix with the intermediate pressure hot water produced in the syngas cooling process. Before entering the gas turbine combustion chamber, the saturated fuel is preheated up to about 400 °F by the intermediate pressure hot water from syngas cooling unit. Figure 3-7 illustrates the mass and energy flows in the sulfur removal and recovery section, and Table 3-7 shows the corresponding unit operations that are simulated in Aspen Plus.

No	Aspen unit ID	Unit parameters	Unit function
0	CO ₂ Sep (Sep)		Simulate the CO_2 capture process in a Selexol unit for the capture plant
1	FuelHet1 (Heater)	Pressure drop=5 psia Temperature=290 F	Simulate the fuel heater
2	FuelExpd (Compr)	Type: Isentropic Discharge pressure=280 psia Isentropic efficiency=0.9	Simulate the fuel expender which reduce the pressure of fuel to match the requirement of gas turbine combustor requirement
3	Satur (Flash 2)	Pressure drop=15 psi Heat duty=0 But/hr	Simulate the fuel saturator, where water steam volume percentage in the fuel is increased up to 16%
4	FuelHet2 (Heater)	Pressure drop=5 psia Temperature=401 F	Simulate the fuel heater which heat the fuel to a temperature of 401 F

 Table 3 - 7
 Sulfur removal and recovery unit description



Figure 3 - 7 Fuel saturation and reheat section flowsheet of the reference plant

For the capture plant, fuel from the Selexol unit is sent to another Selexol unit for CO_2 capture, which is simulated by the block CO_2Sep . The CO_2 capture efficiency and power consumption is calculated based CO_2 capture model, which will be discussed later. Fuel after CO_2 capture is heated up by the hot water from syngas saturator. The fuel heater is simulated by the block FuelHet1. The heated fuel is expended in a turbine expender to generate electricity, and its pressure is reduced to match the pressure at the gas turbine combustor. Fuel from the expender enters the saturator to mix the intermediate pressure saturation water to added warm steam in the syngas. Before entering the gas turbine combustion chamber, the saturated fuel is preheated by the intermediate pressure steam from syngas cooling unit. Figure 3-8 illustrates the mass and energy flows in the sulfur removal and recovery section, and Table 3-7 shows the corresponding unit operations that are simulated in Aspen Plus.



Figure 3 - 8 CO₂ capture, fuel saturation, and reheat section flowsheet of the capture IGCC power plant

3.2.5. Gas turbine section

The gas turbine section design bases on the GE 7FA gas turbine system. Although the original turbine design specifications are based on a natural gas rather than a coal derived syngas, GE heavy-duty gas turbines have operated successfully burning alternate gaseous fuels with heating values ranging from 11.2 to 116 MJ/m³ (300 to 3100 Btu/ft³ lower heating value) [Foster, 2003]. Figure 3-9 illustrates the mass and energy flows in

gas turbine section, and Table 3-8 shows the corresponding unit operations that are simulated in Aspen Plus.

The pressure ratio of GE 7FA is 15.5, hence the air at the ambient conditions (59 F, 14.7 psia, and 60 percent relative humidity) is compressed up to 230 psia at a three-stage compressor. The pressure ratio of each compression stage is one third of the total pressure ratio. The compressor has several extraction points, from which some amount of compressed air is removed and injected into the blades and vanes of the hottest turbine stages for cooling. For GE 7FA gas turbine, approximately 11% of the total air flow rate is used for gas turbine cooling.

The three-stage compressor is simulated by three units, GTComp1, GTComp2, and GTComp3. The outlet pressures for these three stages are 37.82, 93.3 and 230 psia, respectively. Three cooling air streams are moved at the outlet of each stage for turbine cooling.

The saturated and reheated fuel and the air from the last stage of the compressor enter the gas turbine combustion chamber, which is simulated by the block GTBurn. The following chemical reactions are employed to simulate the combustion process:

 $2CO+O_2 \rightarrow 2CO_2$

 $2H_2+O_2 \rightarrow 2H_2O$

 $CH_4+1.5O_2 \rightarrow CO+2H_2O$

 $2H_2S+3O_2 \rightarrow 2H_2O+2SO_2$

 $COS+1.5O_2 \rightarrow CO_2 + SO_2$

$$2NH_3+2.445O_2 \rightarrow 0.1N_2+1.71NO+0.09NO_2+3H_2O$$

 $N_2+1.05O_2 \rightarrow 1.9NO+0.1NO_2$

An amount of intermediate pressure steam from the steam cycle is used for the combustion chamber cooling. The cooling process is simulated by the block GT_Qloss. The heated steam comes back to the steam cycle. The firing temperature of GE 7FA gas turbine is approximately 2350 °F, this temperature is maintained by a Design Specification TIT, which manipulates the inlet temperature of the first stage gas turbine by adjusting the flow rate of the coal slurry.

Hot combustion product gases enter the three-stage turbines at pressures of 228, 92.45, 37.29 psia, respectively. The outlet pressure of the last stage turbine is 15.2 psia. The three turbines are modeled by three blocks, GTTurb1, GTTurb2, and GTTurb3. The exhaust temperature of GE 7FA is 1106 °F, which is maintained through a Design Specific TEXHAUST. The hot exhaust gases enter the HRSG to produce steam for the steam cycle.

The overall mass flow rate in a gas turbine is typically limited by the turbine nozzle. When the March number at the turbine nozzle is unity, the flow at the inlet of gas turbine expender is choked. The choke flow rate calculation used in this model based on the model developed by Frey [Frey, 2001]. The design specification TCHOKE sets the flow rate of air at the compressor inlet to meet the choked flow condition.



Figure 3 - 9 Gas turbine section flowsheet

No	Aspen unit ID	Unit parameters	Unit function
1	GTMix1 (mixer)		Simulate the mixing of fuel and the compressed air for gas turbine combustion
2	GTMix2 (mixer)		Simulate the mixing of cooling air and the combustion products
3	GTMix3 (mixer)		Simulate the mixing of cooling air and the combustion products
4	GTMix4 (mixer)		Simulate the mixing of cooling air and the combustion products
5	GTComp1 (Compr)	Discharge pressure=37.82 psia Isentropic=0.918	Simulate the fist stage of gas turbine compressor
6	GTComp2 (Compr)	Discharge pressure=93.3 psia Isentropic=0.918	Simulate the fist stage of gas turbine compressor
7	GTComp3 (Compr)	Discharge pressure=230 psia Isentropic efficiency=0.918	Simulate the fist stage of gas turbine compressor
8	GTBurn (RStoic)	Pressure=228 psia Heat duty=0 Btu/hr	Simulate the gas turbine combustor
9	GT_Qloss (Heater)	Pressure drop=0 psia Temperature change=16 F	Simulate the heat loss in the combustor during to cooling process
10	GTTurb1 (Compr)	Discharge pressure=92.45 psia Isentropic=0.919	Simulate the first stage of the gas turbine
11	GTTurb2 (Compr)	Discharge pressure=37.49 psia Isentropic=0.919	Simulate the second stage of the gas turbine
12	GTTurb1 (Compr)	Discharge pressure=15.2 psia Isentropic=0.919	Simulate the third stage of the gas turbine
13	GTSplit1 (Split)		This block splits the compressed air from the first stage of the gas turbine compressor for gas turbine first stage rotor cooling
14	GTSplit2 (Split)		This block splits the compressed air from the first stage of the gas turbine compressor for gas turbine second stage vane cooling
15	GTSplit1 (Split)		This block splits the compressed air from the first stage of the gas turbine compressor for gas turbine first stage vane cooling

Table 3 - 8Gas turbine unit description

3.2.6. Steam cycle section

The major components of the steam cycle section include the heat recovery steam generator, the steam turbines (high, intermediate, and low pressure), condenser, the steam bleed for gas turbine cooling, the recycle water pump and heater, and the deaerator. As discussed above, a three-pressure reheat HRSG is adopted for this IGCC system. The major parameters of this HRSG are given in Table 3-9.

Table 3 - 9STAG product line steam turbine throttle and admission steam
conditions

Heat Recovery Steam Cycle	Reheat Three-Pressure
Throttle Pressure (psig)	1400
Throttle Temperature (°F)	1000
Reheat Pressure (psig)	300
Reheat Temperature (°F)	1000
IP Admission Pressure (psig)	300
IP Admission Temperature (°F)	20 Approach to Exhaust Gas Temperature upstream of Superheater
LP Admission Pressure (psig)	40
LP Admission Temperature (°F)	20 Approach to Exhaust Gas Temperature upstream of superheater

Steam cycle process

The three-pressure reheat steam cycle is shown schematically in Figure 3-10, and Table 3-10 gives the corresponding operation units in the Aspen Plus model. The feedwater coming from the steam turbine condenser is preheated up to 221.9 °F in the feed water preheater, which is simulated by the block FWHeat. The heat recovered from the sulfur recovery process/and steam removed from the low pressure turbine is used to preheat the feed water. The preheated feedwater enters the deaerator, which is simulated

by the block Dearer. After deaeration, the feed water is compressed up to 52 psia, and enters the low pressure economizer, which is simulated by the block LPEc. In the low pressure economizer, the feedwater is heated up to 253° F. Part of the low pressure feed water is used to generate the superheated low pressure steam at 42 psia/500 °F through the low pressure evaporator, which is modeled by the block LPEvap, and the low pressure superheater, which is modeled by the block LPSupH. Another amount of the feedwater is compressed up to 360.2 psig, and heated up to 408 °F in the intermediate pressure economizer, which is simulated by the block IPEc2. Part of the intermediate pressure feed water is used to generate the superheated intermediate pressure steam at 303 psia/581 °F through the intermediate pressure evaporator, which is modeled by the block IPEvap, and the intermediate pressure superheater, which is modeled by the block IPSupH. Another part of the intermediate feedwater is compressed to 1824.4 psia, and heated up to 585 °F in the high pressure economizer, which is molded by the block HPEc3. The high pressure hot water enters the high pressure evaporator, which is modeled by the block HPEvap, to generate the high pressure saturation steam.

Before entering the high pressure turbine, the high pressure saturation steam is heated up to 1000 °F in the high pressure superheater, which is simulated by the block HPSupH. Steam from the high pressure turbine at 336 psia/606 °F mixes with the intermediate pressure superheated steam, then is heated up to 1000 °F after flowing through the reheater, which is modeled by the block ReHeat. The reheated steam flows through the intermediate turbine, which is simulated by block IPTur1 and IPTur2. The steam from the intermediate turbine at 40 psia/501.9 °F mixes with the steam from the low pressure superheater, and then passes through the low pressure turbine, which is simulated by block LPTur1, and LPTur2. The steam from the low pressure turbine at 0.67 psia/93.5 °F is condensed at the condenser, which is modeled by the block Cond.



Figure 3 - 10 GE 7FA gas turbine and steam cycle section flowsheet

No	Aspen unit ID	Unit parameters	Unit function	
1	PmpMK (PUMP)	Discharge pressure=20 psia Efficiency=default value	Simulate the make up feed water pump	
2	CndPmp1 (Pump)	Discharge pressure=17 psia Efficiency=default value	Simulate the condensate water pump	
3	FWMix (Mixer)		Indicate the mixing of make up feed water and the condensate water	
4	FWHeat (Heater)	Pressure=17 psia Vapor fraction=0	Simulate the pre-heater of the feed water	
5	DearMix (Mixer)		Indicate the mixing of feed water and the low pressure hot water	
6	Deaer (Flash 2)	Pressure=16.3 psia Vapor fraction=0.005	Simulate the deaerator	
7	LPLoop (Fsplit)		Indicate a amount of feed water is split to the low pressure economizer	
8	LPPump (Pump)	Discharge pressure=52 psia Efficiency=default value	Simulate the low pressure feed water pump	
9	LPEc (Heater)	Pressure drop=4 psia Temperature=253 F	Simulate the low pressure economizer for low pressure steam generation	
10	SP_LPEc (Fsplit)		Indicate a amount of low pressure feed water is split to the deaerator	
11	LPEvap (Flash 2)	Pressure drop=4 psia Vapor fraction=0.99	Simulate the low pressure evaporator, where the blow down is 1% of the inlet water	
12	LPSupH (Heater)	Pressure drop=-2 psia Temperature=500 F	Simulate the low pressure steam superheater	
13	SP_Pmps (Fsplit)		Indicate a amount of feed water is split to the intermediate pressure economizer	
14	IPPmp (Pump)	Discharge pressure=360.2 psia Efficiency= default value	Simulate the intermediate pressure pump	
15	IPEc1 (Heater)	Pressure=342.1 psia Temperature=253 F	Simulate the first intermediate pressure economizer	
16	IPEc2 (Heater)	Pressure=325 psia Temperature=408 F	Simulate the second intermediate pressure economizer	
17	IPEvap (Flash 2)	Pressure=308.8psia Vapor fraction=0.99	Simulate the intermediate pressure evaporator, where the blow down is 1% of the inlet water	
18	IPSupH (Heater)	Pressure=303psia Temperature=581 F	Simulate the high pressure steam superheater	

Table 3 - 10Steam cycle section unit description

No	Aspen unit ID	Unit parameters	Unit function
19	HPPmp (Pump)	Discharge pressure=1824.4 psia Efficiency= default value	Simulate the high pressure pump
20	HPEc1 (Heater)	Pressure=1733.2 psia Temperature=253°F	Simulate the first high pressure economizer
21	HPEc2 (Heater)	Pressure=1646.5 psia Temperature=408°F	Simulate the second high pressure economizer
22	HPEc2 (Heater)	Pressure=1564.2 psia Temperature=585°F	Simulate the third high pressure economizer
23	HPEvap (Flash 2)	Pressure=1486psia Vapor fraction=0.99	Simulate the high pressure evaporator, where the blow down is 1% of the inlet water
24	IPSupH (Heater)	Pressure=1400 psia Temperature=1000°F	Simulate the high pressure steam superheater
25	HPTur (Compr)	Discharge pressure=336 psia Isentropic efficiency=0.92	Simulate the high pressure turbine
26	Re_Mix (Mixer)		Indicate the mixing of superheat intermediate pressure steam and the steam from high pressure turbine
27	ReHeat (heater)	Pressure=300 psia Temperature=1000°F	Simulate the steam reheater
28	IPTur1 (Compr)	Discharge pressure=60 psia Isentropic efficiency=0.92	Simulate the first stage of the intermediate pressure turbine
29	IPTur2 (Compr)	Discharge pressure=40 psia Isentropic efficiency=0.92	Simulate the second stage of the intermediate pressure turbine
30	LPMix (Mixer)		Indicate the mixing of low pressure superheat steam and the steam from the intermediate turbine
31	LPTur1 (Compr)	Discharge pressure=24 psia Isentropic efficiency=0.89	Simulate the first stage of the low pressure turbine
32	BleedLP (Fsplit)		Indicate a amount of low pressure steam is removed to the feed water pre- heater
33	LPTur2 (Compr)	Discharge pressure=0.67 psia Isentropic efficiency=0.89	Simulate the second stage of the low pressure turbine
34	Cond (Heater)	Pressure=0.6252 Vapor fraction=0	Simulate the condenser, where the steam from the last stage of the low pressure turbine is condensed

Table 3 – 10 continued

Design specifications of the steam cycle

• Gas turbine exhaust stack temperature

In general, the HRSG stack temperature should be kept as low as possible to extract as much gas turbine exhaust energy as possible to maximize cycle efficiency. Occasionally, a concern with high sulfur gas turbine fuels is acid condensation on low temperature heat transfer surfaces. In these cases, a low pressure turbine extraction may be used to heat feedwater above the acid dew point prior to feedwater supply to the HRSG economizer. In this study, the stack temperature of the gas turbine exhaust gases is set to be 230 °F. This temperature is maintained by the design specification TSTACK, which can adjust the feed water flow rate of the HRSG to meet the requirement of the stack temperature.

• Pinch, sub-cool, and approach temperature

As shown in Figure 3-11, the pinch temperature is the temperature difference between the gas turbine exhaust temperature and the temperature of saturation water at the inlet of evaporator. The approach temperature is the temperature difference between the main steam temperature and the GT exhaust temperature. The sub-cool temperature is the temperature difference between the water temperature at the outlet of the economizer and the temperature of the saturation water.



Heat Transfer

Figure 3 - 11 Typical exhaust gas temperature profile of one pressure system

If the pinch temperature and the approach temperature are too big, the gas turbine exhaust energy will not be utilized efficiently. On the other hand, if they are too small, heat transfer area will be very large, which will raise the capital cost. Generally, the pinch temperature rage is from 8 to 20 °C, and the approach temperature range from 5 to 20 °C. In order to avoid some of hot water in economizers evaporating, typically the sub-cool temperature range is from 5 to 20 °C. The temperature profile of this three-pressure reheat HRSG is shown by Figure 3-12.

The approach temperature and the sub-cool temperature can be satisfied by setting the main steam temperature and the outlet temperature at the outlet of an economizer. The pinch temperature is satisfied through a design specification TPINCH, which adjusts the feedwater flow rates entering the high, intermediate, and low pressure economizers.



Figure 3 - 12 Typical exhaust gas/steam cycle temperature profile for threepressure reheat HRSG system

3.2.7. Convergence sequence of the IGCC model

Using the sequential-modular (SM) strategy, Aspen Plus performs flowsheet calculations by executing each unit operation block in sequence, and using the calculated output streams of each block as feed to the next block. When flowsheets with recycle loops, design specifications, or optimization problems, it must be solved iteratively. In this study, the convergence sequence is based on eleven design specifications and seven calculators with FORTRAN blocks. Some of them are mentioned in earlier sections of this report and the rest are elaborated upon in this section.

The water to coal ratio is varied by the FORTRAN block H2OCOAL in order to meet the specified coal slurry composition. The elemental composition of coal decomposition is calculated by the FORTRAN block DECOM based on the ultimate composition analysis of the coal.

The FORTRAN block BSAIR maintains the air flow rate entering the combustion reactor of the SCOT process based on the stoichiometric calculation. The FORTRAN block CLAIR maintains the air flow rate entering the first reactor of the Claus process based on the stoichiometric calculation. The flow rate of the make up feedwater for the steam cycle is determined by the FORTRAN block STMAKUP, which takes into account the blowdown in the IGCC system.

3.3. IGCC Cost Model

The cost models for oxygen-blown Texaco quench IGCC systems are developed through updating a previous IGCC cost model developed by Frey [Frey, 1993]. The references used for updating the cost model are given by the following table.

Report No.	Company	Authors	Year	Sponsor	Gasifier
1. Evaluation of Innovative Fossil Fuel Power Plants with CO ₂ Removal	Parsons	W. Owens,	2000	DOE/EPRI	E-gas
2. Texaco Gasifier IGCC Base Cases	EG&G	W. Shelton J. Lyons	2000	NETL	Texaco
3. KRW Gasifier IGCC Base Cases	EG&G	W. Shelton J. Lyons	2000	NETL	KRW
4. Shell Gasifier IGCC Base Cases	EG&G	W. Shelton J. Lyons	2000	NETL	Shell
5. A single IGCC design for variable CO ₂ capture	GE/Texac o	O'Keefe L.F. Griffiths J.	2002		Texaco
6. Market-Based Advanced Coal Power Systems	DOE		1999	DOE	Destec
7. Shift reactors and physical absorption for Low-CO ₂ emission IGCCs		P. Chiesa S. Consonni	1999		Texaco

 Table 3 - 11
 References used for updating the IGCC cost model

For the purpose of estimating the direct capital cost of the plant, the IGCC system is divided into thirteen process areas as listed in the following table. The following section gives the direct cost of each process area in a dollar value at 2000 year.

No.	Cost section
1	Coal handling:
2	Oxidant feed
3	Gasification
4	LTGC
5	Selexol
6	Claus plant
7	Beavon-Stretford
8	Boiler feedwater treatment
9	Process condensate treatment
10	Gas turbine
11	HRSG
12	Steam turbine
13	General facilities

Table 3 - 12IGCC system process areas

3.3.1. Oxidant Feed Section

This process section typically has an air compression system, an air separation unit, and an oxygen compression system. The direct cost depends mostly on the oxygen feed rate to the gasifier, as the size and cost of compressors and the air separation systems are proportional to this flow rate. The direct cost model for the oxidant feed section is:

$$DC_{OF} = 196.2 \frac{N_{T,OF} T_{a}^{0.067}}{1 - \eta_{ox}^{0.073}} (\frac{M_{O,G,i}}{N_{O,OF}})^{0.5618}$$
(R²=0.86)....(3-1)

where $N_{T,OF}$, $N_{O,OF}$ = the total trains of ASU and the total operating trains of ASU, separately.

$$T_a({}^0F) =$$
 Ambient air temperature; $20 \le T_a \le 95$; 0F
 $M_{O,G,i} (lbmole/hr) =$ Gasifier oxygen inlet flow rate;
 $625 \le \frac{M_{O,G,i}}{N_{O,OF}} \le 17000$ lbmole/hr

$$\eta_{ox}$$
 = oxygen purity; $0.95 \le \eta_{ox} \le 0.98$

This regression is based on the equation developed by Frey [Frey, 1993], and revised using data from reports [3], [5-7]. Figure 3-13 gives the data points used for this regression.



Figure 3 - 13 Oxygen flow rate vs. oxidant feed section cost

3.3.2. Coal Handling Section and Slurry Preparation

Coal handling involves unloading coal from a train, storing the coal, moving the coal to the grinding mills, and feeding the gasifier with positive displacement pumps. Slurry preparation trains consist of vibrating feeders, conveyors, belt scale, rod mills, storage tanks, and positive displacement pumps to feed the gasifier. Coal feed rate to gasifier on as-received basis is the most common and easily available independent variable. The direct cost model for the coal handling is based on the overall flow to the plant rather than on per train basis.

$$DC_{CH} = 8.27M_{CF,G,i}$$
 (R²=0.8) (3-2)

where $M_{CF,G,i}(tons/day)$: Gasifier as-received coal feed flow rate; 2,800~25,000 tons/day.

This regression is based on the equation developed by Frey [Frey, 1993], and revised using data from reports [3-5], and [7]. Figure 3-14 gives the data points for coal handling section cost analysis.



Figure 3 - 14 Coal handling section cost vs. coal feed flow rate

3.3.3. Gasification Section

The Texaco quench gasification section of an IGCC plant contains gasifier, gas scrubbing, gas cooling, slag handling, and ash handling sections. The direct capital cost model is a function of the as-received coal flow rate.

$$DC_{G} = 24770N_{T,G} \ln(M_{CG,G,i} / N_{O,G}) - 167453$$
(3-3)

where $N_{T,G}$, $N_{O,G}$ = the total trains of gasifier and the total operating trains of gasifier, separately.

 $M_{CF,G,i}(tons/day) = Gasifier as-received coal feed flow rate; 1,300~3,300 tons/day.$

This regression is based on data from reports [1] and [3]. The data points are given in Figure 3-15.



Figure 3 - 15 Coal flow rate vs. gasifier cost

3.3.4. Low temperature gas cooling

The low temperature gas cooling section consists primarily of a series of shell and tube heat exchangers. The syngas mass flow is assumed to be the major determinant of the process area capital cost as in the original cost model.

$$DC_{LT} = 0.0519N_{T,LT} \left(\frac{M_{syn,LT,i}}{N_{O,LT}}\right)^{0.9}$$
(R²=0.92) (3-4)

where $N_{T,LT}$, $N_{O,LT}$ = the total trains and the total operating trains of low temperature gas cooling, separately.

 $M_{syn,LT,i}(lb/hr) =$ syngas inlet flow rate of low temperature gas cooling section, $650,000 \le \left(\frac{M_{syn,LT,i}}{N_{0,LT}}\right) \le 1,300,000$

This regression is based on the data from reports [3-5]. The data points are shown in Figure 3-16.





3.3.5. Selexol Section

Hydrogen sulfide in the syngas is removed through counter-current contact with the Selexol solvent. The cost of the Selexol section include the acid gas absorber, syngas knock-out drum, syngas heat exchanger, flash drum, lean solvent cooler, regenerator air-cooled overhead condenser, acid gas knock-out drum, regenerator reboiler, and pumps and expanders associated with the Selexol process. The direct capital cost model for the Selexol section is:

$$DC_{s} = \frac{0.304N_{T,S}}{(1 - \eta_{s})^{0.059}} (\frac{M_{syn,S,i}}{N_{O,S}})^{0.98}$$
(R²=0.94) (3-5)

where $N_{T,S}N_{O,S}$ = the total trains and the total operating trains of Selexol section for H₂S capture, separately.

 η_s (%) = the H₂S capture efficiency of Selexol system, 83.5%~99.7%.

 $M_{syn,S,i}$ (lbmol/hr) = syngas inlet flow rate of Selexol section. 2,000~67,300 lbmol/hr

This regression is based on the equation developed by Frey [Frey, 1993], and adjusted with the chemical engineering price index.

3.3.6. Claus sulfur recovery Section

The Claus plant contains a two-stage sulfur furnace, sulfur condensers, and catalysts. It cost is estimated as a function of the element sulfur outlet flow rate of the Claus unit.

$$DC_{c} = 6.96N_{T,C} \left(\frac{M_{s,C,o}}{N_{o,C}}\right)^{0.668}$$
(R²=0.97) (3-6)

where $N_{T,C}$, $N_{O,C}$ = the total trains and the total operating trains of Claus section for sulfur capture, separately.

 $M_{s,c,o}(lb/hr)$ = the element sulfur outlet flow rate of the Claus, 695~18,100.

This regression is based on the equation developed by Frey [Frey, 1993], and adjusted with the chemical engineering price index.

3.3.7. Beavon-Stretford Tail Gas Removal Section

The capital cost of a Beavon-Stretford unit is expected to vary with the volume flow rate of the input gas stream and with the mass flow rate of the sulfur produced.

$$DC_{BS} = 63.3 + 72.8N_{T,BS} \left(\frac{M_{S,BS,o}}{N_{O,BS}}\right)^{0.645}$$
(R²=0.99) (3-7)

where $N_{T,BS}$, $N_{O,BS}$ = the total trains and the total operating trains of Beavon-Stretford for sulfur capture, separately.

 $M_{S,BS,o}(lb/hr)$ = the element sulfur outlet flow rate of Beavon-Stretford, 75~1,200.

This regression is based on the equation developed by Frey [Frey, 1993], and adjusted with the chemical engineering price index.

3.3.8. Boiler Feedwater System

The boiler feedwater system consists of equipment for handling raw water and polished water in the steam cycle, including a water mineralization unit for raw water, a dimineralized water storage tank, a condensate surge tank for storage of both dimineralized raw water and steam turbine condensate water, a condensate polishing unit, and a blowdown flash drum. The cost model considers both raw water flow rate through the demineralization unit and the polished water flow rate through the polishing unit. The polished water includes steam turbine condensate and makeup water, and condensate from the miscellaneous process users such as waste water treatment.

$$DC_{BFW} = 0.16M_{rw}^{0.307}M_{pw}^{0.435}$$
(R²=0.99) (3-8)

where M_{rw} (lb/hr) = the flow rate of raw water, 24,000~614,000.

 M_{pw} (lb/hr) = the flow rate of polished water in the steam cycle, 234,000~3,880,000

This regression is based on the equation developed by Frey [Frey, 1993], and adjusted with the chemical engineering price index.

3.3.9. Process Condensate Treatment

The process condensate treatment area consists of strippers, air cooled heat exchangers, and knock-out drums. It is expected that the process condensate treatment direct cost will depend primarily on the scrubber blowdown flow rate.

$$DC_{PC} = 10670 \left(\frac{M_{SBD}}{300000}\right)^{0.6}$$
(3-9)

where $M_{SBD}(lb/hr) =$ the blowdown flow rate.

This regression is based on the equation developed by Frey [Frey, 1993], and adjusted with the chemical engineering price index.

3.3.10. Gas Turbine Section

A number of design factors affect the cost of a gas turbine in an IGCC system. In this study, the cost model for the gas turbine was developed for a GE Frame 7F gas turbine.

$$DC_{GT} = 168MW_e$$
 (R²=0.92) (3-10)

where, MW_e = the net output of GE7F gas turbine (MW).

This regression is based on the data from reports [3] and [7]. The data points used for regression are given in Figure 3-17.



Figure 3 - 17 Gas turbine cost vs. gas turbine net output

3.3.11. Heat Recovery Steam Generator

The HRSG is a set of heat exchangers in which heat is removed from the gas turbine exhaust gas to generate steam, including the superheater, reheater, high pressure steam drum, high pressure evaporator, and the economizers. The direct cost of the HRSG is a simple regression model based on the high pressure steam flow rate to the steam turbine.

$$DC_{LT} = -5943 + 7.98 \times 10^{-3} N_{T,HR} P_{hps,HR,0}^{1.526} \left(\frac{M_{hps,HR,0}}{N_{O,HR}}\right)^{0.242}$$
(R²=0.96) (3-11)

where $N_{T,HR}$, $N_{O,HR}$ = the total trains and the total operating trains of HRST, separately.

 $P_{hns HR o}(lb/hr)$ = the high pressure steam mass flow rate of HRSG,

$$66,000 \le (\frac{M_{\text{hps,HR,o}}}{N_{\text{O,HR}}}) \le 64,0000$$

This regression is based on the equation developed by Frey [Frey, 1993], and adjusted with the chemical engineering price index.

3.3.12. Steam Turbine

A typical steam turbine consists of the high-pressure, intermediate-pressure, and low-pressure turbine stages, a generator, and an exhaust steam condenser. The cost model is given by

$$DC_{GT} = 0.145W_e$$
 (R²=0.92) (3-12)

where W_e = the net output of gas turbine (kW), 200,000~500,000

This regression is based on the data from reports [6-7], which are shown in Figure 3-18.



Figure 3 - 18 Steam turbine cost vs. steam turbine net power output
3.3.13. General Facilities

The general facility section includes cooling water system, plant and instrument air, potable and utility water, and electrical system. Most studies assume that the direct cost of the general facilities is approximately 14%-17% of the direct costs of other sections. In the present study the direct cost of the general facilities is assumed to be approximately 15% of the total direct cost of the above 12 sections. Based on the direct cost of each section, the process facility cost of each section is estimated as 1.2 times of its direct cost [Frey, 1993].

3.3.14. Total Capital Requirement of IGCC systems

The following cost and parameter estimation of IGCC systems follows the principles given by the EPRI Technical Assessment Guide (1993), which is widely considered the industry standard and has long been an authoritative source of cost and performance information on advanced and conventional power generation, storage, transmission and distribution.

The total process facilities cost (PFC) of the IGCC system is the summation of the individual process facility costs. Based on the PFC, the engineering and home office costs can be estimated. The engineering and home office costs include the costs associated with: (1) engineering, design, and procurement labor; (2) office expenses; (3) licensing costs for basic process engineering; (4) office burdens, benefits, and overhead costs; (5) fees or profit to the architect/engineer. EPRI recommends that a value of 7 to 15 percent of the process facility cost as the engineering and home office cost. Therefore, a value of 10 percent is used here as a default.

Project contingency costs reflect the expected increase in the capital cost estimate that would result from a more detailed cost estimate for a specific site. Usually, project contingency is assigned as a multiplier of the process facility cost. A typical value for the project contingency for a preliminary level cost estimate, as defined by the EPRI Technical Assessment Guide, is 20 percent.

Another major cost item is the process contingency. The process contingency is used in deterministic cost estimates to quantify the expected increase in the capital cost of an advanced technology due to uncertainty in performance and cost for the specific design application. In the EPRI cost method, the process contingency is estimated based on separate consideration of contingencies for each process section. The contingency is expressed as a multiplier of the sum of the plant facility cost for each process area. The process contingency decreases as the commercial experience with a process area increases. For example, in a fully commercialized process that has been used in similar applications, the process contingency may be zero. The ranges of process contingency factors for IGCC systems are shown in Table 3-13.

Cost section	Process contingency
Coal handling:	0.05
Oxidant feed	0.05
Gasification	0.15
LTGC	0
Selexol	0.1
Claus plant	0.05
Beavon-Stretford	0.1
Boiler feedwater treatment	0
Process condensate treatment	0.3
Gas turbine	0.125
HRSG	0.025
Steam turbine	0.025
General facilities	0.05

 Table 3 - 13
 Process contingency of cost sections

The total plant cost (TPC) is the sum of process facility cost, the engineering cost, the process contingency, and the project contingency. An allowance for funds during construction (AFDC) is calculated based on the TPC as a function of the amount of time it would take to construct the plant. Methods for computing the AFDC are documented elsewhere [EPRI, 1993] and are not repeated here. The total plant investment (TPI) represents the sum of the total plant cost and the AFDC.

The final measure of the capital cost is the total capital requirement (TCR). The TCR includes the total plant investment plus owner costs for royalties, startup costs, and initial inventories of stock feed. Preproduction costs typically include one month of both fixed and variable operating costs and two percent of total plant investment. Inventory capital is estimated as 0.5 percent of total process capital excluding catalyst. The initial catalyst cost requirement is estimated based on the unit price of the catalysts and their

volumes. The total capital cost and O&M cost calculation processes are given in Table 3-14.

Capital cost elements	Value
Total process facilities cost	Sum of the PFC of each section
Engineering and home office	10% PFC
General facilities	15% PFC
Project contingency	20% PFC
Process contingency	See Table
Total plant cost (TPC) = PFC+Engineering fee-	-General facilities+Project & Process
Allowance for funds during construction (AFDC)	Calculated based on discount rate and construction time
Royalty fees	0.5% PFC
Preproduction fees	1 moth fee of VOM&FOM
Inventory cost	0.5% TPC
Total capital requirement (TCR) = TPC+AFDC	+Royalty fees+Preproduction fee+Inventroy
Fixed O&M cost (FOM)	
Total maintenance cost	2% TPC
Maintenance cost allocated to labor	40% of total maintenance cost
Administration & support labor cost	30% of total labor cost
Operation labor	\$25/hour
Variable O&M cost (VOM)	
Fuel cost	Depends on coal type
Consumable	See Table 3-15

Table 3 - 14Capital cost elements of an IGCC power plant

The fixed operation and maintenance costs, including labor, administration and support cost are estimated as a fraction of the total plant cost. The variable cost and expenses associated with operating the plant include: the consumable and fuel cost. The unit costs of consumable are given by Table 3-15.

Material	Unit cost	Unit	
Sulfuric acid	119.52	\$/ton	
NaOH	239.04	\$/ton	
Na2 HPO4	0.76	\$/lb	
Hydrazine	3.48	\$/lb	
Morpholine	1.41	\$/lb	
Lime	86.92	\$/ton	
Soda ash	173.85	\$/ton	
Corrosion Inh	2.06	\$/lb	
Surfactant	1.36	\$/lb	
Chlorine	271.64	\$/ton	
Biocide	3.91	\$/lb	
Selexol Solv.	1.96	\$/lb	
Claus catalyst	478.08	\$/ton	
B/S catalyst	184.71	\$/ft^3	
Fuel oil	45.64	\$/bbl	
Plant air ads.	3.04	\$/lb	
Water	0.79	\$/Kgal	

 Table 3 - 15
 Unit costs of consumables (Source: IECM manual)

REFERENCES (CHAPTER 3)

- IEA Greenhouse Gas R&D Program, 2003: Potential for improvement in gasification combined cycle power generation with CO₂ capture, IEA report, report number PH4/19
- 2. Brdar R.D., Jones R.M., 2003: GE IGCC Technology and Experience with Advanced Gas Turbines, GE Power Systems, GER-4207
- 3. Chase D.L., Kehoe P.T., 2003: GE Combined-Cycle Product Line and Performance, GE Power Systems, GER-3574G
- 4. IEA, 2000: Modeling and simulation for coal gasification, IEA Coal Research 2000, ISBN 92-9029-354-3
- 5. Foster A.D., Doering H.E., and Hilt M.B., 2003: Fuel flexibility in heavy-duty gas turbines, GE Company, Schenectady, New York
- IEA Greenhouse Gas R&D Program, 2003: Potential for improvement in gasification combined cycle power generation with CO₂ capture, IEA report, report number PH4/19
- 7. Brdar R.D., Jones R.M., 2003: GE IGCC Technology and Experience with Advanced Gas Turbines, GE Power Systems, GER-4207
- 8. Chase D.L., Kehoe P.T., 2003: GE Combined-Cycle Product Line and Performance, GE Power Systems, GER-3574G
- 9. IEA, 2000: Modeling and simulation for coal gasification, IEA Coal Research 2000, ISBN 92-9029-354-3
- 10. Foster A.D., Doering H.E., and Hilt M.B., 2003: Fuel flexibility in heavy-duty gas turbines, GE Company, Schenectady, New York
- Frey H. C., 2001: Probabilistic modeling and evaluation of the performance, emissions, and cost of Texaco gasifier-based integrated gasification combined cycle systems using ASPEN, Janu. 2001
- 12. Owens W., 2000: Evaluation of Innovative Fossil Fuel Power Plants with CO₂ Removal, 2000
- 13. Shelton W., Lyons J., 2000: Texaco Gasifier IGCC Base Cases, PED-IGCC-98-001, last revision June 2000
- 14. Shelton W., Lyons J., 2000: NETL, KRW Gasifier IGCC Base Cases, PED-IGCC-98-005, last revision June 2000

- 15. Shelton W., Lyons J., 2000: Shell Gasifier IGCC Base Cases, PED-IGCC-98-005, last revision June 2000
- 16. O'Keefe L.F. Griffiths J., 2002: A single IGCC design for variable CO₂ capture, Fifth European Gasification Conference, April, 2002
- 17. Office of Fossil Energy, U.S. Department of Energy, 1999: Market-based advanced coal power systems, Final report, May 1999
- Chiesa P., Consonni S., 1999: Shift reactors and physical absorption for Low-CO₂ emission IGCCs, Journal of engineering for gas turbines and power, 121 (2): 295-305 APR 1999
- 19. Zhu, Yunhua, 2004: Evaluation of Gas Turbine and Gasifier-Based Power Generation Systems, PhD Dissertation, North Carolina State University August 2004.

Chapter 4. PERFORMANCE AND COST MODEL OF WATER GAS SHIFT REACTION SYSTEM

4.1. Introduction

The water-gas shift (WGS) reaction is an industrially important reaction, which is also a key part of the CO₂ capture system in an IGCC power plant, for it converts almost all the CO in syngas into CO₂ for CO₂ capture before combustion. Without this conversion via the water gas shift reaction, the pre-combustion CO₂ capture from IGCC would not be an attractive option due to the low CO₂ partial pressure of CO₂ in the raw syngas. The reaction is given as follows [David, 1980].

$$CO + H_2 0 \leftrightarrow CO_2 + H_2$$
 (4-1)



Figure 4 - 1 Effects of temperature and CO/steam on the CO conversion of the WGS reaction (This figure is derived based on that the original molar concentration ratios of CO to H₂O are 1.5, 2, 2.5, and 3, and the original concentrations of CO₂ and H₂ equal zero)

This reaction is exothermic. Figure 4-1 shows the effect of the reaction temperature of the water gas shift reaction on the equilibrium conversion of CO. Equilibrium will favor CO conversion to CO_2 at low temperatures. The equilibrium will also favor CO conversion at high steam-to-CO ratios. The steam-to-CO ratio is determined by the

chemical process. For IGCC systems with CO_2 capture, the steam required is supplemented by existing upstream steam or water quench addition.

4.2. Effects of operation temperature and two-stage shift reaction system

Practically, the water gas shift reaction occurs in an adiabatic system with the presence of a catalyst accelerating the reaction rate. In an adiabatic system, the CO slip is determined by the exit temperature of the shift reactors, because low temperatures result in low equilibrium levels of CO. On the other hand, favorable kinetics occurs at higher temperatures. Either high steam-to-gas ratio or low temperature can improve CO conversion percentage, but it also requires higher capital and operation cost. Hence, there is a tradeoff between CO conversion percentage and costs.

Conversion in a single reactor is equilibrium limited. As the reaction proceeds, the rise in temperature due to the exothermal reaction eventually restricts further reaction. This limitation can be overcome with a two stage water gas shift reaction--a high temperature shift reactor followed by a low temperature shift reactor. An inter-bed cooling process is employed between the two reactors to keep the reaction occurring at low temperature in the second reactor. Attainment of low equilibrium CO slip from the low temperature reactor is critical to the efficient and economic operation of plants. A typical CO variation in a two stage shift reactors is shown in Figure 4-2.



Figure 4 - 2 Typical CO variation in high temperature shift and low temperature shift catalyst beds [Frank, 2003a]

4.3. Clean shift catalysts

Gases used in water gas shift reactors often contain sulfur component, such as H_2S and COS. These sulfur components have a detrimental effect on the activation of some shift catalysts, which will be poisoned and lose activation in the presence of sulfur components. On the other hand, sulfur components are necessary to maintain the activation of some other shift catalysts. For the former type of shift catalysts, sulfur components must be removed from reaction gases before the water gas shift reaction. Hence this type of catalysts is so-called "clean shift catalyst". A schematic flowsheet of coal gasification system with a clean water gas shift reaction is given in Figure 4-3. The syngas from the gasifier is cooled down, and fed to the soot scrubber to remove the bulk of the carbon. Then it is further cooled for sulfur removal. Before passing to the shift reactors, steam is added to the syngas to meet requirements of steam-to-carbon ration. The inlet temperature of the second stage of the shift reaction is controlled by the feed/effluent heat exchanger.



Figure 4 - 3 Coal gasification system with a clean water gas shift reaction

As mentioned above, a low operating temperature will give the most favorable thermodynamic equilibrium and hence the minimum slip of carbon monoxide. For a twostage shift reaction system, the ideal operation is the low temperature shift reactor with the lowest possible inlet temperature. There are two boundaries which limit the operation temperature of the low temperature shift reactor. One is the activity of the catalyst at lower temperature; the other is the dew-point of the process gases because condensation on shift catalyst will weaken the clean catalyst pellets at low temperatures [Frank, 2003].

It has been reported that a low pressure plant was able to operate the low temperature shift reactor at an inlet temperature of only 340 °F, because of the low dewpoint of the process gas [Frank, 2003a]. The dew-point of the process gas increases with the increase of pressure. Hence, for high pressure cases it is the dew point, not the activity of the catalyst, is more likely to be an operating limitation. IGCC systems usually operate at a pressure high enough to allow the low temperature shift reaction to be operated close to the dew point of the process gas. A safety margin above the dew point should be used to ensure complete evaporation of water droplets that may form in the cooler. This is adopted as a design criterion for a water gas shift reaction in an IGCC system with a clean shift reaction.

For a two-stage shift reaction with clean shift catalysts, the iron-based catalyst is the common commercially available high temperature catalyst. The commonly used low temperature clean shift catalysts are copper-based. Both high temperature and low temperature catalysts require activation by in situ pre-reduction steps. Since both catalysts burn up when exposed to air (pyrophoric), they must be sequestered during system shutdown when only air flows through the system [Frank 2003a].

The lifetimes of Cu-based catalysts and Fe-based catalysts are determined by the poison-absorbing capacity of the catalysts. These poisons are inevitably present in the process gas, such as syngas from coal gasification, or introduced with steam. As mentioned above, the key poison in syngas is sulfur. Hence a sulfur removal process is required upstream of the water gas shift reaction.

4.4. Sulfur tolerance shift catalysts

The so-called sour shift catalysts are sulfur tolerant, and sulfur is required in the feed gas to maintain the catalyst in the active sulphided state. This type of catalyst is usually cobalt-based.

Figure 4-4 shows the schematic process of a gasifier system with a sour shift reaction. The syngas from the gasifier is quenched, and then the saturated syngas is fed to the soot scrubber, to remove the bulk of particles before passing to the sour shift reactors. The inlet temperature of the second stage of the shift reaction is controlled by the cooling process. After heat recovery, the cooled syngas from the second shift reactor is passed to the sulfur removal system.



Figure 4 - 4 Schematic process of a gasifier system with a sour shift

The sour shift catalyst has demonstrated its high and low temperature performance, ranging from 210°C to 480°C, and work properly up to a pressure as high as 1160 psia [Frank, 2003b]. Because the catalyst is not impregnated with a water-soluble promoter it can be operated closer to the dew point and will not lose activity when wetted occasionally.

In a gasification plant, the average catalyst life in the first stage shift reactor was 2.5 years, and 5-8 years in the second reactor [Frank, 2003b]. The difference in catalyst life in the two reactors is highly influenced by the gas quality. These data of catalysts' lifetime are adopted for the estimation of the operation and maintenance cost of the water gas shift reaction system.

4.5. Performance model of the water-gas shift reaction process

This section presents the performance model developed for the WGS reaction process. This is a general performance model for a two-stage shift system with either clean shift catalysts or sulfur tolerant shift catalysts. The purpose of the performance model is to characterize the change in syngas composition and flow rate as a function of inlet condition to the WGS reactor and key design parameters of the WGS system. The performance model also characterizes the heat integration between the shift reaction system and the steam cycle system. A general water gas shift reaction process model is illustrated in Figure 4-5. The black box in this figure includes a high temperature reactor, a low temperature reactor and several heat exchangers for heat recovery. The performance of the shift reaction was first modeled in the Aspen Plus. In this model, the syngas from a gasifier is mixed with steam or quenched at a given temperature and pressure, and then fed into the high temperature reactor. Most of the CO in the syngas is converted to CO_2 in the high temperature reactor at a fast reaction rate. Because the water gas shift reaction is exothermic, the syngas from the high temperature reactor. Further CO conversion is achieved in the low temperature reactor. The shifted syngas from the low temperature reactor is cooled down again for subsequent CO_2 capture in a Selexol process, and the other part of the heat is integrated into the steam cycle.

In this model, the reactions in the two reactors are assumed to achieve equilibrium states. On the other hand, the shift reaction in a real reactor only approaches an equilibrium state. In order to compensate for the difference between the equilibrium state assumption and the real state in a reactor, the approach temperature method is used to adjust the model equilibrium temperatures. The difference between the model temperature and the design reaction temperature is referred to as the approach temperature. The approach temperature is determined through comparing model outputs with practical data from shift reactors in the industry field. Thus, with the approach temperature, the reactor model is assumed to reach an equilibrium state at a higher temperature than the design temperature, which makes the CO conversion efficiency in the model to match the realistic situation.



Figure 4 - 5 Mass and energy flow of the water gas shift reaction system

The Aspen model had been executed thousands of times with varying the inlet temperature, pressure and syngas composition. The value ranges of these parameters are given in Table 4-1 which covers the possible ranges of gasification operation. The inlet temperature was varied in a step of 30 F, and the inlet pressure was varied by a step of 100 psia. At the same time, 50 different syngas compositions were used. A total of 9000 cases were run. Based on the Aspen simulation results, statistical regression methods were then used to develop relationships between the inlet conditions and the final products of the WGS reaction. Using these regression relationships, the entire water gas shift reaction system can be treated as a "black box" when it is used in the IECM framework.

Parameter	Inlet temperature (F)	Inlet pressure (psia)	CO in the syngas (vol%)	H ₂ in the syngas (vol%)	CO ₂ in the syngas (vol%)	H ₂ O in the syngas (vol%)	CH ₄ in the syngas (vol%)
Value	440-755	150-1500	20-60	15-55	5-30	5-30	0.5-20

Table 4 - 1Range of model parameter values for the WGS reaction system

4.5.1. Input and output parameters of the WGS performance model

The input and outlet parameters of this model include the temperature, pressure, and flow rates of the inlet and the outlet syngas as shown in Table 4-2. The input parameters are used to calculate reaction rates and the composition changes after the reaction.

Table 4 - 2Input and output parameters of the WGS reaction system

Input parameter		Output parameter		
Syngas Temperature (F)		Shifted	Temperature (F)	
from gasifier Pressure (psia)	syngas	Pressure (psia)		
gasinei	Flow rate (lb-mole/hr)		Flow rate (lb-mol/hr)	
	Molar concentrations of		Molar concentration	
	CO, CO ₂ , H ₂ O, H ₂ , N ₂ , CH ₄		CO, CO ₂ , H ₂ O, H ₂ , N ₂ , CH ₄	
Steam/carb	oon molar ratio	Reaction rate	e & Catalyst volume (ft ³)	
Feed	Pressure (psia)	HP & IP	Temperature (F)	
water	Temperature (F)	steam	Flow rate (lb-mol/hr)	

4.5.2. Performance model output

This section discusses the performance outputs of this model. In this section, the CO to CO_2 conversion is defined and calculated using the chemical equilibrium constant. The outlet temperatures and syngas composition of the two shift reactors are regressed from Aspen model simulation results. The heat released from the syngas cooling is also quantified for the energy balance calculation of the whole IGCC system. The detailed calculation processes of CO conversion efficiency and catalyst volumes are given in

Appendix A and B. Appendix C shows the practical utilization of this model through a case study.

Shifted syngas composition

The water gas shift reaction occurring in both the high and low temperature reactors changes the concentration of syngas species and the temperature of the syngas. The CO conversion efficiency (ξ) can be used to show how much CO is converted into CO₂ in one reactor or in two reactors.

$$\xi = \frac{\text{CO flowrate in(lb \cdot mol/hr)} - \text{CO flowrate out(lb \cdot mol/hr)}}{\text{CO flowrate in(lb \cdot mol/hr)}}$$
(4-2)

A numerical model is set up to calculate the CO conversion in a shift reactor for given inlet parameters. The detailed calculation process is given in Appendix A.

Based on the definition of the CO conversion and stoichiometric factors of the reaction, the CO concentration of syngas exiting the high temperature reactor is given by,

$$[CO]_{h,o} = [CO]_0 \cdot (1 - \xi_h)$$
(4-3)

where $[CO]_{h,o}$ = the molar concentration of CO in the syngas exiting the high temperature reactor

 $[CO]_0$ = the molar concentration of CO in the syngas entering the high temperature reactor

 ξ_h = the CO conversion in the high temperature reactor

Based on the shift reaction (Eq. 4-1) and the definition of CO conversion, the molar concentrations of H_2 , CO_2 and H_2O after the high temperature reactor are given by,

$$[CO_2]_{h,o} = [CO_2]_0 + [CO]_0 \cdot \xi_h$$
(4-4)

$$[H_2]_{h,o} = [H_2]_0 + [CO]_0 \cdot \xi_h$$
(4-5)

$$[H_2O]_{h,o} = [H_2O]_0 - [CO]_0 \cdot \xi_h$$
(4-6)

Using the CO conversions definition and Equation (4-3), the CO concentration of shifted syngas after the low temperature reactor is to be given by,

$$[CO]_{l,o} = [CO]_0 \cdot (1 - \xi_{tot})$$
(4-7)

where $[CO]_{l,o}$ = the molar concentration of CO in the syngas exiting the low temperature reactor

ξ_{tot} = the total CO conversion in the high and low temperature reactors

Then the concentrations of H_2 , CO_2 and H_2O after the low temperature reactor are given by,

$$[H_2]_{1,0} = [H_2]_0 + [CO]_0 \cdot \xi_{tot}$$
(4-8)

$$[CO_2]_{l,o} = [CO_2]_0 + [CO]_0 \cdot \xi_{tot}$$
(4-9)

$$[H_2O]_{l,o} = [H_2O]_0 - [CO]_0 \cdot \xi_{tot}$$
(4-10)

Flow rate of high pressure saturation steam

In the following two sections, temperature changes and flow rates of water and syngas are calculated, and then used for the following cost model.

Syngas from the high temperature reactor is cooled down to a temperature which is determined by the dew point of syngas before it is fed into the low temperature reactor. According to the heat integration design, heat from the exothermic reaction is recovered to generate high pressure saturated steam for the steam cycle.

The temperature of the saturation steam is determined by the high pressure steam cycle in the power block. Using the data from the ASME steam and water table (1967), the temperature is given by the following regression equation:

$$T_{w,sat}(F) = 328.34 + 0.3565P_{sc} - 0.0002P_{sc}^2 + 6 \cdot 10^{-8}P_{sc}^3 - 7 \cdot 10^{-12}P_{sc}^4$$

$$(R^2=0.99)$$
 (4-11)

where P_{sc} (psia) = the pressure of steam cycle, (300 ~ 3000 psia)

The heat released by the syngas after the high temperature reactor is determined by,

$$Q_{\text{HE1}}(\text{Btu}/\text{hr}) = q_{\text{HE1}} \cdot f_{\text{SG},0}$$
(4-12)

where $f_{SG,0}$ = the total molar flow rate of syngas entering the high temperature reactor (lb-mole/hr);

 q_{HE1} = the heat released per lb-mole syngas after the high temperature reactor, which is regressed and given by (Btu/lb-mole),

$$q_{HE1}(Btu / lbmol) = P_0^{0.0360} T_0^{1.2874} [CO]_0^{1.14347} [CO_2]_0^{-0.4734} [H_2O]_0^{0.3150} [H_2]_0^{0.0003} [N_2]_0^{0.0139}$$
(R²=0.95) (4-13)

where P_0 = the pressure of syngas entering the high temperature reactor (psia)

 T_0 = the temperature of syngas entering the high temperature reactor (F)

 $[i]_0$ = the molar concentration of species i entering the high temperature reactor

Based on the total heat available and the saturation temperature, the flow rate of the saturation high pressure steam (f_{HPS} , lb-mole/hr) can be calculated by the following equation,

$$f_{HPS} = \frac{Q_{HE1}}{(h_{T_{w,sat}} - h_{T_0})}$$
(4-14)

where $h_{T_{w,sat}}$ = the enthalpy of the steam at the saturated temperature (Btu/lb-mole)

 h_{T_0} = the enthalpy of high pressure feed water at the inlet temperature (Btu/lb-mole).

Flow rate and temperature of the intermediate pressure steam

The syngas from the low temperature reactor is cooled to 100 F for sulfur removal, and the heat is recovered to generate the intermediate pressure steam. The total heat Q_{tot} (Btu/hr) released when the syngas from the low temperature reactor is cooled down to 100 °F is given by,

$$\begin{aligned} Q_{tot} = f_{l,o} (9.255 \cdot T_{l,o} - 0.316 \cdot P_{l,o} - 13861 \cdot [CO]_{l,o} - 297.779 \cdot [CO_2]_{l,o} - 1485.34 \cdot [H_2]_{l,o} \\ + 1759587 \cdot [H_2O]_{l,o} - 1439.29 \cdot [N_2]_{l,o} - 331.533 \cdot [CH_4]_{l,o}) \end{aligned}$$

 $(R^2=0.95)$ (4-15)

where $f_{l,o}$ = the molar flow rate of syngas exiting the low temperature reactor (lb-mole/hr);

 $T_{l,o}$ = the syngas temperature at the outlet of the second reactor

 $P_{l,o}$ = the syngas pressure at the outlet of the second reactor

 $[i]_{l,o}$ = the molar concentration of species i at the outlet of the second reactor

In order to meet the approach temperature requirement in the superheater, the final temperature of the intermediate pressure steam (T_{HPS}) is set to be 10 F lower than the outlet temperature of the syngas from the second shift reactor, and the feedwater temperature is set to be 59 F. Hence the flow rate of the intermediate pressure steam $(f_{IPS}, \text{lb-mole/hr})$ is given by,

$$(f_{HPS} + f_{IPS}) \cdot (h_{IP_{sat}} - h_{FW}) + f_{IPS} \cdot (h_{IPS} - h_{IP_{sat}}) = Q_{tot}$$
(4-16)

where f_{HPS} = the flow rate of the high pressure saturation steam (lb-mole/hr)

 $h_{IP_{sat}}$ = the enthalpy of the intermediate pressure saturation water at the inlet temperature (Btu/lb-mole)

 h_{FW} = the enthalpy of the feedwater (Btu/lb-mole)

 h_{IPS} = the enthalpy of the final intermediate pressure steam (Btu/lb-mole)

4.6. Cost model of WGS reaction process

This section presents the economic model developed for the water gas shift reaction process. The cost model is comprised of the capital cost model and the annual operating and maintenance (O&M) cost model. The capital cost of the WGS reaction system includes the following major process areas: the first stage shift reactor, the second shift reactor and the cooling units. For each of these major areas, its process facilities cost model is developed at first.

4.6.1. Process facility cost

The process facility cost of the reactor includes the reaction vessel, structural supports, dampers and isolation valves, ductwork, instrumentation and control, and installation costs. The reactor vessels are made of carbon steel. The process facility costs of the shift reactors are estimated based on the reactor volumes, which is assumed to be 1.2 times the catalyst volume [Doctor, 1994]. The catalyst volume calculation process is described in Appendix A.

Process facility cost of shift reactors

The process facility costs of the high and low temperature shift reactors are regressed as a function of reactor volume and operation pressure using the data in Table 4-3.

$$PFC_{R} = 0.9927 \cdot N_{T,R} [17.6487 (\frac{1.2V_{cat.}}{N_{O,R}})^{0.4883} P_{R}^{2.028}]$$
(R²=0.9) (4-17)

where PFC_R = the process facility cost of the reactor (US\$ in 2000)

 $N_{T,R}$ = the total number of the reactor trains

 $N_{O,R}$ = the number of the reactor operating trains

 $V_{cat.}$ = the volume of catalyst (m³)

 P_R = the operation pressure of the reactor (atm)

Table 4 - 3Water gas shift reactor cost data adjusted to the dollar value in 2000
[Doctor, 1996]

Cost (\$ in 2000)	Reactor volume(m ³)	Pressure(atm)
82864.8	22.6	31.1
38692.2	34	18.7
59189.0	9.684	31.0
21495.0	11.553	18.7

Process facility cost of heat exchangers

In this model, two types of heat exchangers are used, which are the gas-liquid type, and the gas-gas type. Generally, the cost of a heat exchanger depends on its heat exchange surface, which is determined by the heat load of the exchanger and the temperature difference between the hot and cold flows. To allow for variations in these parameters, the process facility cost of the gas-liquid type heat exchanger was regressed using the data in Table 4-4,

$$PFC_{HE1} = 1.0064 \cdot N_{T,HE} \cdot [13.7528 (dT_{HE})^{-0.6714} (\frac{Q_{HE}}{N_{O,HE}})^{0.6855}] \qquad (R^2 = 0.91) \quad (4-18)$$

where PFC_{HE1} = the process facility cost of the gas-liquid heat exchanger (US k\$ in 2000)

 $N_{T,HE}$ = the number of total train of the heat exchanger

 $N_{O,HE}$ = the number of the operating train of the heat exchanger

 Q_{HE} = the heat load of the heat exchanger (kW)

 dT_{HE} = the log mean temperature difference (C)

Table 4 - 4Gas-liquid heat exchanger cost data adjusted to the dollar value in 2000
[Doctor, 1996]

Cost (K\$ in 000)	Pressure (atm)	Log mean temperature difference (C)	Heat load (kW)
625.4	30.7	68.2	16421.6
615.0	30.7	90.8	21052.4
210.2	18.7	190.4	9298.0
168.2	19.4	148.6	5036.0
472.9	19.4	121.0	19534.9
315.3	19.4	13.7	1293.1
210.2	18.7	190.4	9298.0
99.8	19.4	153.5	2407.3
210.2	20.4	190.4	9298.0
634.6	68.1	52.0	12119.7
210.2	157.8	190.4	9298.0

Based on the data in Table 4-5, the process facility cost of the gas-gas type heat exchanger is given by,

$$PFC_{HE2} = 0.9927 \cdot N_{T,HE2} [24.4281 \cdot P_{HE2}^{0.2804} (dT_{HE2})^{-0.1143} (\frac{Q_{HE2}}{N_{O,HE2}})^{0.3881}]$$

where PFC_{HE2} = process facility cost of gas-gas heat exchanger (US k\$ in 2000)

 $N_{T,HE}$ = the total train number of the heat exchanger

 $N_{O,HE}$ = the operating train number of the heat exchanger

 Q_{HE} = the heat load of the heat exchanger (kW)

 dT_{HE} = the log mean temperature difference in the heat exchanger

Table 4 - 5Gas-gas heat exchanger cost data adjusted to the dollar value in 2000
[Doctor, 1996]

Cost (k\$ in 2000)	Pressure (atm)	Log mean temperature (C)	Heat load (kW)	
1757.3	30.7	98.0	17319.5	
1757.3	30.7	90.7	16776.2	
2205.4	19.4	10.0	42480.7	
3131.2	30.7	318.4	100832.3	
2606.0	31.6	340.4	95833.1	
897.1	897.1 68.1		1223.6	
2193.5	18.7	31.8	25641.0	
1294.8	18.7	19.4	4034.0	
644.3	20.4	69.1	2407.3	
849.9	20.4	71.4	5036.0	
692.1	20.4	57.5	2407.3	
966.5	18.7	51.2	5036.0	

4.6.2. Total capital requirement of WGS reaction system

The total process facilities cost of the water gas shift reaction system is the summation of the individual process facility costs above plus the cost of initial catalyst charge. This is added because it is also a large and integral part of the reaction system. Following the EPRI Technical Assessment Guide (1993), the total capital requirement and O&M cost of the WGS reaction system is given in the following table.

Value				
Sum of the PFC of each equipment				
10% PFC				
15% PFC				
20% PFC				
5% PFC				
General facilities+Project & Process				
Calculated based on discount rate and construction time				
0.5% PFC				
1 month of VOM&FOM				
0.5% TPC				
+Royalty fees+Preproduction fee+Inventory cost				
2% TPC				
40% of total maintenance cost				
30% of total labor cost				
1 jobs/shift				
\$250/ft ³ , replaced every 2.5 years				
\$250/ft ³ , replaced every 6 years				

Table 4 - 6Cost parameters of water gas shift process

REFERENCES (CHAPTER 4)

- 1. Campbell J.S., 1970: Influences of catalyst formulation and poisoning on activity and die-off of low temperature shift catalyst, Industrial & engineering chemistry process design and development, 9(4): 588
- 2. Davis R.J., 2003: All That Glitters Is Not AuO, Science, Vol. 301, Issue 5635
- 3. Dmitrievich A., 2002: Hydrodynamics, mass and heat transfer in chemical engineering. Taylor & Francis Press, New York
- 4. Doctor R.D., 1994: Gasification combined cycle: carbon dioxide recovery, transport, and disposal, ANL/ESD-24, Argonne National Laboratory, Argonne, IL
- Doctor R.D., 1996: KRW oxygen-blown gasification combined cycle carbon dioxide recovery, transport, and disposal, ANL/ESD-34, Argonne National Laboratory, Argonne, IL
- Enick R.M. and Busfamante F., 2001: Very High-Temperature, High-Pressure Homogenous Water Gas Shift Reaction Kinetics, 2001 AIChE Annual Meeting, Reno
- 7. Frank P., 2003a: Low Temperature Shift Catalysts for Hydrogen Production, Johnson Matthey Group
- 8. Frank P., 2003b: Sulfur Tolerant Shift Catalyst -Dealing with the Bottom of the Barrel Problem, Johnson Matthey Group
- Newsome D.S., Kellogg P., 1980: The Water-Gas Shift Reaction, CATAL. REV.-SCI. ENG., 21(2)
- Park J.N., Kim J.H., 2000: and Ho-In Lee, A Study on the Sulfur-Resistant Catalysts for Water Gas Shift Reaction IV. Modification of CoMo/ g-Al2O3 Catalyst with K, Bull. Korean Chem. Soc. Vol. 21, No. 12
- 11. ASME steam table (saturation: pressure), <u>http://www.e-cats.com/databook/Page%2047.htm</u>
- 12. Twigg M.V., 1989: Catalyst handbook, second edition, Wolfe publishing Ltd

Chapter 5. PERFORMANCE AND COST MODEL OF SELEXOL PROCESS FOR CO₂ CAPTURE

5.1. Introduction to the Selexol absorption process

The Selexol process uses a physical solvent to remove acid gas from the streams of synthetic or natural gas. It is ideally suited for the selective removal of H₂S and other sulfur compounds, or for the bulk removal of CO₂. The Selexol process also removes COS, mercaptans, ammonia, HCN and metal carbonyls [Epps, 1994].

The Selexol process, patented by Allied Chemical Corp., has been used since the late 1960s. The process was sold to Norton in 1982 and then bought by Union Carbide in 1990 [Epps, 1994]. The Dow Chemical Co. acquired gas processing expertise, including the Selexol process, from Union Carbide in 2001. The process is offered for license by several engineering companies—the most experienced of which with the process is probably UOP [UOP, 2002].

The Selexol process has been used commercially for 30 years and has provided reliable and stable operations. As of January 2000, over 55 Selexol units have been put into commercial service [Kubek, 2000], which cover a wide variety of applications, ranging from natural gas to synthetic gas. By now, Selexol process has been the dominant acid-gas removal system in gasification project. Moreover, increasingly interests to control CO_2 emission in the world may lead to Selexol application widely, particularly for coal gasification plants. Actually, the use of the Selexol solvent has a long history in gasification process, and was chosen as the acid-gas removal technology for the pioneering work in this area. Due to its outstanding record, the Selexol process continues to be the preferred choice for acid-gas removal today, and has recently been selected for

several large projects around the world [Breckenridge, 2000]. Relevant experiences for gasification are as follows [Kubek, 2000].

- About 50 Selexol units have been successfully commissioned for steam reforming, partial oxidation, natural gas, and landfill gas. Of these, 10 have been for heavy oil or coal gasifiers.
- The 100 MW Texaco/Cool Water (California) 1,000 t/d coal gasifier plant for IGCC demonstration was operated continuously for about five years in the 1980s. The Selexol unit performed extremely well. The process delivered H₂S-enriched acid gas to a Claus plant while removing 20 to 25% of the CO₂ and treating a high CO₂/H₂S ratio feed gas.
- The TVA/Muscle Shoals (Alabama) 200 t/d coal gasifier demonstration plant was operated continuously for about five years in the early 1980s. It employed a Texaco gasifier, a COS hydrolysis unit, and a Selexol unit to convert coal to clean synthesis gas, and CO₂ as an alternative feed to an existing ammonia-urea plant. The COS hydrolysis and Selexol units were stable and had a high on-stream factor. The Selexol unit delivered an H₂S-enriched acid gas to elemental sulfur production, a pure (< 1 vppm total sulfur) synthesis gas to NH₃ synthesis, and removed part of the CO₂ to provide high-purity CO₂ for urea production.

In this section, the technical background information of Selexol process is reviewed. This information is used to provide a basis for the development of performance models of Selexol systems for CO₂ emission control of IGCC plants.

5.2. Selexol solvent property

The Selexol acid gas removal process is based on the mechanism of physical absorption. The solvent used in the Selexol acid removal system is a mixture of dimethyl ethers polyethylene glycol with the formulation of $CH_3(CH_2CH_20)_nCH_3$, where n is between 3 and 9 [Epps, 1994]. The general properties of the glycol solvent is given in Table 5-1 [Sciamanna, 1988; Newman, 1985].

Property	Value
Viscosity @25C,cp	5.8
Specific gravity@25C,kg/m^3	1030
Mole weight	280
Vapor pressure @25C, mmHg	0.00073
Freezing point C	-28
Maximum operating Temp., C	175
Specific heat@25C Btu/lb F	0.49

Table 5 - 1Property of glycol solvent

The performance of a physical solvent can be predicted by its solubility. The solubility of an individual gas follows the Henry's law—the solubility of a compound in the solvent is directly proportional to its partial pressure in the gas phase. Hence, the performance of the Selexol processes enhances with increasing the partial pressures of sour gases. This is one of the major advantages of physical solvents, such as Selexol, over chemical solvents, such as methyldiethanolamine (MDEA), for acid gases removal from the high pressure syngas. As shown in Figure 5-1, compared to physical solvents, chemical solvents, such as methyldiethanolamine (MDEA) and diethanolamine (DEA), have higher absorption capacity at relatively low acid gas partial pressures. However, their absorption capacities plateau at higher partial pressures. The solubility of an acid

gas in physical solvents increases linearly with its partial pressure. Therefore, chemical solvent technologies are favorable at low acid gas partial pressures and physical solvents are favored at high acid gas partial pressures. Furthermore, the physical absorption allows for the solvent to be partially regenerated by pressure reduction, which reduces the energy requirement compared to chemical solvents.



Figure 5 - 1 Characteristics for Chemical and Physical Solvents [Sciamanna, 1988]

Higher partial pressure leads to higher solubility in physical solvents of all components of a gas stream, but the attractiveness of the Selexol system is that it has a favorable solubility for the acid gases versus other light gases. Comparing with some acid gases, H_2 and CO have much lower solubility in the solvent. For instance, as shown in Table 5-2, CO₂ is 75 times more soluble than H_2 , and H_2 S is 670 times more soluble than H_2 in Selexol.

Table 5 - 2Relative solubility of gases in Selexol solvent [Doctor, 1994]

Gas	CO ₂	H ₂	CH ₄	СО	H_2S	COS	SO ₂	NH ₃	N ₂	H ₂ O
Solubility	1	0.01	0.0667	0.028	8.93	2.33	93.3	4.87	0	733

Table 5-3 shows the actual solubility of various gases at 25°C in the Selexol solvent. The solubility data in Table 5-3 are based on single component solubility. It would be expected that these values should be approximately the same for non-polar components even in acid gas loaded solvents [Korens, 2003].

Table 5 - 3Solubility of Gases in the Selexol Solvent [Korens, 2002]

Gas	CO ₂	H ₂	CH ₄	СО	H_2S	COS	HCN	C ₆ H ₆	CH ₃ SH	H ₂ O
Solubility, Ncm2/g.bar @25°C	3.1	0.03	0.2	0.08	21	7.0	6600	759	68	2200

The solvent may be regenerated by releasing the absorbed sour gases. The regeneration step for Selexol can be carried out by either thermally, or flashing, or stripping gas. In addition to its solubility, the Selexol solvent has some other positive advantages to gasification applications [Kubek, 2000].

- A very low vapor pressure that limits its losses to the treated gas
- Low viscosity to avoid large pressure drop
- High chemical and thermal stability (no reclaiming or purge) because the solvent is true physical solvent and do not react chemically with the absorbed gases [Shah, 1988]
- Nontoxic for environmental compatibility and worker safety

- Non-corrosive for mainly carbon steel construction: the Selexol process allows for construction of mostly carbon steel due to its non-aqueous nature and inert chemical characteristics
- Non-foaming for operational stability
- Compatibility with gasifier feed gas contaminants
- High solubility for HCN and NH3 allows removal without solvent degradation.
- High solubility for nickel and iron carbonyls allows for their removal from the synthesis gas. This could be important to protect blades in downstream turbine operation.
- Low heat requirements for regeneration because the solvent can be regenerated by a simple pressure letdown

5.3. Technical Overview Selexol process for acid gas removal

This section presents a technical overview of Selexol absorption processes for sour gases removal, with particular focus on the effects of the sour remove requirements on the design of Selexol process.

Although a Selexol process can be configured in various ways, depending on the requirements for the level of H₂S/CO₂ selectivity, the depth of sulfur removal, the need for bulk CO₂ removal, and whether the gas needs to be dehydrated, this process always includes the following steps—sour gas absorption, solvent regeneration/sour gas recovery, and solvent cooling and recycle. These general steps for the Selexol process for acid gas removal are described by the following cases.





Figure 5 - 2 Selexol Flow Diagram for Selective H₂S Removal [Kubek, 2000]

A typical Selexol flow diagram for selective H_2S removal is shown in Figure 5-2. The feed gas and the lean solvent counter currently contact at high pressure and lower temperatures in an absorber, where desired levels of H_2S , COS and CO₂ are absorbed into the solution. Regeneration of the acid gas rich solvent is fulfilled through a combination of flashing and thermal regeneration. Acid gases absorbed in the solvent released first from one or more flash tanks at reduced pressures, then from the stripper by thermal regeneration with steam stripping at elevated temperatures and low pressure. A solvent heat exchange is employed to cool down the solvent. The regenerator overhead vapors (acid gas and steam) are routed to a condenser plus knockout drum, and the condensed water is returned to the unit to maintain water balance. The high- pressure flash gas vapors are compressed and returned to the absorber for greater H_2 and CO recovery and to provide H_2S -enrichment of the acid gas for the Claus plant.

5.3.2. Selexol process for H₂S and CO₂ removal

Through taking advantage of the high H_2S to CO_2 selectivity of Selexol solvent, Selexol solvent processes can also be configured to capture H_2S and CO_2 together with high levels of CO_2 recovery. This is usually accomplished by staging absorption for a high level of H_2S removal, followed by CO_2 removal. Figure 5-3 shows a Selexol process layout for synthesis gas treating where a high level of both sulfur and CO_2 removal are required. H_2S is selectively removed in the first column by a lean solvent, and CO_2 is removed from the H_2S -free gas in the second absorber. The second-stage solvent can be regenerated with air or nitrogen if very deep CO_2 removal is required.



Figure 5 - 3 Selexol Process for Sulfur and CO₂ Removal [Kohl, 1985]

A COS hydrolysis unit may be required if a high level of H_2S and COS removal is required. At the Sarlux IGCC plant in Italy, which gasifies petroleum pitch, the Selexol unit allows a COS hydrolysis step and gives an acid gas that is 50-80 vol.% H_2S to the Claus plant. This acid gas composition is the result of an H_2S enrichment factor of about 2 to 3 through the Selexol unit. The H_2S content of the purified gas from the Selexol absorber at that plant is about 30 ppmv [Korens, 2002].

5.3.3. An optimal design for Selexol process for sulfur and CO₂ capture from IGCC systems

A variety of flow schemes of Selexol processes permits process optimization and energy reduction. The following is a description of an optimal design of a Selexol process which removal sulfur and CO_2 from syngas from IGCC systems. This optimal design is based on revising a Selexol process, originally designed by UOP, for H₂S and CO₂ removal from syngas for the production of ammonia (UOP, 2002).

The H₂S Absorption flowsheet for the optimized configuration is shown in Figure 5-4. Syngas from the gas cooling section of the gasification process enters the H₂S absorber where it is contacted with CO₂-saturated Selexol solvent from the CO₂-removal portion of the facility. The pre-saturated solvent from the CO₂ removal area is chilled with refrigeration before fed into the absorber, which can increase the CO₂ and H₂S loading capacity of the solvent. The use of pre-loaded solvent prevents additional CO₂ absorption in the H₂S absorber, and it also minimizes the temperature rise across the tower, which negatively affects the H₂S solubility and the selectivity of the solvent. H₂S is removed from the CO₂ absorber. Therefore, significantly bulk CO₂ is removed in this pre-contacting stage which reduces the loading in the CO₂ Absorber. The rich solvent from the H₂S absorber is fed to the H₂S solvent regeneration facility.


Figure 5 - 4 Optimized Selexol absorption process for H₂S removal

Figure 5-5 presents a process flow diagram for the optimized H_2S solvent regeneration section. The rich solvent from the H_2S absorber is pumped to high pressure and heated in the lean / rich exchanger. The solvent then enters the H_2S solvent concentrator, which operates at a pressure higher than the H_2S absorber, thus the recycle gases can be recycled to the H_2S absorber without compression. Due to the relative difference in solubility in Selexol solvents, CO_2 is removed from solution preferentially over H_2S , which results in an enriched H_2S concentration in the solvent. The CO_2 removed in the H_2S solvent concentrator is the majority of the recycle gases back to the H_2S absorber. The enriched solvent from the H_2S solvent concentrator is flashed down to lower pressure. The flash gas again contains a higher proportion of CO_2 than H_2S . This stream is also recycled back to the H_2S absorber. This recycle stream is relatively small because much of the CO_2 was removed at high pressure. The solvent from the flash drum enters the Selexol stripper for regeneration.



Figure 5 - 5 Optimized H₂S Solvent Regeneration

The optimized CO_2 absorption flowsheet is shown in Figure 5-6. In this optimization design, the entire CO_2 solvent flow is contacted with the H₂S absorber overhead stream in the pre-contacting stage, which can unloads the CO_2 absorber. The heat of absorption is removed from this pre-contacting stage in a refrigeration chiller. The relatively high temperature of this stream allows setting high temperature refrigeration, which reduces the power consumption of the refrigeration system. The solvent is cooled to optimum absorption temperatures when the pressure is reduced in the flash regeneration portion of the facility. A portion of the rich CO_2 solvent is returned to the H₂S absorber as pre-saturated solvent. The remainder of the solvent is flash regenerated which will be presented below. The top bed of the tower uses lean solvent from the H₂S regeneration facility to contact the syngas. This allows for the CO_2 to be removed to levels lower than could be achieved using only flash regenerated (semi-lean) solvent.



Figure 5 - 6 Optimized Selexol process for CO₂ absorption

Rich CO₂ is flash regenerated as shown in Figure 5-7. The flash regeneration uses one sump tank, one or two power recovery turbines, and three stages of flash. The CO₂ rich solvent leaving the bottom of the CO₂ absorber is let down to the sump tank at a reduced pressure, where most H_2 and a tiny amount of CO₂ captured in the Selexol are released and recycled back to the pre-contacting stage.

Then the CO_2 rich solvent with high pressure is let down to one or two hydraulic power recovery turbines to recover the pressure energy before it is fed into three flash drums, where CO_2 is released at staged pressures to reduce the power consumption of CO_2 compression later.



Figure 5 - 7 Optimized Selexol regeneration through CO₂ flash

A key limitation of Selexol systems is the operating temperature requirement. The operating temperature for Selexol systems is typically approximately 100°F. Hence a reasonable location of Selexol process in an IGCC system is at the down stream of syngas cooling section.

5.4. Performance model of Selexol process

As a patented commercial solvent, the detailed characteristics of the Selexol solvent are not available. Hence in this section, a semi-analytical, semi-regression performance model of Selexol systems for CO_2 capture is presented.

5.4.1. Performance model of Selexol process for CO₂ capture

This section discusses the methodology of setting up a performance model of Selexol process for CO_2 capture. A cost model of the Selexol process is further developed on the basis of this performance model.

Temperature effect on solubility of gases in Selexol

The solubility of a gas in Selexol depends on its partial pressure and temperature. The solubility of CO_2 as a function of temperature is regressed based on published data [Doctor 1996, Black 2000] and given by,

$$\chi_{CO_2} = 0.0908 - 0.0008 \cdot T \tag{R^2=0.95} \tag{5-1}$$

where χ_{CO_2} = the solubility of CO₂ in the Selexol (SCF/gallon-psia)

T = the temperature of solvent with a range of $30 \sim 77$ °F

The solubility of other gases at different temperature is not available. Here the relative solubility of other gases to CO_2 at different temperature is assumed to be constant.

Solvent flow rate of the Selexol process

The input and output parameters of this model are given in Table 5-4. For the performance simulation, the first step is to calculate the flow rate of the solvent. In order to do this calculation, the whole Selexol process can be simplified as Figure 5-8. Stream 1 is the syngas fed into the absorber at a given temperature, and α percent of CO₂ is removed from the syngas. Stream 4 is the lean solvent at a design temperature. Due to

heat transfer between the solvent and syngas and the absorption heat, the temperature of the rich solvent (stream 3) will be increased by ΔT . For the given CO₂ removal percentage α , the flow rate of solvent, fuel gas and CO₂ can be calculated as follows.

Input parameter			Output para	meter		
	Flow rate (mole/s)	f_1			Flow rate (mole/s)	f_2
	Pressure	p ₁			Pressure	p ₂
	Temperature	T ₁			Temperature	T ₂
		[CO] ₁			[[CO] ₂
Syngas		[CO ₂] ₁		Fuel oas		[CO ₂] ₂
input	Molar concentrations	[H ₂] ₁		output		[H ₂] ₂
		[CH ₄] ₁		Molar	[CH ₄] ₂	
		$[H_2S]_1$			concentrations	$[H_2S]_2$
		[COS] ₁				$[COS]_2$
		[NH ₃] ₁				[NH ₃] ₂
		[H ₂ O] ₁				[H ₂ O] ₂
CO ₂ removal percentage				CO ₂ flow	Flow rate (mole/s)	f_5
				- 2	Pressure	P ₅
				Refrig. power	Power recovery	Comp. power

 Table 5 - 4
 Input and output parameters of Selexol model



Figure 5 - 8 Simplified Selexol process

As mentioned in the above section, the solubility of gases in Selexol is a function of temperature. For calculating the flow rate of solvent, the first step is to estimate the temperature change of solvent in the absorber. Assuming the flow rate of solvent is ω lb-mol/hr, the temperature increase of solvent in the absorber is given by

$$\Delta T = \Delta T_1 + \Delta T_2 \tag{5-2}$$

where ΔT = the temperature increase of solvent in the absorber (°F)

 ΔT_1 = solvent temperature increase caused by the heat transfer (°F)

 ΔT_2 = solvent temperature increase due to the solution heat of gases (°F)

According to the amount of heat transferred between the syngas and solvent, and the specific heat of the solvent, the temperature increase due to heat transfer is calculated by

$$\Delta T_{1} = \frac{Q_{1}}{\omega \cdot MW_{sel} \cdot C_{p,s}}$$
(5-3)

where MW_{Sel} = the molar weight of Selexol (280 lb/lb-mol)

$$C_{p,s}$$
 = the specific heat of Selexol (0.49 Btu/lb °F)

 Q_1 = the heat released by the syngas, which can be estimated according to the energy balance and given by,

$$Q_{1} = (T_{SG,i} - T_{SG,o})f_{SG,i} \{2.02 \cdot [H_{2}]_{1} \cdot C_{p,H_{2}} + 16 \cdot [CH_{4}]_{1} \cdot C_{p,CH_{4}} + 28 \cdot [CO]_{1} \cdot C_{p,CO} + 44 \cdot [CO_{2}]_{1} \cdot (1 - \alpha) \cdot C_{p,CO_{2}} \} + 44 \cdot (T_{SG,i} - T_{SG,o} - \Delta T) \cdot f_{SG,i} \cdot \alpha \cdot [CO_{2}]_{1} \cdot C_{p,CO_{2}}$$
(5-4)

where $T_{SG,i}$ = the syngas temperature at the inlet of the absorber (°F)

 $T_{SG,o}$ = the syngas temperature at the outlet of the absorber (°F)

 $f_{SG,i}$ = the molar flow rate of syngas at the inlet of the absorber (lb-mole/hr)

 $[i]_1$ = the molar concentration of species i in syngas at the inlet of the absorber

 $C_{p,i}$ = the specific heat of species i (Btu/lb °F), which is given in Table 5-5.

Table 5 - 5Specific heat of gases in the syngas

Gas	СО	CO ₂	H ₂	CH ₄	Ar	N_2	H_2S	NH ₃
Specific heat (Btu/lb F)	0.248	0.199	3.425	0.593	0.125	0.249	0.245	0.52

In Eq. 5-2, ΔT_2 is caused by the solution heat. Here only the solution heat of CO₂ is calculated. The solution heat of other gases is negligible because the amount of other gases captured by Selexol is much less than that of CO₂.

$$\Delta T_{2} = \frac{44f_{SG,i} \cdot [CO_{2}]_{1} \cdot \alpha \cdot \psi_{CO_{2}}}{\omega \cdot MW_{Sel} \cdot C_{p,Sel}}$$
(5-5)

where $f_{SG,i}$ = total flow rate of syngas entering the absorber (lb-mole/hr)

 $\alpha = CO_2$ removed from the syngas (%)

 ω = Selexol flow rate (lb-mole/hr)

MW_s = Selexol molecular weight (lb/lb-mole)

 $[CO_2]_1 = CO_2$ molar concentration at the inlet of absorber

 ψ_{CO_2} = solution heat of CO₂ in Selexol (Btu/lb-solute), and the solution heat of several gases is given in Table 5-6 [Korens, 2002].

 Table 5 - 6
 Solution heat (Btu/lb-solute) of gases in the Selexol

Gas	CO ₂	H_2S	CH ₃
Heat of solution (Btu/lb-solute)	160	190	75

In the flash tanks, the residual time is long enough to assume that equilibrium can be achieved in these tanks. In the last flash tank, the solvent temperature is about $(30+\Delta T_1)$, hence the volume and mass flow rate of the residual CO₂ in the lean solvent (S4 stream in Figure 5-8) can be given by:

$$V_{CO_2, res}(SCF/hr) = SV_{sel} \cdot \omega \cdot p_{CO_2} \chi_{CO_2}$$
(5-6)

$$m_{CO_2, res}(lb \cdot mol/hr) = \frac{V_{CO_2, res}}{SV_{CO_2}}$$
(5-7)

where SV_{sel} = the specific volume of Selexol (32.574 gallon/lb-mol);

 SV_{CO_2} = the specific volume of CO₂ (377.052 SFC/lb-mol);

 ω = the flow rate of Selexol (lb-mol/hr);

 p_{CO_2} = the partial pressure of CO₂ (psia);

 $\chi_{CO_2,4}$ = the solubility of CO₂ in Selexol at temperature of 30+ ΔT_1 .

According to the CO_2 capture percentage in the absorber, the amount of CO_2 that need be captured by the solvent is,

$$V_{CO_2,abs}(SCF/hr) = SV_{CO_2} \cdot f_{SG,i} \cdot [CO_2]_1 \cdot \alpha$$
(5-8)

In the absorber, the equilibrium cannot be achieved due to the limited residual time. The flow rate of solvent used in the absorber is larger than that of the solvent required to capture α percentage of CO₂ at equilibrium. The ratio of the actual flow rate to the equilibrium flow rate of the solvent was regressed based on published data [Doctor, 1994, 1996, Sciamanna, 1988].

$$\gamma = \frac{1.26}{\left(1 - \alpha\right)^{0.07}} - 0.0002 p_1 \tag{5-9}$$

where P_1 = the pressure of syngas at the inlet of absorber (psia).

Then the flow rate of Selexol for capturing α percentage of CO₂ is given by

$$\omega(lb \cdot mol / hr) = \frac{\gamma(V_{CO_2, res} + V_{CO_2, abs})}{SV_{sel} \cdot p_1 \cdot [CO_2]_1 \cdot \chi_{CO_2, 1}}$$
(5-10)

where $V_{CO_2,res}$ = volume flow rate of residual CO₂ in the lean solvent (lb-mole/hr)

 $V_{CO_2,abs}$ = volume flow rate of CO₂ captured in the absorber (lb-mole/hr)

$$\chi_{CO_2,1}$$
 = the solubility of CO₂ in Selexol at temperature of 30+ ΔT (°F)

Based on the above discussion, the calculation process for the flow rate of Selexol is concluded as in the following. First assuming the temperature of the Selexol solvent in the absorber is increased by $(\Delta T_1 + \Delta T_2)$, then the solubility of CO₂ at this increased temperature can be calculated. Second the solubility of CO₂ at the solvent in the last flash tank is calculated at the temperature $(30+\Delta T_1)$. Given the amount of CO₂ needed to be required, the flow rate of the solvent is calculated based on the solubility difference between the solvent in the absorber and in the last stage flash tank. Then the new values of ΔT_1 and ΔT_2 are computed using the calculated solvent flow rate of solvent. Such calculation process continues until the flow rate of the solvent is convergent. This calculation process is represented by Figure 5-9:



Figure 5 - 9 Calculation process for the flow rate of Selexol

Composition and flow rate of fuel gas

After CO_2 capture, the syngas is converted into the fuel gas, the main component of which is hydrogen. The composition and flow rate of the fuel gas can be calculated as follows.

With knowing the Selexol flow rate and solubility of gases, the volume and mass amount of species i which is captured by the solvent is:

$$V_{i}(SCF/hr) = SV_{sel} \cdot \omega \cdot p_{i} \cdot \chi_{i}$$
(5-11)

$$m_{i}(lb \cdot mol/hr) = \frac{V_{i}}{V_{i}}$$
(5-12)

where V_i = the volume flow rate of species i captured in the Selexol (SVF/hr);

 SV_{sel} = the specific volume of Selexol (gallon/lb-mol);

 v_i = the specific volume of CO₂ (SFC/lb-mol)

 ω = the flow rate of Selexol (lb-mol/hr);

 p_i = the partial pressure of species i in the syngas (psia);

 χ_i = the solubility of species i in Selexol at temperature of 30+ ΔT °F;

In the sump tank, most of the H₂, CH₄ captured in the Selexol are released and recycled to the absorber again. Because of the much higher solubility, only a tiny amount of CO₂ is released in the sump tank. The operating pressure of the sump tank is a design parameter. For this study, the operating pressure is determined to keep the loss of H₂ to Selexol solvent no more than 1% of H₂ in the syngas. The calculation process for the sump tank is as the follows: assuming the operating pressure is p_{sump}, the volume of species i released from the sump tank is $V_i^{'}$, then the partial pressure $p_{i,sump}$ can be given by Eq. (5-13). According to mass conservation, the total volume of species i captured in the absorber equals the volume released in the sump tank plus the volume retained in the solvent in the tank, expressed as Eq. (5-14). Now recalling the Eq. (5-11), the volume of species is retained in the solvent in the tank can calculated as Eq. (5-15). Iteratively calculating Eq (5-13), (5-14), and (5-15) until the partial pressures are converged. If at the given operating pressure, the H₂ volume retained in the solvent does not meet the design value, then the operating pressure is adjusted and the calculation is run again. The calculation procedure is given by Figure 3-10.

$$p_{i,sump} = \frac{V'_i}{\sum_i V'_i} p_{sump}$$
(5-13)

$$\mathbf{V}_{i} = \mathbf{V}_{i,\text{sump}} + \mathbf{V}_{i}^{'} \tag{5-14}$$

$$V_{i,sump} \left(SCF / hr \right) = 32.574 \omega \cdot p_{i,sump} \cdot \chi_i$$
(5-15)



Figure 5 - 10 Calculation process for the operating pressure of the sump tank Composition and flow rate of CO₂ rich flow

At each stage of the flash tanks, the flash pressure is given. At this pressure, the residual gases in the lean solvent can be calculated based on their solubility. Based on mass conservation, the composition and flow rate of CO_2 rich flow from the flash tanks can also be calculated, and the calculation procedure is similar to that shown in Figure 5-10.

5.4.2. Power consumption model of Selexol process

There is no heat duty in the Selexol process because the solvent is regenerated through pressure flashing, but the power input is required to compress the recycling gas from the sump tank, the lean solvent from the flash tank 3, and the CO_2 rich product. At the same time, some electricity can be generated through the power recovery hydro turbine. The total power consumption is the difference between the power input and the recovered power from the turbine.

Power recovery

In this performance model, the pressure of the high-pressure rich solvent from the absorber is reduced and the energy is recovered through one or two hydro turbines. According to the designs in other studies [Doctor, 1994, 1996, Sciamanna, 1988, Black, 2000], a thumb rule of design is concluded here. If the pressure of CO_2 rich Selexol flow is larger than 240psia, two power recovery turbines will be used. Otherwise, only one power recovery turbine will be used. Generally, this outlet pressure ($P_{o,1}$, psia) of the turbine can be determined based on the system pressure as following:

$$\mathbf{P}_{0,1} = 0.0402 \mathbf{P}_{1,1}^{1.415} \tag{5-16}$$

where $P_{0,1}$ = the outlet pressure of power recovery turbine 1 (psia).

 $P_{i,1}$ = the pressure of the CO₂-rich Selexol at the inlet of turbine 1 (psia), (150 $\leq p_{i,1} \leq 1000$).

If the pressure of the CO_2 rich Selexol flow is larger than 240 psia, then the outlet pressure of the second turbine is given by,

$$p_{0,2} = 35.619 \ln(p_{1,1}) - 169.88 \quad (240 \le p_1 \le 1000) \tag{5-17}$$

where $P_{0,2}$ = the outlet pressure of power recovery turbine 2 (psia)

 $P_{i,1}$ = the pressure of the CO₂-rich Selexol at the inlet of turbine 1 (psia), (240 ≤ $p_{i,1} \le 1000$)

The power recovered from the liquid solvent is calculated from the following expression [Doctor, 1994],

$$hp_{tur} = H_{Sel} \cdot \frac{f_{Sel_2}}{1714} \cdot \eta_{tur}$$
(5-18)

where hp_{tur} = the power recovered through the power turbine (hp)

$$H_{Sel}$$
 = the total dynamic head (lb/in²)

 f_{Sel_2} = the flow rate of CO₂ rich Selexol entering the turbine (gal/min)

 η_{tur} = the efficiency of the turbine

The temperature change of the solvent in the turbine can be calculated based on the change in enthalpy, which equals flow work, $\int v dp$. For the default efficiency of turbines, 78%, the temperature can be given by,

$$dT_{tur} = 0.0047 \cdot dP_{tur} - 0.0715$$
(5-19)

where dT_{tur} = the decreased temperature of the Selexol in the power recovery turbine (°F);

 dP_{tur} = the decreased pressure of the Selexol in the power recovery turbine (°F)

CO_2 compression

There are three flashing pressure levels for CO_2 release. The design of the flashing pressures in the three flashing tanks is an optimal problem, but a preliminary study showed that the effect of flashing pressures on the power consumption of the Selexol processes is not considerable. Hence, some default values are adopted here for the process design. If the system pressure is larger than 240 psia, the first flashing pressure equals the outlet pressure of the second turbine. If the system pressure is less than 240 psia, the first flashing pressure is set to be 25 psia. The second flashing pressure is set to be 14.7 psia, and the last flashing pressure is set to be 4 psia.

In each flashing tank, the gases released from solvent are calculated. CO_2 released from flash tank 2 and tank 3 is compressed to the flashing pressure of tank 1. The CO_2 stream is finally compressed to a high pressure (>1000psia) for storage using a multistage, inter-stage cooling compressor. The power required by the CO_2 compressors is estimated by [Doctor, 1994],

$$hp_{comp.} = \frac{0.00436}{\eta_{comp.}} \cdot VF_{gas} \cdot P_{comp,i} \cdot (\frac{k}{k-1}) \cdot [(\frac{P_{comp.,o}}{P_{comp.,,i}})^{(k_{gas}-1)/k_{gas}} - 1]$$
(5-20)

where $hp_{comp.}$ = the power consumption of the CO₂ compressor (hp)

 $\eta_{comp.}$ = the overall efficiency of the compressor

 VF_{gas} = the inlet rate of the CO₂ stream (ft³/min)

 $P_{comp,i}$ = the inlet pressure of the compressor (psia)

 $P_{comp,o}$ = the outlet pressure of the compressor (psia)

$$k_{gas} = \frac{C_{p,gas}}{C_{v,gas}}.$$

Solvent compression work

The CO_2 -lean solvent is pumped back to the absorber operating pressure by a circulation pump. The power required by the circulation pump is estimated in a similar way as Eq. (5-18),

$$hp_{pump} = H_s \frac{f_{Sel}}{1714\eta_{pump}}$$
(5-21)

where H_s = the total dynamic head (psia)

 f_{Sel} = the flow rate of CO₂ lean Selexol (gal/min)

 η_{pump} = the efficiency of the pump

Recycle gas compression work

The gases from the sump tank are recycled to the absorber. A compressor is used to compress the gases to the operating pressure of the absorber. The power of the compressor is estimated using Eq. (5-20).

Solvent refrigeration

Before the CO_2 -lean solvent fed into the absorber, it has to be cooled down to the absorber operating temperature (30F) by refrigeration. The refrigeration power is estimated by [Doctor, 1994],

$$W_{ref.} = \frac{refrigeration \ load(Btu/hr)}{1000(9 + \frac{T_{evap}}{10})}$$
(5-22)

where W_{ref} = the power consumption of the solvent refrigeration process (kW)

 T_{evap} = the evaporation temperature of the refrigerant (°F)

Makeup of the Selexol solvent

The vapor pressure of the Selexol solvent $is1.35 \times 10^{-5}$ psia at 77F, which is very low. The real vapor pressure is even lower because the operating temperature is usually lower than 77F. Hence, the loss of solvent due to evaporation is negligible. On the other hand, due to leakage, especially in the start on and turn off processes, a certain amount of solvent is lost. Here the annual loss of solvent is assumed to be approximate 10% of the total solvent in the system [UOP, 2003].

5.5. Cost model of the Selexol process

Similar to the cost model of the WGS reaction system discussed in Chapter 4, the outputs of this cost model include the process facility cost, total plant cost, total plant investment, total capital requirement, and O&M cost.

5.5.1. Process facility costs of the Selexol system for CO₂ capture

The major process facility costs of the Selexol system for CO_2 capture are considered as in the following.

CO₂ absorption column

Using the data in Table 5-7, the process facility costs of the absorption column is regressed as a function of the operating pressure, the flow rates of the solvent and syngas,

$$PFC_{abso} = N_{T,abso} \cdot [-1375 .356 + 16.536 P_{abso}] + 0.127628 (0.5 f_{Sel} + 0.5 f_{SG})]$$
(R2=0.90) (5-23)

where PFC_{avso} = the process facility cost of the absorber(US k\$ in 2000)

 $N_{T,abso}$ = the total train number of absorbers

 $P_{abso.i}$ = the inlet pressure of absorber (atm)

 f_{Sel} = the flow rate of the Selexol(lb-mole/hr)

 f_{gas} = the flow rate of the syngas (lb-mole/hr)

Table 5 - 7Absorber cost data adjusted to the dollar values in 2000 [Doctor, 1996]

PFC (2000\$)	P(atm)	Flow rate of syngas(lb- mol/h)	Selexol flow rate(lb- mol/hr)
6.3E+05	30.35	11771.88	11815.53
9.2E+05	10.21	12418.46	20802.84
1.5E+06	16.88	17614.58	23000
1.3E+06	68.05	17614.58	6900

Power recovery turbine

Based on the data in Table 5-8, the process facility cost of the power recovery turbine is given by,

$$PFC_{tur} = 219.086 + 0.080912 \cdot hp_{tur} + 0.020086P_{tur,o}^2$$
 (R²=0.91) (5-24)

where PFC_{tur} = the process facility cost of power recovery turbine (US k\$ in 2000)

 hp_{tur} = power output of the turbine (hp)

 $P_{tur,o}$ = the outlet pressure of the turbine (atm)

PFC (2000 k\$)	Outlet pressure	Power output(hp)
277.23	13.60	649
235.64	3.40	404
246.66	5.10	293
263.21	3.40	451
246.66	1.70	293
317.14	51.03	567
317.14	6.80	567

Table 5 - 8Power recovery turbine cost data adjusted to the dollar value in 2000
[Doctor, 1996]

Sump tank

The process facility cost of the sump tank is regressed as a function of the solvent flow rate,

$$PFC_{slump} = 2.0049 \cdot N_{T,slump} \cdot \left(\frac{f_{Sel}}{N_{O,slump}}\right)^{0.7446}$$
(R²=0.87) (5-25)

where PFC_{sump} = the process facility cost of the sump tank (US k\$ in 2000)

 $N_{T,sump}$ = the total train number of sump tanks

 $N_{O,sump}$ = the operating train number of the sump tanks

 f_{Sel} = the flow rate of the CO₂-rich Selexol entering the sump tank (kg/s), 400~800/train

PFC (2000 k\$)	Selexol flow rate (kg/s)
179.04	416.85
272.83	733.92
205.11	811.44
205.22	811.44

Table 5 - 9Sump tank cost data adjusted to the dollar value in 2000 [Doctor, 1996]

Recycle compressor

The process facility cost of the recycle compressor is given by,

$$PFC_{RC} = 4.45519 \, hp_{RC}^{0.7784}$$
 (R²=0.98) (5-26)

where PFC_{RC} = the process facility cost of the recycle compressor (US k\$ in 2000)

 hp_{RC} = the power consumption of the recycle compressor (hp)

Table 5 - 10	Recycle compressor cost data adjusted to the dollar value in 200)0
	[Doctor, 1996]	

PFC (2000 k\$)	Compressor capacity (hp)
576.64	537
361.19	259
212.55	151
212.55	151.3

Selexol pump

The process facility cost of the Selexol pump is given by,

$$PFC_{SP} = 1.2286 hp_{SP}^{0.7164}$$
 (R²=0.92) (5-27)

where PFC_{sp} = the process facility cost of the Selexol pump (US k\$ in 2000);

 hp_{sp} = the power consumption of the Selexol pump (hp).

PFC (2000 US k\$)	Pump capacity (hp)
301.52	2205
207.29	1282
326.63	2388
326633.3	2388

Table 5 - 11Selexol pump cost data adjusted to the dollar value in 2000 [Doctor,
1996]

CO₂ compressor

The process facility cost of the CO₂ compressor is regressed as,

$$PFC_{comp1} = 7.0321 h p_{comp}^{0.6769}$$
(R²=0.83) (5-28)

where PFC_{comp1} = the process facility cost of the CO₂ compressor (US k\$ in 2000)

 hp_{comp} = the power consumption of the compressor (hp)

Table 5 - 12CO2 compressor cost data adjusted to the dollar value in 2000 [Doctor,
1996]

PFC (2000, US k\$)	Compressor capacity (hp)
323.1754	600.41
311.5061	255
216.2418	155.52
190.1031	120.54
1026.139	1086
576.6455	539.71

CO₂ final product compressor

The process facility cost of the multi-stage CO₂ compressor is given by,

$$PFC_{comp2} = 13.0969 hp_{comp}^{0.64}$$
 (R²=0.85) (5-29)

where PFC_{comp2} = the process facility cost of the compressor (US k\$ in 2000)

 hp_{comp} = the horse power consumption of the compressor (hp)

Table 5 - 13	CO_2 final compressor cost data adjusted to the dollar value in 2000
	[Doctor, 1996]

PFC (2000 US K\$)	Compressor capacity (hp)
2162.421	2582
2851.544	2913
2565.347	3369
2382.109	3217

Refrigeration

The process facility cost of the refrigeration unit is regressed as,

$$PFC_{refr} = 1.0019 \cdot N_{T,refr} \cdot [16.4796 \cdot (\frac{f_{Sel}}{N_{O,refr}})^{0.3618} (\Delta T_{Sel})^{0.4064}] \qquad (R^2 = 0.97) \quad (3-30)$$

where PFC_{refr} = the process facility cost of the refrigeration unit (US k\$ in 2000);

 $N_{T,reft}$ = the total train number of the refrigeration unit;

 $N_{O,reft}$ = the operating train number of the refrigeration unit;

 f_{Sel} = the flow rate of the solvent entering the refrigeration unit (lb-mol/h), 70000~23000 /train;

 ΔT_{Sel} = the Selexol temperature difference between the inlet and outlet of the refrigeration unit (°C), 1~5 °C.

Table 5 - 14Refrigeration unit cost data adjusted to the dollar value in 2000 [Doctor,
1996]

PFC (2000 k\$)	Solvent flow rate (lb-mol/h)	Temperature difference (C)
657.73	12000	2.171
613.81	20802	1.017
771.71	7016	4.706
771.71	23397	1.667

Flash tank

The process facility cost of flash tanks is given by,

$$PFC_{tan k} = 0.9832 \cdot N_{T, tan k} \left(\frac{f_{Sel}}{N_{O, tan k}}\right)^{0.8005}$$
(R²=0.89) (5-31)

where $PFC_{tan k}$ = the process facility cost of the flash tank (US k\$ in 2000);

 $N_{T,\tan k}$ = the total train number of the flash tank;

 $N_{O,\tan k}$ = the operating train number of the flash tank;

 f_{Sel} = the flow rate of the Selexol entering the flash tank (kg/s), 400~800 /train.

 Table 5 - 15
 Flash tank cost data adjusted to the dollar value in 2000 [Doctor, 1996]

PFC (2000 \$)	Solvent flow rate (kg/s)	
129745.5	416.85	
197707.4	733.92	
205227.8	811.44	

5.5.2. Total Capital Requirement of the Selexol process

Here the default values for the capital cost calculation of the Selexol process for CO_2 capture are given by the following Table 5-16.

 Table 5 - 16
 Parameters for TCR of Selexol process

Total process facilities cost (PFC)	Sum of PFC of the major units in the process			
Engineering and home office	10% PFC			
General facilities	15% PFC			
Project contingency	15% PFC			
Process contingency	10% PFC			
Total plant cost (TPC) = sum of the above values				
Interest during construction	Calculated			
Royalty fees	0.5% PFC			
Preproduction fees	1 moth fee of VOM&FOM			
Inventory cost	0.5% TPC			
Total capital requirement (TCR) = sum of above values				
Fixed O&M cost (FOM)				
Total maintenance cost	2% TPC			
Maintenance cost allocated to labor	40% of total maintenance cost			
Administration & support labor cost	30% of total labor cost			
Operation labor	2 jobs/shift			
Variable O&M cost (VOM)				
Selexol solvent	\$ 1.96/lb			

REFERENCES (CHAPTER 5)

- Black W.B., Pritchard V., Holiday A., Ong J.O. and Sharp C., 2000: Use of SELEXOL Process in Coke Gasification to Ammonia Project By Presented at the Laurance Reid Gas Conditioning Conference, The University of Oklahoma, Norman, Oklahoma
- 2. Doctor R.D., 1994: Gasification combined cycle: carbon dioxide recovery, transport, and disposal, ANL/ESD-24
- 3. Doctor R.D., 1996: KRW oxygen-blown gasification combined cycle carbon dioxide recovery, transport, and disposal, ANL/ESD-34, 1996
- 4. Dow Chemical Company, 2004: Selexol solvent for gas treating, <u>www.dow.com</u>
- 5. Epps R., 1994: Use of Selexol Solvent for Hydrocarbon Dewpoint Control and Dehydration of Natural Gas, presented at the Laurance Reid Gas Conditioning Conference, Norman, OK
- 6. Gas Processes, 2002: Hydrocarbon Processing, UOP LLC, Des Plaines, Illinois.
- 7. IEA, 2003: Potential for improvement in gasification combined cycle power generation with CO₂ capture, Report number PH4/19
- 8. Newman S. A., 1985: Acid and sour gas treating processes: latest data and methods for designing and operating today's gas treating facilities, Gulf Publishing Co.
- 9. Kohl A.L. and Riesenfeld F.C., 1985: Gas Purification, Fourth Edition, Gulf Publishing Company
- Korens N., Simbeck D.R., Wilhelm D.J., 2002: Process Screening Analysis of Alternative Gas Treating and Sulfur Removal for Gasification, Revised Final Report, December 2002, Prepared by SFA Pacific, Inc. Mountain View, California
- 11. Personal communication with UOP, 2003
- Sciamanna S. and Lynn S., 1988: Solubility of hydrogen sulfide, sulfur dioxide, carbon dioxide, propane, and n-butane in poly(glycol ethers), Ind. Eng., Chem. Res., 27
- 13. Shah V.A., 1988: Low-cost ammonia and carbon recovery, Hydrocarbon Process., 67(3)
- 14. UOP, 2002: Use of SELEXOL Process in Coke Gasification to Ammonia Project, UOP report

Chapter 6. GREENFIELD IGCC POWER PLANT WITH AND WITHOUT CO₂ CAPUTRE

Parameter	Value	
Design ambient temperature	59 °F	
Design ambient pressure	14.7 psia	
ASU oxygen purity	95%	
Steam cycle	1400 psi/1000°F/1000°F	
Condenser pressure	0.67 psia	
Syngas sulfur removal efficiency	99%	
NO _x control	fuel gas moisturization	
Gasifier operation conditions	615 pisa/2450 °F	
Spare gasifier number	1	
Fuel type	Pittsburgh #8	

 Table 6 - 1
 Technical design assumption of the IGCC power plant

Table 6 - 2Economic and financial assumption of the IGCC power plant

Capacity factor	75%		
Fixed charge factor	14.8%		
Cost year	2000		
Construction period	4 years		
Lifetime	30 years		
Fuel price	1.26 \$/MBtu		
For CO ₂ capture plant			
CO ₂ capture efficiency	90%		
CO ₂ product final pressure	2100 psia		
CO ₂ transport and storage	10 \$/tonne		

This section applies the IGCC models in Aspen Plus to investigate factors influencing the performance and costs of IGCC power plants with and without CO_2 capture. At first, the effects of the quality of coals are studied. Then effects of CO_2

capture, plant size, and capital structures are also studied. The general technical design assumptions are given in Table 6-1 and the economic and financial assumptions are given in Table 6-2

6.1. The effects of coal types on IGCC performance

For a Texaco gasifier, coal is prepared in a slurry form. The composition of the slurry (for a given type of coal, the water percentage in the slurry by weight), may influence the gasifier efficiency and the efficiency of a whole IGCC power plant. To investigate the effects of water percentage in the slurry, an IGCC system with two GE 7FA gas turbines and two operating gasifiers was studied.

As an important factor determining the actual operation, as well as the economic feasibility of using a gasifier system, the gasification efficiency is defined as,

$$\eta_{\text{gasifier}} = \frac{H_g \cdot Q_g}{H_s \cdot M_s} \cdot 100$$
(6-1)

where $\eta_{gasifier}$ = gasification efficiency (%)

 H_g = is heating value of the gas (kJ/m³);

 $Q_g =$ is volume flow of gas (m³/s);

 H_s = is the heating value of gasifier fuel (kJ/kg);

 M_s = is the gasifier solid fuel consumption (kg/s).

If the lower heating values of the syngas and the fuel are used in the above equation, the gasification efficiency is the lower heating value gasification efficiency. Otherwise, it is the higher heating value efficiency.

Texaco gasifiers require the coal to be prepared in a slurry form for transport. The amount of water added depends on the composition of a coal, especially the carbon, ash and moisture percent in the coal. At first, the effects of total water percent in slurry on the performance of IGCC systems are studied. For Pittsburgh #8 coal, Figure 6-1 gives the effects of water percentage in slurry by weight on the gasification efficiency, the net plant thermal efficiency and the heat rate of the IGCC plant. The gasification efficiency is as low as 45% if no extra water is added to the coal, because at a given gasification temperature there is not enough oxygen to partially oxidize all the carbon in the feedstock. With the increase of the water percentage in the slurry, the gasification efficiency increases, and reaches the peak point, approximately 79%, at a total water percentage of 27% in the slurry by weight. The gasification efficiency decreases with further increasing the water percentage due to the increase of water content in the syngas. The thermal efficiency of the whole IGCC system shows the same trend as the gasification efficiency, which also shows that the gasification efficiency is a major factor influencing the performance of IGCC systems.



Figure 6 - 1 Effect of water percentage in slurry on IGCC performance

The effects of the water percentage in the slurry by weight on the total capital requirement and the cost of electricity are given by Figure 6-2. It is not a surprise to find that the there is an optimal water percentage for the COE and TCR of an IGCC power plant, because COE and TCR are heavily depends on thermal efficiency. However, the optimal value of water percentage in the slurry in this case is pure hypothetical and without considering the requirement of the slurryability. The slurryability of a given type of coal has a minimum requirement of water percentage in the slurry for transportation in pipes and pumps. For instance, in order to ensure the slurryability for transportation, the total water percentages in the slurry for Pittsburgh #8, Illinois #6, PRB and ND Lignite should be no less than 34%, 37%, 44% and 50%, respectively [Breton, 2002]. The hypothetical testing of all the four types of coal shows that the amount of water added in the slurry should based on the minimum requirement of the slurryability to avoid that the slurry composition is far away from the optimal value.



Figure 6 - 2 Effect of water percentage in slurry on TCR and COE

Although an entrained flow gasifier, like the Texaco gasifier, is able to gasify all types of coals regardless of coal rank, caking characteristics, or amount of coal fines, coal rank may influence the performance of gasifiers and IGCC systems. Here four coals are used to investigate this influence. These four coals represent bituminous coal, subbituminous coal, and lignite. The compositions of these coals are given in Table 6-3. The major feedstock parameters are carbon content, ash content, and oxygen content. The primary energy of coal is from the carbon content, which is reflected in the heating value of coal. Ash content in coal is a heat sink in gasification, and the oxygen content influences the oxygen requirement of gasifier performance; at this time there is a lack of empirical data for alternative (low rank) coals.

Dry basis						
	Pittsburgh#8	Illinois#6	Wyoming PRB	ND Lignite		
Coal rank	Bituminous	Bituminous	Sub-bituminous	Lignite		
HHV (Btu/lb)	13965	12529	11955	8989		
MOISTURE	5.05	13.00	30.24	33.03		
ASH	7.63	12.64	7.63	23.77		
CARBON	77.74	70.34	69.07	52.32		
HYDROGEN	5.14	4.83	4.74	4.00		
NITROGEN	1.50	1.33	1.00	1.15		
CHLORINE	0.06	0.20	0.01	0.13		
SULFUR	2.24	3.74	0.53	1.73		
OXYGEN	5.70	6.92	17.02	16.89		
Wet basis	Wet basis					
MOISTURE	5.05	13.00	30.24	33.03		
HHV	13260	10900	8340	6020		
LHV ²	12761	10381	7722	5431		
ASH	7.24	11.00	5.32	15.92		
CARBON	73.81	61.20	48.18	35.04		
HYDROGEN	4.88	4.20	3.31	2.68		
NITROGEN	1.42	1.16	0.70	0.77		
CHLORINE	0.06	0.17	0.01	0.09		
SULFUR	2.13	3.25	0.37	1.16		
OXYGEN	5.41	6.02	11.87	11.31		
Water percentage in slurry						
Water%	34	37	44	55		

 Table 6 - 3
 Compositions of the four coals and their water percentage in slurry

² LHV calculation is based on the following formula given by [George Booras, 2004]: LHV = $HHV - (91.1436 * H + 10.3181 * H_2O + 0.3439 * O)$, where H, H₂O, and O are on an asreceived basis.

Figure 6-3, using Pittsburgh #8 coal as the reference case, compares the gasification efficiency, thermal efficiency and heat rate of the IGCC power plant using the four types of coal. From this figure, it is clear that the rank of coal significantly influence the gasification efficiency and the thermal efficiency of the power plant, which increase with the increase of the heating value of coal. The heat rate of the IGCC power plant using lignite coal (ND lignite with a high heating value of 6020 BTU/lb) is about 33% percent higher than that of the IGCC plant using bituminous coal (Pittsburgh #8 with a high heating value of 13260 BTU/lb).



Figure 6 - 3 Effect of coal rank on the efficiency and heat rate of IGCC plants

The rank of coal also influences the economic factors of IGCC power plants. Figure 6-4 shows that low quality coal significantly increases the capital cost of an IGCC power plant. For instance, the total capital cost (\$/kW) of an IGCC power plant using ND coal is about 68% higher than that of an IGCC power plant using Pittsburgh #8. On the other hand, the lower quality coal has the lower fuel price (except ND lignite), which offsets the effect of coal quality on the cost of electricity. For instance, the cost of electricity of

an IGCC using the PRB coal is only 8.6% higher than that of an IGCC using the Pittsburgh #8 coal.



Figure 6 - 4 The effect of coal rank on the TCR and COE of IGCC plants (For the COE calculation, the coal price ratios based on the actual mine month coal price are: Pittsburgh #8: Illinois #6: PRB: ND Lignite=1:0.667:0.2:0.265)

The relative feed rates of oxygen and coal per MWh output of an IGCC using the four coals are compared in Figure 6-5. For a unit power output, the oxygen flow rate of the lower rank coal is bigger. For instance, the oxygen flow rate per MWh output of Illinois #6, PRB, and ND lignite are 1.2, 1.3 and 2.2 times of that of Pittsburgh #8. The relative flow rate of coal per MWh output shows the similar trend. The relatively higher feed rates of stock and oxygen require more capital cost and auxiliary power consumption for an IGCC power plant, which explain why lower rank coal deteriorates the performance of IGCC plants.


Figure 6 - 5 Relative oxygen and coal mass flow rate per MWh power generation

An IGCC power plant using the lower rank coal emits more CO_2 because of its lower energy efficiency Figure 6-6 shows that the CO_2 emission rate (kg CO_2/MWh) of an IGCC plant using ND lignite coal is more than 1.3 times higher than that of an IGCC using the Pittsburgh #8 coal.



Figure 6 - 6 Relative CO₂ emission for per MWh power generation

6.2. Effects of CO₂ capture efficiency

Many studies of CO_2 capture from IGCC power plants typically assumed a constant CO_2 capture efficiency in a range of 75% to 92%. CO_2 capture efficiencies used in these studies were determined by the study authors, and no studies published investigate the effect of different CO_2 capture efficiency on the performance of IGCC power plants. In this section, the performance of IGCC power plants, including the CO_2 avoidance cost, energy penalty, capital cost and cost of electricity, are studied with different CO_2 capture efficiencies. An optimal criterion is explored to determine the least-cost CO_2 capture efficiency for an IGCC power plant. The configuration of the IGCC system for this study is based on one GE 7FA gas turbine and one operating gasifier.

This study is based on the two-stage Selexol process for sulfur removal and CO_2 capture described in Chapter 5. At the first stage, 99% of sulfur content well as 7% of CO_2 is removal and vented into the atmosphere at the sulfur removing unit. After a two-stage shift reaction, there is approximately 0.5% CO not converted into CO_2 , and this additional amount is also emitted as CO_2 when the fuel gas is burned in the combustor. Hence, the maximum total CO_2 removal efficiency is approximately 92.5%. Here the total CO_2 removal efficiency is defined as:

$$CO_2$$
 removal efficiency = $\frac{CO_2 \text{ captured(mole)}}{\text{Total carbon in syngas from gasifier(mole)}}$

For the CO_2 captured, this study considers three situations: one is the CO_2 captured in the Selexol without compression; one is that the CO_2 captured is compressed to 2100 psia; the last one is that the CO_2 captured is compressed to 2100 psia, and transported and stored with a cost of 10\$/tonne CO_2 .

Figure 6-7 shows the power requirement and capital cost of the Selexol process for CO_2 capture. The power consumption for CO_2 capture varies slowly when the total CO_2 removal efficiency is lower than 80%. The power consumption rises quickly when the total CO_2 removal efficiency is higher than 80% because the total flow rate of Selexol increases quickly for very high CO_2 removal efficiency. Compared with the power consumption for compressing the CO_2 stream to 2100 psia, which is about 74 kWh/tonne- CO_2 , the power consumption of the Selexol process with 90% total CO_2 removal efficiency is about 44% of the power consumption for CO_2 compression.

The capital cost (k/tonne-CO₂ captured per hour) of the Selexol process (excluding CO₂ compression) reaches the lowest value of 49.2 when total CO₂ capture efficiency is in a range from 85% to 90%. Out of this range, the capital cost increases sharply.



Figure 6 - 7 Power requirement and capital cost of Selexol process for CO₂ capture

The thermal efficiency of an IGCC power plant with CO_2 capture is given in Figure 6-8. The thermal efficiency decreases with the increase of the total CO_2 removal efficiency. Compared with the thermal efficiency without CO_2 compression, compressing the captured CO_2 to 2100 psia reduces the thermal efficiency by 2 percent points when the total CO_2 removal efficiency is 90%.





Energy penalty is defined to study the influence of the CO_2 capture on the energy performance of an IGCC power plant as in the following,

$\mathbf{EP} = \frac{\mathbf{reference\ plant\ efficiency} - \mathbf{capture\ plant\ efficiency}}{\mathbf{reference\ plant\ efficiency}}$

Figure 6-9 gives the energy penalty of an IGCC power plant with different total CO_2 removal efficiency. Without CO_2 compression, the energy penalty is about 8% when the total CO_2 removal efficiency is 70%, and it rises to 10% when the total CO_2 removal efficiency is 90%. CO_2 compression further increases the energy penalty. For instance,

when the total CO_2 removal efficiency is 90%, the energy penalty including compression is up 15%.



Figure 6 - 9 Energy penalty for CO₂ removal (The thermal efficiency of the IGCC reference plant without in this case is 0.371)

The capital cost of an IGCC power plant is also significantly influenced by CO_2 capture. Figure 6-10 gives the total capital requirement (TCR) of an IGCC power plant with CO_2 capture. When the total CO_2 removal efficiency is lower than 0.9, the total capital requirement increases slowly with the increase of the CO_2 removal efficiency. When the total CO_2 removal efficiency is 0.9, the total capital requirement without CO_2 compression is about 1800 \$/kW, and it's approximately 11% higher when CO_2 compression is considered.



Figure 6 - 10 Total capital cost of an IGCC power plant with CO₂ capture

Figure 6-11 shows the TCR increase percentage between the capture plant and the reference plant. Without CO_2 compression, the TCR is increased by 16% when the total CO_2 removal efficiency is 90%. The TCR would be increased by about 30% when the captured CO_2 is compressed up to 2100 psia after capture.



Figure 6 - 11Total capital cost increase percentage of IGCC power plants with CO₂ capture (The total capital requirement of the reference plant is 1547 \$/kW in this study)

Cost of electricity (COE) is an essential factor to evaluate the economic performance of a power plant. Figure 6-12 shows the COE increase percentage under the three different situations. Compared to the COE of the reference plant, when the total CO_2 removal efficiency is 90%, the COE increase percentage without CO_2 compression, with CO_2 compression, and with CO_2 transportation and storage is 15%, 25% and 41%, respectively.



Figure 6 - 12 COE increase percentage of IGCC plants with CO₂ capture (In this case, the COE of the reference plant is 56 \$/MWh)

 CO_2 avoidance cost is used to evaluate the price paid for CO_2 capture, which is defined as:

$$\frac{\text{COE of capture plant} - \text{COE of reference plant}}{\text{reference plant CO}_2 \text{ emission rate} - \text{capture plant CO}_2 \text{ emission rate}}$$

Figure 6-13 shows the CO_2 avoidance cost of an IGCC power plant. When the total CO_2 removal efficiency is 0.9, comparing to the case without CO_2 compression, the CO_2 avoidance cost with CO_2 compression is increased by 1.7 times. When the transportation and storage cost is included, the CO_2 avoidance cost is 2.7 times of the cost without

compression. It is also noticed that no matter with CO_2 compression or storage and transportation, the avoidance cost always reaches the lowest point when the total CO_2 removal efficiency is around 90%.



Figure 6 - 13 CO₂ avoidance cost of IGCC plants

Figure 6-14 compares the CO_2 emission rates of the capture plant. The relative CO_2 emission rate quasi-linearly decreases with the increase of the total CO_2 removal efficiency. When the total CO_2 removal efficiency is 0.9, the CO_2 emission rate of the capture plant is 0.091kg/kWh for the capture only case and, it goes up to 0.097 kg/kWh for the capture and compression case, which is about the 11.7% of the emission rate of the reference plant.



Figure 6 - 14 CO₂ emission rate of the capture IGCC plant

6.3. Effects of plant size

Generally, the performance and cost of a power plant will vary with a change in the size of the plant because in a certain range, relatively large plants will be benefit from economy of scale and higher efficiency. This section shows the influence of the plant size on IGCC systems and CO_2 capture.

Here three sizes are investigated: one gasifier with one GE 7FA gas turbine, two gasifiers with two GE 7 FA gas turbines, and three gasifiers with three GE 7FA gas turbines. There is one spare gasifier for each plant. For the capture plant, the CO_2 capture efficiency is 90%, and the final CO_2 product is compressed to 2100 psia.

The cost of electricity, thermal efficiency, total capital requirement and the net output of each plant without CO_2 capture are shown in Figure 6-15. The plant size has notable influence on the total capital requirement. The capital requirement of the biggest plant is about 280 k less than that of the smallest one. Beside of the effect of economy of scale on the equipment, the lower capital cost percentage of the spare gasifier

in the bigger IGCC plant is also a major reason for the lower capital cost. The thermal efficiency is also improved with the increase of the plant size. For instance, the efficiency of the biggest plant is about 0.5 percentage points higher than that of the smallest one. Hence the cost of electricity also decreases with the increases of the plant sizes due to the lower capital requirement and higher efficiency.

The effects of the plant size on the cost of electricity, thermal efficiency, total capital requirement and net power output of IGCC plants with CO_2 capture, which are given in Figure 6-16, are similar to the effects shown in Figure 6-15.



Figure 6 - 15 Cost of electricity, thermal efficiency and total capital requirement of different size IGCC plants without CO₂ capture



Figure 6 - 16 Cost of electricity, thermal efficiency, and total capital requirement of different size IGCC plants with CO₂ capture (the COE of the capture plant includes the CO₂ transportation and storage cost at a value of 10 \$/tonne-CO₂)

The CO_2 avoidance cost, as shown in Figure 6-17 slightly decreases with the increase of the plant size. For example, the avoidance cost of the biggest plant is 29 \$/tonne-CO₂ captured, which is approximately \$2 lower than that of the smallest one.





6.4. Finance analysis of IGCC systems

Although IGCC systems show advantages in energy efficiency and emissions, investments to design and build commercial IGCC power plants in the world have not

solidly stepped forward due to financing, cost and risk concerns [Rosenberg, 2004]. One of the major issues hindering the application of IGCC is difficulty with financing. A key challenge with financing IGCC technology is that there is not enough information on which to make comparisons, or not enough experience bases in the marketplace.

Due to the large capital investment required by an IGCC power plant, typically, neither the manufacturer nor the owner can self-finance, or secure adequate financing using their non-project assets. So, project financing with an affordable capital structure is often the only way that IGCC technology can be built.

The term capital structure refers to the mix of debt and equity that is used to finance projects. A typical capital structure for a utility company is given in the following Table 6-4.

Title	Units	Value
Percent Debt	%	45
Percent Equity (Preferred Stock)	%	10
Percent Equity (Common Stock)	%	45

Table 6 - 4Capital structure of a typical power plant project (source: IECM
manual)

Cost of capital refers to the weighted costs of common stock, preferred stock (equity returns) and long term debt (debt interests) used to finance a project. For a project financed by debt and equity, the average capital cost is given by [Ross, 2005],

$$\mathbf{r}_{\text{WACC}} = (\frac{S}{S+B}) \times \mathbf{r}_{s} + (\frac{B}{S+B}) \times \mathbf{r}_{B} \times (1 - T_{C})$$

where r_{WACC} = average cost of capital after tax for the project

S = the amount of equity

B = the amount of debt

 r_s = the cost of equity

 r_B = the cost of debt (borrowing rate)

 T_c = the tax rate

Both the cost of equity and the cost of debt depend on the perceived risk of a project. As an emerging technology for power generation, IGCC is generally viewed to have higher risk than more mature power generation systems, such as PC power plants. Hence IGCC faces higher financing cost in the absence of incentive policies.

In order to stimulate deployment of IGCC technology, a 3-Party Covenant has been proposed, which is a financing and regulatory program aimed at reducing financing costs and providing a technology risk-tolerant investment structure [Rosenberg, 2004]. The 3-Party Covenant would be an arrangement between the federal government, state Public Utility Commission (PUC), and equity investors. The proposal would work as follows First, Federal legislation authorizes a federal loan guarantee to finance IGCC projects. The terms of the federal guarantee require that a proposed project obtain from a state PUC an assured revenue stream to cover return of capital, cost of capital, taxes and operating costs. The state PUC provides this revenue certainty through utility rates in states with traditional regulation of retail electricity sales. The equity investors (electric utility or an independent power producer) negotiate performance guarantees to develop, construct, and operate the IGCC plant. A fair equity return is determined and approved by the state PUC before construction begins [Rosenberg, 2004].

In short, the function of this 3-Party Covenant is to adjust the capital structure (debt to equity ratio) and reduce the interest rate of debt through federal guarantee. For instance, a typical interest rate of a mid-grade utility debt ranked as BBB was 6.5 percent in early 2004. With the federal guarantee, the debt would be ranked as AAA, and its interest would be reduced to 5.5 percent [Rosenberg, 2004].

Six different capital structures are used here to investigate the influence of the proposed 3-Party Covenant on the capital costs and energy costs of IGCC power plants, which are given in Table 6-5. The capital structures and resulting cost of capital from Case A to Case E reflect different debt-to-equity ratios. Case F gives a conventional capital structure for a power plant project.

Title	Unit	Case A	Case B	Case C	Case D	Case E	Case F
Real Bond Rate	%	5.5	5.5	5.5	5.5	5.5	6.5
Real Equity Return	%	11.5	11.5	11.5	11.5	11.5	11.5
Percent Debt	%	50	60	70	80	90	45
Percent Equity	%	50	40	30	20	10	55
Debt/Equity Ratio		1.0	1.5	2.3	4.0	9.0	0.8
Federal Tax Rate	%	35	35	35	35	35	35
State Tax Rate	%	4	4	4	4	4	4
Property Tax Rate	%	2	2	2	2	2	2
Cost of Capital							
(Before Taxes)	%	8.50	7.90	7.30	6.70	6.10	9.25
Fixed Charge Factor (FCF)	%	13.00	12.04	11.12	10.22	9.35	13.88

 Table 6 - 5
 Capital structures and cost of capital for IGCC financing

In the following simulation cases, the IGCC system has two GE 7FA gas turbines and two operation gasifiers. For comparison, the performance and costs of a PC power plant and a NGCC plant are also calculated using the IECM computer model. The PC power plant is super-critical with in-furnace NO_x control, cold-side ESP for particulate control, and a flue gas desulfurization system for SO_x control. The gross output of the PC plant is 500 MW. The fuel for the PC plant is also the Pittsburgh #8 coal with a price of 1.27 \$/MBtu. The NGCC power plant uses two GE 7FA gas turbines, and its steam cycle heat rate is 9496 kJ/kWh. The natural gas price is 3.797\$/GJ. The capacity factors for all these three type plants are 75%.

Figure 6-18 shows the total capital requirement of the reference IGCC power plant with the A to F capital structures and the total capital requirement of the PC plant with the F capital structure. With the same capital structure F, the total capital requirement of the IGCC plant is about 11% higher than that of the PC power plant. With the incentive 3-Party Covenant capital structures, the total capital cost of the IGCC plant reduces with an increasing debt-to-equity ratio, because higher debt percentage in the capital structure lowers the Allowance for Funds during Construction (AFDC).



Figure 6 - 18 Total capital requirement of the IGCC and PC plant based on different capital structures

The effect of the capital structure on the cost of electricity is show in Figure 6-19. Among the IGCC, PC and NGCC plants with the same conventional capital structure (Case F), the IGCC power plant has the highest COE, and NGCC has the lowest one. However, the COE of the IGCC with the capital structure of Case A is 43.9 \$/MWh, which is almost break-even with that of the PC plant. When the debt-to-equity ratio in the capital structure increases from Case A to Case B, the COE of the IGCC plant is even lower than that of the NGCC plant.



Figure 6 - 19 Cost of electricity of IGCC plant with different capital structures

The IGCC plant shows its advantage if CO_2 capture is included. From Figure 6-20, it is clear that even with the common capital structure, the total capital requirement of the IGCC capture plant is still lower than that of the PC capture plant.



Figure 6 - 20 Total capital requirement of IGCC and PC capture plants under different capital structures

With the same conventional capital structure, the COE of the IGCC capture plants, as shown in Figure 6-21, is about 22 percent lower than that of the PC power plant, and about 12 percent higher than that of the NGCC plant. When the debt-to-equity ratio

increases to 2.3, the COE of IGCC is the same as that of NGCC plant. However, it should be noticed that the natural price used here is a relatively low value based on recent U.S. gas prices. If the nature gas price goes up to 4.66 \$/MJ, calculation shows that with the same conventional capital structure, the COE of the NGCC capture plant will be same as that of the IGCC capture plant. Considering the highly volatile price of natural gas in the foreseeable future, even without any incentive policies an IGCC capture plant could be competitive with an NGCC capture plant.



Figure 6 - 21 COE of IGCC, PC and NGCC capture plants with different capital structures

In conclusion, many factors influence the performance and cost of IGCC power plants. For the current commercial gasifier designs that employ slurry coal feeding, the use of low rank coal significantly reduces the thermal efficiency of an IGCC plant. The plant size also is an important factor influencing the total capital cost of an IGCC plant due to the economy of scale. For CO₂ capture, there is an optimal CO₂ capture efficiency that minimizes the CO₂ avoidance cost. Based on the current CO₂ capture procedure, this optimal CO₂ capture efficiency is in a range from 85% to 90%. Finally, without an incentive financing approach, the IGCC power plant without CO₂ capture is less competitive than the PC and NGCC power plants in terms of both the total capital requirement and the COE. An incentive financing policy for IGCC power plants, like the 3-Party Covenant proposed by Rosenberg [2004], can help IGCC power plants enter into commercial operation more widely. Due to the advantages of IGCC plants for CO₂ capture, even without incentive financing policies IGCC capture plants are competitive with PC and NGCC capture plants.

REFERENCES (CHAPTER 6)

- Booras G., and Holt N, 2004: Pulverized Coal and IGCC Plant Cost and Performance Estimates, Gasification technologies, Washington, DC, October 3-6, 2004
- 2. Breton D. L. and Amick P., 2002: Comparative IGCC cost and performance for domestic coals, gasification technology conference, Oct., 2002, San Franscio
- 3. Rosenberg W.G., Alpern D.C., and Walker M.R., 2004: Deploying IGCC in This Decade With 3-Party Covenant Financing, ENPR discussion paper, discussion paper 2004-7, Harvard University, Cambridge, MA
- 4. Jaffe R.W., 2005, Corporate Finance, Seventh Edition, McGraw-Hill
- 5. IECM User Manual, 2005, Department of Engineering and Public Policy, Carnegie Mellon University, Pittsburgh, PA

Chapter 7. IGCC REPOWERING WITH CO₂ CAPTURE

North America has over 320,000 MWe of existing coal-fired power plants, which accounts for 35% of the total installed capacity and 49.8% of the total annual power generation in North America [Simbeck, 2001; Smock, 1990, EIA, 2004]. Most of the existing coal-fired power plant capacities are pulverized coal (PC) boilers that are 25-35 years old. These existing coal-based power plants have the highest CO₂ emission rate, due to the use of high carbon fuel (coal) and a relatively low thermal efficiency. What is the technical and economic potential to reduce CO₂ emissions from these existing power plants in the event that new environmental regulations place limits on carbon emissions? One recent study looked at retrofitting plants with an amine scrubber, and found this to be a costly measure that would substantially degraded plant performance [Rao, 2002]. However, IGCC repowering with CO2 capture offers a substantially different option to this problem.

IGCC repowering can be defined as the integration of gasification units, gas turbine generator units and heat recovery units into an existing steam power plant. Compared to other repowering technologies, IGCC repowering without CO_2 capture is usually considered to be less attractive due to the expense of the gasification units [Brander, 1992]. However, it does present several advantages. IGCC repowering can substantially increase the capacity and thermal efficiency of an old PC plant. The net output of a repowered IGCC plant can be up to three times or more of the original PC plant's output. At the same time, the emissions of NO_x , SO_x , Hg and solid waste can be dramatically reduced [Daledda, 1995; Bajura, 1995]. Shorter construction time and re-use of existing equipment (cooling system, steam turbine/generator units), infrastructure (road/railroad

connections, office building), and existing transmission capacity will reduce the capital cost relative to a new IGCC plant. Furthermore, re-use of the existing plant land can simplify the complicated site studies and authorization procedures [Makansi, 1994]. If the purpose of repowering is to mitigate CO₂ emissions, IGCC repowering can reduce CO₂ emissions while also improving capacity and efficiency.

This section provides an overview of the available options of using IGCC technology for repowering PC power plants. Then the decision factors which should be considered for an IGCC repowering project are discussed. Finally, the cost and performance of IGCC repowering are preliminarily analyzed, and results are summarized to show how IGCC repowering might be an attractive option for improving the performance of existing power plants.

7.1. IGCC repowering options

There are four major approaches for IGCC repowering, which are site repowering, feedwater heating repowering, boiler hot windbox repowering and heat recovery repowering [Sullivan, 1994; Najjar, 1994; Stenzel, 1995]. Each of them is discussed below.

7.1.1. Site Repowering

Site repowering is the simplest repowering option. It is to reuse the existing site to construct a new IGCC power plant or other types of power plants after demolishing existing units, except for keeping some reusable facilities, such as the cooling water system, switchyard and buildings. Site repowering has the advantage of being able to utilize the best available combined-cycle technology without having to make compromises to match the older existing components or systems. When compared to constructing a new unit on a new site, there can be savings in the permitting process, transmission access, and socioeconomic considerations for the local area that can make the site repowering a preferred option. The repowered plant performance would usually be identical to a new unit.

7.1.2. Feedwater heating repowering

Feedwater heating repowering uses the gas turbine exhaust to heat feedwater in an existing PC power plant. The steam previously extracted from steam turbines for feedwater heating is used to generate more power in steam turbines if the existing steam turbine design limits are not exceeded, or used to augment power output in the combustion turbine. In order to increase the availability, existing feedwater heaters can be retained to allow conventional operation when the combustion turbine or the gasifier is out of service. Feedwater heating repowering can improve the efficiency of the steam unit by about 15% [Brander, 1992].

7.1.3. Boiler windbox repowering

Windbox repowering utilizes the gas turbine exhaust as the combustion air for the existing boiler. Boiler windbox repowering technologies can add up to 25% additional capacity to the unit, improve the efficiency by 10-20%, improve the part load efficiency and cycling capability, and reduce NO_x emissions, but windbox repowering appears to be the highest degree of technical complexity of all the combustion-turbine-based repowering options [Stenzel, 1995].

A variant to the boiler windbox repowering approach includes a HRSG to reduce the temperature of the combustion turbine exhaust and produce additional steam. With this approach the existing windbox can be retained but will need to be enlarged, or the boiler will not produce the full steam output. This repowering configuration is commonly known as warm-windbox repowering and is used primarily to achieve heat rate reductions.

7.1.4. Heat recovery repowering

In the heat recovery repowering, the plant's existing boiler is replaced by a gasifier, combustion turbine and Heat Recovery Steam Generator (HRSG). Heat recovery repowering uses the gas turbine exhaust to generate steam in a HRSG. High efficiency is obtained by exchanging condensate, feedwater, and steam between the gasification system and the heat recovery steam generator.

7.1.5. Evaluation of repowering options

The site repowering is somewhat like building a greenfield IGCC power plant, which can be roughly estimated based on the performance and cost of a greenfield IGCC plant. For the feedwater heating and boiler windbox repowering, the existing boilers have to be kept, and it is necessary to control CO_2 emissions from the existing boilers as well as from the gasifier. Therefore, these two approaches do not fully take advantage of the low CO_2 capture cost of the gasification process. In the heat recovery repowering, the existing boiler is completely replaced by a gasifier, which is the only source of CO_2 . Hence, for the goal of CO_2 capture, only the heat recovery repowering approach is an attractive choice for IGCC repowering with CO_2 capture.

7.2. Decision factors for IGCC repowering

Evaluations for repowering projects must include a wide range of business aspects, load growth forecasts, financial parameters, environmental regulations, fuel cost ranges, fuel availability, legal issues and many other factors. A repowering analysis usually follows steps similar to those summarized below [Stenzel, 1995; Weinstein, 1999]:

- Determining the generation system goals; e.g., the amount and value of the needed additional power, emission reductions, fuel availability and costs, transmission requirements and/or limitations, forecasted generation load schedules, target electricity market price and/or other requirements and goals.
- Determining the existing plants that can be repowered to meet the generation goals by identifying the important site restrictions (e.g., emission limits), conditions of the existing equipment, and other important information.
- Identifying candidate repowering technologies and perform an initial analysis to reduce the repowering options to the most competitive technologies.
- Developing the design, operation parameters, capital costs, schedules and economics (the saving potential and simple pay-out time) for applicable repowered plants and optional new plants.
- Selecting the best option(s) based on economics and other factors.

There are a number of significant differences in considering IGCC repowering applications, which include the following.

Available space: Reusing old sites is one of the advantages of repowering, but IGCC repowering with CO_2 capture needs more equipment than other repowering approaches. For an IGCC repowering, the area for new units, the distance from gas turbine to existing steam turbines are important factors to consider. Hence, site space could be at a premium for some locations, and installation costs may be increased due to fitting new units in available space and more complicated layouts. Based on the space availability, for heat recovery repowering, the existing boilers can be demolished to provide more space for new units, or retired in place, or retained in standby for increased reliability states [Weinstein, 1999].

Heat rejection capability: Although the heat rejection from the steam turbine cycle is almost the same before and after the repowering, the low-energy, non-recyclable waste heat from the air separation unit and gasification process increases the total amount of heat rejection. In some cases this additional heat generated by gasification could exceed the heat rejection limitation permitted for a plant where condenser cooling is provided from a river, ponder or estuary. For cooling tower installations, this will result in an increase in condenser pressure, circulation water temperature, and tower evaporation. This system should be checked to assure that any cooling tower makeup water flow limitations are not exceeded and that certain critical auxiliary cooling water users, such as the generator coolers, do not exceed maximum temperature limits. Any such permitting limitations should be evaluated [Sullivan, 1994].

Transmission constraint on bulk transmission system: IGCC repowering can triple the capacity of an existing plant and the total capacity of the repowered plant may surpass the capacity of the original switchyard and transmission system.

Demineralized water availability: For IGCC repowering, power augmentation from the gas turbine and steam turbine is desired, steam or water is available with an associated increase in the demineralized water requirements.

Air emissions: Emissions from IGCC systems are typically controlled to meet strict environmental standards. Emissions limitations for the existing boiler can vary significantly depending on the control technology used, local permitting requirements, the age of the boiler, and other site specific conditions. Typically, IGCC emissions will be less than that of the boiler being replaced. This reduction in total emissions will also benefit the utility by allowing offsets in emissions at other sites.

Steam turbine capabilities: the conditions, capabilities, and limitations of the existing steam turbine are the most significant factors in determining the feasibility of IGCC repowering. Optimizing the existing steam turbine performance with the new combined-cycle components is important for the repowered unit to be able to compete with a new unit, even if it has lower capital costs. Selecting an appropriate size combustion turbine to match an existing steam turbine is a key factor to reach the optimal result. The following section will discuss this key issue: how to select and match a combustion turbine to an appropriately sized steam turbine for IGCC repowering with and without CO_2 capture.

7.3. Heat recovery repowering design

For a PC power plant, steam is generated in a one, two, or three-pressure boiler for delivery to a steam-generator. The boiler feed water is heated by steam extracted from the steam turbine. Figure 7-1 gives the schematic process of a PC power plant with a single-

pressure, non-reheat system cycle. This is the simplest steam cycle that can be applied in a PC plant. It results in a low installed cost. Although it does not produce the highest combined-cycle thermal efficiency, it is a sound economic selection when fuel is inexpensive.



Figure 7 - 1 One-pressure, non-reheat steam cycle with steam extraction for feedwater heating

Multi-pressure (two or three) steam cycles are used to maximize energy recovery from the boiler. Two or three-pressure steam cycles achieve better efficiency than the single pressure systems, but their installed cost is higher. They are the economic choice when fuel is more expensive or if the duty cycle requires a high load factor. Figure 7-2 shows a two-pressure, non-reheat steam cycle. Three-pressure, reheat steam cycle is shown in Figure 7-3. This cycle can achieve the highest energy efficiency, but the capital cost is also higher than other steam cycle options.



Figure 7 - 2 Two-pressure, non-reheat steam cycle with steam extraction for feedwater heating



Figure 7 - 3 Three-pressure, reheat steam cycle with steam extraction for feedwater heating

For a combined cycle power plant, like an IGCC plant or a NGCC plant, its steam cycle is similar to the PC power plant. Depending on the design criteria, the steam cycle could be a simple one-pressure style if the capital cost is more concerned than the thermal

efficiency (or fuel cost). It can also be a three-pressure, reheat style if the thermal efficiency is more concerned than the capital cost.

Unlike steam turbines, gas turbines are only available in discrete sizes. For the IGCC heat recovery repowering option, the capacity of the gas turbines and steam turbine should match well to fully utilize the waste heat from the gasification process and the gas turbine. For a greenfield power plant, it is not a problem to product a steam turbine with an appropriate size to match a given gas turbine. For a repowering project, however, a steam turbine has existed with a fixed maximum flow capability (power generation capacity). Once this is reached, no further output capability exists at the site.

There is a range of the steam turbine power output that that can be repowered with a given gas turbine. The range depends on the temperature and flow rate of the gas turbine exhaust, the throttle pressure and loading limitation of the existing steam turbine, and the heat recovery process employed. The low boundary of the range is achieved under the most restrictive condition—the steam turbine limitations are so severe that the repowering is only simple replacement of a non-reheat boiler by a gasifier, a gas turbine, and a HRSG with no modification to either the steam turbine or the feedwater heating system. This configuration is illustrated in Figure 7-4, which shows the repowered plant configuration with all existing feedwater heaters in service. This represents the minimum capital cost approach, which also results in the lowest output, lowest thermal efficiency alternative.



Figure 7 - 4 IGCC repowering with all feedwater heaters (minimum repowering)

If sufficient steam turbine low-pressure section flow passing capability is available, the low pressure feedwater heaters or all feedwater heaters can be removed from service as shown in Figures 7-5. These systems require additional heat transfer surface to be installed in the HRSG to heat the feedwater, which increases the capital cost. The increased plant output and efficiency may justify the added expense [Brander, 1992].

The maximum power output is achieved under the most ideal condition---the existing steam turbine has sufficient design margins, and the temperature of a gas turbine exhaust is high enough so that it can incorporate a three-pressure, reheat HRSG and eliminate all the feedwater heaters, as shown in Figure 7-6.



Figure 7 - 5 IGCC repowering with removing some feedwater heaters (medium repowering)



Figure 7 - 6 IGCC repowering without feedwater heaters (maximum repowering case)

7.4. IGCC repowering economic and performance analysis

As discussed above, a wide range of factors have to be considered when evaluating the performance and cost of a repowering project at a given site. Due to the variability from site-to-site, it is clear that there will be a wide range in the results of economic evaluation. As an example, consider using a Texaco quench oxygen-blown gasifier and a GE MS7001F gas turbine to repower an old PC plant using a steam turbine operating at 1400 psig throttle condition. Two simulation models are set up in Aspen Plus to evaluate the repower range of this configuration. One model simulates the most restrictive condition, or the minimum case—replacing the existing boiler with a gasifier, a gas turbine and a HRSG and no modification to the steam turbine and the feedwater system. In this case, the heat recovered from syngas cooling is only used to reheat and saturate the syngas fed into the gas turbine. Another model simulates the most favorable condition, or the maximum case—the steam turbine has sufficient design margins so that it can be incorporated into a three-pressure reheat HRSG, and remove all the feedwater heaters. In this case, part of the heat recovered from syngas cooling is used to reheat and saturate syngas, and the left heat is used for steam generation. The two models are further revised to incorporate the CO₂ capture function.

For the cost analysis, all existing equipment is assumed to be fully amortized, and the reusable utilities are assumed to be the coal handling facility, the dematerialized water unit, the boiler feed water system, the steam turbine and the generator. Other assumptions are given in Table 7-1.

Fixed charge factor	14.8%	Years of construction (yr)	3.5
Capacity factor	75%	Lifetime (yr)	30
Fuel type	Pittsburgh #8	Fuel price (\$/MBtu)	1.27
CO_2 transport and			
storage (\$/tonne)	10	CO ₂ final pressure (psia)	2100

 Table 7 - 1
 Economic and financial assumption for repowering studies

The performance of the repowering cases is given in Table 7-2. A Texaco quench gasifier and a GE MS7001F gas turbine without CO₂ capture can satisfy the steam requirements of a 60 MW steam turbine if a straight boiler replacement is done. If the cycle can be optimized using three-pressure reheat HRSG, the corresponding steam turbine size is approximately 110 MW. The total capital cost of the repowered plants without CO₂ capture ranges from \$1201/kW (maximum case) to \$1410/kW (minimum case) as compared to \$1547/kW of the greenfield plant. For the repowering plants with CO₂ capture, the capital costs range from \$1656/kW (maximum case) to \$2108 (minimum case) as compared to \$1995/kW of the greenfield plant.

	Case	TCR (\$/kW)	COE (\$/MWh)	ST power (MW)	Net plant output (MW)	Thermal efficiency (HHV)	CO ₂ emission (kg/kWh)
	Min. case	1410	57.5	60.1	225.5	31.1	0.986
Without	Max. case	1201	48.8	110.3	274.4	36.7	0.835
CO ₂ capture	Greenfield	1547	55.7	112.1	276.1	36.9	0.830
	Min. case	2108	92.7	60.5	192.1	24.2	0.126
With	Max. case	1656	72.5	120.4	251.6	31.3	0.098
CO ₂ capture	Greenfield	1995	78.3	122.3	253.4	31.5	0.097

 Table 7 - 2
 Study results of IGCC repowering with and without CO₂ capture

The repowering option with lower capital cost also has worse energy efficiency. The energy efficiencies of the repowering plants without CO_2 capture range from 31.1% (minimum case) to 36.7% (maximum case) as compared to 36.9% of the greenfield plant.

Without CO_2 capture, the COE of the minimum repowering case is slightly higher than that of the greenfield IGCC power plant. However, the COE of the maximum repowering case is around 12.3% lower than that of the greenfield plant. For the capture plant, the maximum case also has the lowest COE, but the COE of the minimum repowering case is much higher than the other two cases.





Using the greenfield IGCC without CO_2 capture as the reference plant, as shown in Figure 7-7, the maximum repowering IGCC plant with full CO_2 capture (capture, compression, storage and transportation) reduces the CO_2 avoidance cost from \$31/tonne to \$23/tonne. On the other hand, the minimum repowering case raises the avoidance cost to \$53/tonne. For CO_2 capture only without compression, the CO_2 avoidance cost is only \$3/tonne.

According to the above discussion, IGCC repowering with and without CO₂ capture may be an economically attractive option for existing PC power plants. Compared to building greenfield IGCC plants, IGCC repowering is less capital intensive and has a shorter construction period. Hence it also provides an option for introducing new power generation technology with lower risk to utilities. Under suitable conditions, IGCC repowering may be a cost-effective and attractive option for reducing CO_2 emissions from existing coal-fired plants. However, the cost and feasibility of repowering is very site specific. Hence, further research is needed to identify the most promising applications of IGCC repowering based on detailed site-specific assessments.
REFERENCES (CHAPTER 7)

- 1. Simbeck D.R., CO₂ Mitigation Economics for Existing Coal-Fired Power Plants, the Eighteenth Annual International Pittsburgh Coal Conference, 4 December, 2001, Newcastle, NSW, Australia
- 2. Robert Smock, Performance improvement of old plants gains favor, Power engineering, Nov. 1990
- 3. Rao, A.B. and Rubin, E.S. (2002). "A Technical, Economic and Environmental Assessment of Amine-based Carbon Capture Technology for Power Plant Greenhouse Gas Control," Environ. Sci. Technol. 36(20), 4467-4475.
- 4. J. A. Brander, D. L. Chase, Repowering application considerations, Journal of Engineering for Gas Turbines and Power, Oct. 1992, Vol. 144
- 5. Kenneth Daledda, Repowering cuts air pollution, improves station's efficiency, Power, April 1995
- 6. Rita A. Bajura and J. Christopher Ludlow, Repowering with coal gasification and gas turbine systems: an acid rain control strategy option, Morgantown Energy Technology Center, DOE
- 7. Jason Makansi, Repowering: options proliferate for managing generation assets, Power, June 1994
- 8. T.M. Sullivan and M.S. Briesch, Repowering: a ready source of new capacity, Energy Engineering, Vol. 91, No.2 1994
- 9. Y.S. H. Najjar and M. Akyurt, Combined cycles with gas turbine engines, Heat recovery systems & CHP, Vol. 14, No.2, 1994
- 10. William C. Stenzel, Dale M. Sopocy, and Stanley E. Pace, Repowering Existing Fossil Steam Plants
- 11. Richard E. Weinstein, Robert W. Travers Advanced Circulating Pressurized Fluidized Bed Combustion (APFBC) Repowering Considerations

Chapter 8. PERFORMANCE AND COST UNCERTAINTY ANALYSIS OF IGCC SYSTEMS

An IGCC plant is a complex chemical treatment and energy conversion system. Large scale, commercial experience with IGCC and Selexol systems for CO₂ capture is still limited. Consequently, there are substantial uncertainties associated with using the limited performance and cost data available to predict the commercial-scale performance and cost of a new IGCC plant. There are several types of uncertainty associated with a developing technology, such as the IGCC technology. These uncertainties include statistical errors, systematic errors, variabilities, and the lack of an empirical basis for concepts that have not been tested [Frey, 1994]. Uncertainties may apply to different aspects of the process, including performance variables, equipment sizing parameters, process area capital costs, requirements for initial catalysts and chemicals, indirect capital costs, process area maintenance costs, requirements for consumables during plant operation, and the unit costs of consumables, byproducts, wastes, and fuel. Model parameters in any or all of these areas may be uncertain, depending on the development state of the technology, the level of the performance and cost estimates, future market conditions for new chemicals, catalysts, byproducts, and wastes, and so on.

Given limited performance and cost data, as well as uncertainties associated with the complexity of IGCC systems, this chapter undertakes a systematic evaluation of performance and cost uncertainties and a ranking of the importance of different factors in terms of their potential contribution to the total uncertainty. In this study, the term uncertainty is used loosely to include variability (for example, in nominal process design values) as well as true uncertainty in the value of a particular parameter.

8.1. Methodology for uncertainty analysis

In this study, the parameter uncertainty analysis assumes that the total uncertainty can be calculated from an estimate of uncertainty in each of the parameters used in the performance and cost models. The technique of parameter uncertainty analysis provides a quantitative way to estimate the uncertainty in model results. The general approach to perform the parameter uncertainty analysis is given in the following steps [IAEA, 1989]:

- Define the assessment endpoint.
- List all uncertain parameters (include additional parameters if necessary to represent uncertainty in model structure).
- Specify maximum range of potential values relevant for uncertain parameters.
- Specify a subjective probability distribution for values occurring within this range.
- Determine and account for correlations among parameters.
- Using either analytical or numerical procedures, propagate the uncertainty in the model parameters to produce a probability distribution of model predictions.
- Derive quantitative statements of uncertainty in terms of a subjective confidence interval for the unknown value.
- Rank the parameters contributing most to uncertainty in the model prediction by performing a sensitivity analysis.

• Present and interpret the results of the analysis.

According to the above procedure, to perform a quantitative uncertainty analysis, the first step is to estimate uncertainties in specific process parameters, which involves the following several steps:

- Review the technical basis for uncertainty in the process
- Identify specific parameters that should be treated as uncertain
- Identify the source of information regarding uncertainty for each parameter

Depending on the availability of information, the estimate of a parameter uncertainty can be based on published judgments in the literature, published information that can be used to infer a judgment about uncertainty, statistical analysis of data, or elicitation of judgments from technical experts.

8.2. Probability distribution estimation of uncertainty parameters

For this study, reviewing the technical basis for uncertainty and identifying specific parameters that should be treated as uncertain had been completed along with the development of the technical and economic models. Then a probability distribution must be assigned to each of the uncertain parameters. Some of the probability distributions of parameters came directly from published judgments in the literature. Most of the other probability distributions were estimated through statistical analysis of data from reviewing published information. We note that using histograms of published literature values can sometimes provide a misleading estimate of uncertainty because some published literature values may have little bearing on how a system actually performs once it has been deployed. With this in mind, much more attention was paid to collect data from project reports and papers published by industrial companies with real-world experience.

8.2.1. Data visualization

After data collection, the data set for each parameter was visualized through plotting the data in figures. Specific techniques for evaluating and visualizing data include calculating summary statistics, plotting empirical cumulative distribution functions, representing data using histograms, and generating scatter plots to evaluate dependencies between parameters. The purposes of visualizing data sets include [Frey, 2002]:

- evaluating the central tendency and dispersion of the data;
- visually inspecting the shape of empirical data distribution as a potential aid in selecting parametric probability distribution models to fit to the data;
- identifying possible anomalies in the data set (such as outliers);
- identifying possible dependencies between variables.

8.2.2. Probability distribution selection

In choosing a distribution function to represent an uncertainty parameter, besides the data visualization, prior knowledge of the mechanism that impacts a quantity plays an important role. Probability distribution selection in this work was done in three steps.

As the first step, most of the probability distributions were represented by uniform distributions or triangular distributions to screen the most important parameters. Uniform probability is useful when it is possible to specify a finite range of possible values, but no information is available to decide which values in the range are more likely to occur than others. Triangular distributions also specify a range, but a mode is also specified. It is useful when we can specify both a finite range of possible values and a most likely (mode) value. For instance, for some input parameters, values toward the middle of the range of possible values are considered more likely to occur than values near either extreme. When this is the case, the triangular distribution provides a convenient means of representing uncertainty [Morgan, 1998]. Uniform and triangular distributions are excellent for screening studies and relatively easy to obtain judgments for relevant values. In addition to being simple, the shape of the uniform and triangular distributions can be a convenient way to send a signal that the details about uncertainty in the variable are not well known. This may help to prevent over-interpretation of results or a false sense of confidence in subtle details of model results [Morgan, 1998].

Once a particular distribution for an uncertainty parameter has been selected, a key step is to estimate the parameters of the distribution. The most widely used techniques for estimating the parameters are the method of maximum likelihood estimation (MLE), the method of least squares, and the method of matching moments [Morgan, 1998]. MLE was used in this study when a distribution more complicated than uniform and triangle is employed.

The fitted parametric distributions may be evaluated for goodness of fit using probability plots and test statistics. In this study, the empirical distribution of the actual

data set was compared visually with the cumulative probability functions of the fitted distributions to aid in evaluating the probability distribution model that described the observed data.

8.2.3. Distribution functions of uncertain parameters

During the model development process of IGCC system with CO₂ capture, a number of variables are determined as the uncertain parameters for preliminary uncertainty screening, which are given in Table E-1, E-2 and E-3 of Appendix E. Several of the parameters in the above tables were found to be correlated or expected to be correlated. The probabilistic simulations were exercised both with and without considering parameter correlations to determine if model results are sensitive to parameter correlation. Simulations using parameter correlations produced only minor effect on the results. Therefore, for convenience, the following case study presents the results based on uncorrelated sampling.

After preliminarily investigating uncertainty ranges of these parameters and their effects on performance and cost, this thesis focus on two key parameters, capacity factor and fuel price, for the following reasons. First, preliminary study shows that the uncertainties associated with these two factors have a significant influence on the CO₂ avoidance cost and on the cost of electricity (which is arguably the most important criterion for a power plant). Second, uncertainties (including variability) associated with most of the parameters of the IGCC process and the capture process would disappear or shrink after the specific plant is designed and installed, but during the operation period the capacity factor and the fuel price may still change frequently and widely due to changes in load requirements and fuel price volatility, as described below.

Capacity factor distribution

IGCC plants, as other coal-fired power plants, typically provide base load service, with nominal design capacity factor of 85% assumed in many recent studies (vs. the historical average of 67% for U.S. coal plants). To consider the possible range of capacity factors over the lifetime of IGCC plants, historical capacity factor data for power plants with capacity larger than 250 MW and age less than 30 years were collected from the DOE/NETL coal-fired power plant database, for the year 2000, and used to simulate the capacity factor uncertainty of an IGCC power plant.

Using the methodology expressed above, a distribution function for the capacity factor of an IGCC power plant was represented as a Weibull distribution. Table 8-1 compares the statistic properties of the data points with that of the fitted Weibull distribution. The empirical cumulative distribution function of the data points and the cumulative distribution function of the Weibull distribution are compared in Figure 8-1. These comparisons show that the Weibull distribution is a suitable presentation of the uncertainty (variability) associated with the capacity factor.

Table 8 - 1Statistical description of power plant capacity factor data and the fitted
Weibull distribution

Dataset	Mean	Median	Standard Deviation	Skewness
Coal Plant Data (>250 MW, <30 years old)	0.762	0.771	0.104	-0.712
Weibull distribution	0.764	0.775	0.106	-0.568



Figure 8 - 1 Empirical cumulative distribution functions of the capacity factor data and the distribution of the Weibull(8.5, 0.81) with Trunc(0, 1)

Fuel price distribution

Figure 8-2 represents the historical Central Appalachian coal prices in the New York Mercantile Exchange (NYMEX). From 1990 to 2000, coal prices decreased, but after 2000 coal prices began to increase and price volatility became notably larger. The historical coal prices shows the importance of considering the risk of an IGCC power plant exposed to volatile coal prices. Modeling and predicting the future coal prices are beyond the scope of this thesis. Rather, this thesis focuses on analysis of the effect of uncertainty of coal prices on IGCC systems based on the historical coal prices in the recent 15 years.



Figure 8 - 2 Central Appalachian coal price in the New York Mercantile Exchange (The original coal prices were given for July of each year; the prices shown here were inflation adjusted to the dollar value in 2000)

The distribution of coal prices is represented as a lognormal distribution. Table 8-2 gives the general statistical properties of the data points and the lognormal distribution. The empirical cumulative distribution function of the data points and the lognormal

distribution are given in Figure 8-3.

<i>Table 8 - 2</i>	Statistic description of coal price data and the fitted lognormal
	distribution

Dataset	Mean	Median	Standard Deviation	Skewness
Coal Price Data	1.14	1.02	0.38	2.35
Lognormal	1.17	1.14	0.27	0.64



Figure 8 - 3 Comparison of empirical cumulative distribution function of Central Appalachian coal price data with the distribution of Lognormal(1.169, 0.273)

8.3. Uncertainty analysis results

In order to analyze uncertainties, a probabilistic modeling environment is required. In this study, the uncertainty analysis was performed using the IECM computer model, which employs Monte Carlo simulation for uncertainty analysis. In a Monte Carlo simulation, a model is run repeatedly, using different values for each of its uncertain inputs each time. The values of each of the uncertain inputs are generated based on the probability distribution assigned to uncertain parameters, using Latin Hypercube sampling.

The following simulation results are based on an IGCC capture plant with two GE 7FA gas turbines, two operation gasifiers and one spare gasifier. Other design parameters and assumptions are the same as those in Chapter 6.

Figure 8-4 shows the uncertainties associated with the total capital requirement of the IGCC capture plant. The deterministic total capital requirement in this case is 1714 \$/kW. The value of the total capital requirement varies from 1660 to 1790 \$/kW, with a

90% confidence interval of [1687, 1760] when the uncertainties due to the IGCC process (parameters given in Table E-1) are taken into account. From this figure, it is clear that most of the uncertainty in the total capital cost comes from the IGCC process, rather then the capture process.



Figure 8 - 4 Cumulative distributions of the total capital requirement of the IGCC plant (Unc. of cap. process is given by Table E-2; Unc. of IGCC is given by Table E-1 in Appendix E; All factors take into account the uncertainties from Table E-1 and E-2)

Then, the effect of uncertainties of capacity factor and coal price on the cost of electricity is given in Figure 8-5. The deterministic value of COE is 69.9 \$/MWh. The uncertainties associated with the fuel price cause the COE to vary from 65 to 79 \$/MWh, with a 90 percentile range of 67~76 \$/MWh. If other parameters are fixed, there is a 63% possibility that the IGCC plants would have a higher COE than the deterministic estimate due to the assumed variability of the fuel price. Compared to the uncertainty of fuel prices, the assumed uncertainty of the capacity factor contributes more to the volatility of the COE. The capacity factor distribution makes the COE change from 59 to 106 \$/MWh, with a 90 percentile range of 63~94 \$/MWh. The combined uncertainty of fuel price and

capacity factor causes the COE to change from 58 to 108 \$/MWh, with a 90 percentile range of 62 to 96 \$/MWh. In this scenario, the possibility that the COE will be higher than that of the deterministic result is as high as 65%. In this scenario, the weighted average COE, which takes into account the value of COE and its probability, is 75.5 \$/MWh. The weighted average COE is 5.6 \$/MWh higher than the deterministic result. Because the weighted average COE takes into account the potential operating uncertainties of an IGCC plant, it maybe a more suitable measure of performance of the IGCC plant than the deterministic result based on a static situation.





The distributions of CO_2 avoidance costs are given in Figure 8-6. The distribution of the avoidance cost values are calculated based on the deterministic COE of the IGCC reference plant, and the probabilistic COE values of the IGCC capture plant. If both the fuel price and the capacity factor uncertainties are taken into account, the range of CO_2 avoidance cost is from 13.4 to 78.6 \$/tonne CO_2 , with a weighted average cost of \$36.6 (weighted average 1 in the figure).

It is interesting to compared this weighed average costs with two other CO₂ avoidance cost measures. One is the deterministic CO₂ avoidance cost which is calculated based on the deterministic COE values of the IGCC reference plant and the IGCC capture plant. The other measure is a weighted average CO₂ avoidance cost (weighted average cost 2 in the figure) which is calculated based on the weighted average COE of the IGCC reference plant and the IGCC capture plant. Hence, the difference between the three CO₂ avoidance cost is that the deterministic cost is calculated without considering uncertainties in both the reference plant and the capture plant; the weighted average 1 cost is calculated with considering uncertainties only in the capture plant; and the weighted average 2 cost is calculated considering uncertainties in both the capture plant and the reference plant. It is found that the deterministic CO₂ avoidance cost has the lowest value (\$29.5/ton), the weighted average 1 cost has the highest cost (\$36.6/ton), and the weighted average 2 cost has an intermediate cost of \$30.9/ton. Because the weighted average 2 cost takes into account the uncertain operating conditions in both the reference and the capture plant, this value is arguably the most realistic CO₂ avoidance cost. Not considering the uncertainty in the reference plant and the capture plant (or just considering the uncertainty in one of the two plants) will lead to either higher or lower estimates of the CO₂ avoidance cost.



Figure 8 - 6 Cumulative distribution of the CO₂ avoidance cost

The uncertainty and variability in IGCC systems with CO_2 capture come from the limited experience in producing, constructing and operating IGCC power plants with CO_2 capture. Most of the uncertainties associated with the capital cost of an IGCC capture plant come from the IGCC process itself. Assumptions about the fuel price and the capacity factor (especially the capacity factor) can change the estimated cost of electricity and the CO_2 avoidance cost significantly. Hence, using the most realistic capacity factor estimates to evaluate the performance of IGCC plants and CO_2 capture is especially important.

REFERENCES (CHAPTER 8)

- 1. Frey C., 1994: Probabilistic Performance, Environmental, and Economic Evaluation of an Advanced Coal Gasification System, Proceedings of the 87th Annual Meeting, Cincinnati, OH, June 19-24, 1994
- International Atomic Energy Agency (IAEA). 1989. Evaluating the Reliability of Predictions Made Using Environmental Transfer Models. IAEA Safety Series 100. Vienna, Austria
- 3. Rubin E.S. and A.B. Rao, 2002: Uncertainties in CO₂ capture and sequestration costs, GHGT-6 paper, 2002
- 4. Frey H.C. and J.Y. Zheng, 2002: Quantification of variability and uncertainty in air pollutant emission inventories: method and case study for utility NOx emissions, Journal of the air & waste management association, Vol. 52, Sep. 2002
- 5. Cullen A.C., H.C. Frey, 1999: Probabilistic techniques in exposure assessment—a handbook for dealing with variability and uncertainty in models and inputs, Plenum Publishing Corporation, 1999
- Morgan M.G., M. Henrion, M. Small, 1998: Uncertainty—a guide to dealing with uncertainty in quantitative risk and policy analysis, Cambridge University Press, 1998
- 7. Marco K., Carlo W., 2004: Fuel Flexibility, the European Gasification Conference, May 2004
- 8. McDaniel J.E., Hornick M., 2003: Polk Power Station ICGG 7th year of commercial operation, Gasification technologies, San Francisco, California
- 9. Keeler C.G., 2003: Operating Experience at the Wabash River Repowering Project, Gasification technologies, San Francisco, California
- 10. Ignacio M.V., 2003: Elcogas Puertollano IGCC Update, Gasification Technologies, San Francisco, California
- 11. Yamaguchi M., 2004: First Year Operation Experience with the Negishi IGCC, Gasification Technologies 2004, Washington DC
- 12. JGC Corporation, 2003: NPRC Negishi IGCC Startup and Operation, Gasification Technologies, San Francisco, California
- 13. Daslay C., BrkicBrkic C., 2003: The Experience of Snamprogetti's Four Gasification Projects, Gasification Technologies, San Francisco, California

14. Kaptur C.J., 2004: Trends in U.S. Domestic Coal Markets Are Higher Prices and Higher Price Volatility Here to Stay? Pinecock Perspectives, Issue No.58, Sep, 2004

Chapter 9. IGCC SYSTEMS WITH ADVANCED TECHNOLOGIES

While current IGCC power plants show relatively high energy efficiency and low environmental emissions, there is still much room for improvement in the performance and cost of IGCC plants. There are substantial R&D programs in the U.S. to improve the efficiency and cost-effectiveness of IGCC technology. During the next decade or so, IGCC technology is expected to make significant improvement in the following five areas [Todd, 2002; O'Brien, 2004]:

- Advanced gasifier concepts with higher efficiency, reliability, and higher operating pressure for more economic CO₂ capture;
- Advanced air separation units with better thermal integration with IGCC systems;
- Syngas cleanup process with less expensive particulate removal systems or hot gas filtration;
- Advanced gas turbines with high energy efficiency and capacity of burning syngas and hydrogen-rich fuels;
- Optimal integration with new technologies and components.

In particular, two research areas are likely to produce significant improvements in the performance and capital cost of the next generation IGCC power plants: advanced air separation processes and advanced gas turbines. The following sections discuss these novel technologies and their influence on the development and application of IGCC technologies in the near future.

9.1. Ion Transportation Membrane (ITM)-Based Air Separation Unit

The use of oxygen instead of air for gasification removes excess nitrogen from the gasifier and results in higher gasification efficiency, higher syngas heat value, lower capital costs (of gasifiers, heat recovery and downstream gas cleanup systems), substantially lower NO_x emissions, and better potential for CO_2 capture. Cryogenic air separation, pressure swing absorption, and polymeric membranes are common commercially available technologies for oxygen production. After making numerous refinements over a long time period, cryogenic air separation has now evolved as the most efficient way to produce oxygen at large scale, and has become the typical air separation process for oxygen-blown IGCC.

Current cryogenic air separation units of an IGCC system still account for about 15% of the plant capital cost, and consume about 10% of the gross power output [Stiegel, 2005]. Hence, reducing capital cost and increasing efficiency of ASU are important to improve economic viability, and to stimulate commercial deployment of IGCC power plants. However, the overall thermodynamic efficiency of cryogenic ASU is approaching its theoretical limit. So few significant technical breakthroughs are expected that would lead to dramatic oxygen cost reduction [Air Products, 2004].

A promising air separation alternative consists of highly selective and active membranes with high flux and selectivity to oxygen. Oxygen can be recovered at high temperatures by passing hot air over non-porous, mixed conducting ceramic membranes. These membranes, known as ion transport membranes (ITM), utilize an oxygen partial pressure differential across the membrane to cause oxygen ions to migrate through the membrane [Air Products, 2002].

The separation mechanism of this technology is illustrated in Figure 9-1. Oxygen molecules cling to the membrane surface on the high oxygen partial pressure side. Due to the catalytic properties of the surfaces of specialized ceramic materials, oxygen atoms are ionized by electrons. Then the oxygen ions diffuse across membrane due to the oxygen partial pressure differential across the membrane. Oxygen ions diffused through the membrane relinquish electrons and reform as oxygen molecules on the other side of the membrane (low oxygen pressure) [Air Products, 2004].



Figure 9 - 1 Separation mechanism of membrane-based oxygen production [Mathieu, 2002]

Due to the highly selective property of the membrane, impurities, such as nitrogen, are rejected by the membrane. The product gas from the ceramic membrane systems is virtually 100% pure oxygen. In addition, when ITM devices are built in practice, they have three valuable properties from operating point of view. First, the ITM devices require no moving parts, which lead to better reliability. Second, the ceramic membranes are insensitive to supply air contaminants. All the other air separation technologies, such as cryogenic air separation, suffer form sensitivity to moisture or the minor constituents

of air. Third, the deterioration and failure of a ceramic membrane can be readily detected due to a fall-off in the pressure of the output oxygen pressure [Air Products, 2003].



Figure 9 - 2 Process temperature and oxygen purity of different air separation technologies [Prasas, 2002]

When integrated with IGCC systems, ITM technology has another important advantage to improve the energy efficiency of IGCC systems over other air separation technologies. As shown in Figure 9-2, other air separation processes suffer from the lack of thermal synergy between the low temperature oxygen production and the high temperature gasifier operation. Oxygen produced from ITM is in the temperature range at which coal gasifiers operate. Hence, IGCC processes can be developed so as to include a significantly high level of thermal integration with air separation process. This will increase the overall process efficiency and reduce the cost of electricity.

Presently, there is a keen competition among several manufacturers to develop the membrane-based oxygen production technology. It is being developed under different nametags by different players, such as ITM (Air Products), OTM (Praxair), COGS

(Litton Life Support). Currently, it appears that Air Products Inc. is holding a leading position in this field.

With the support of the U.S. Department of Energy, Air Products has been leading a R&D program for ITM process since 1998 [Richards, 2001]. The goal of the three-phase program is to cut the cost of oxygen production by approximately one-third compared to conventional technologies, and demonstrate all the necessary technical and economic requirements for commercial scale-up. The research team has successfully addressed all technical and economic requirements for scale-up ITM technology and demonstrated over 2,300 hours of performance and stability of thin-film membrane structures in several experiments [Air Products, 2004]. According to Phase III of this program, a precommercial scale ITM process with approximately 25-50 ton-per-day (TPD) design capacity will be demonstrated in year 2008. After Phase III, the process may be introduced into the market at the 100's-of-TPD scale. The technology is expected to be ready for use in the large tonnage oxygen market (1000's-of-TPD) within a decade [Air Products, 2004].

9.2. GE H-class turbines

Improvement of the power block efficiency of IGCC system can further reduce the cost of electricity. Hence, the next generation of gas turbines is expected to enhance the economic competitiveness of IGCC plants. Key features of an advanced gas turbine technology to improve economical power generation of IGCC plants are [Smith, 2001]: low installed cost resulting from the capacity of the unit matching with a single large gasifier; high efficiency which reduces fuel consumption and plant cost by reducing the capacity of the gasification and cleanup system per unit of generation capacity

In 1995, GE introduced its new generation of gas turbines—the GE H System. This H System technology is the first gas turbine to achieve 60% fuel efficiency (LHV basis). Compared to the current gas turbines, like GE F-class turbines currently used in IGCC plants, GE H system is a state-of-the-art turbine system. The H System's pressure ratio is 23:1 which was selected to optimize the combined-cycle performance. This is a major change from the GE F-class gas turbines, which used a 15:1 pressure ratio. The firing temperature of H System is 2600 °F/1430 °C, which is about 200 °F/110 °C higher than the firing temperature of F class turbines [Matta, 2000].

The unique feature of an H turbine is the integrated heat transfer system, which combines the steam plant reheat process and gas turbine bucket and nozzle cooling [Matta, 2000]. This feature allows the turbine to be operated at a higher firing temperature and pressure ratio, which in turn produces dramatic improvements in fuel-efficiency. However, higher temperatures in the combustor also increase NOx emission. Using closed-loop steam cooling, GE H System solved the NO_x problem, and is able to raise firing temperature by 200 °F over the current GE F class of gas turbines and keep the NO_x emission levels at the GE F class levels.

In conventional gas turbines, the stage 1 nozzle is cooled with compressor discharge air. This cooling process causes a temperature drop across the stage 1 nozzle of up to 280 °F. In H System gas turbines, cooling the stage 1 nozzle with a closed-loop steam coolant reduces the temperature drop across that nozzle to less than 80 °F [Matta, 2000]. This results in a firing temperature 200 °F higher, and with no increase in combustion temperature.

An additional benefit of the H System is that the steam cooling the nozzle recovers heat for use in the steam turbine, transferring the heat was traditionally waste heat into usable output. The third advantage of closed-loop cooling is that it minimizes extraction of compressor discharge air, thereby allowing more air to flow to the combustor for fuel premixing [Matta, 2000].

Table 9-1 compares the performance of H class turbines and the F class turbines. The technology improvements shown in the GE H turbines are expected to yield substantial improvements in performance and significant reductions in the capital cost of IGCC systems [Brdar, 2000].

Table 9 - 1Performance characteristics of H-class and F-class turbines [Matta,
2000]

Gas turbine type	9FA	9Н	7FA	7H
Firing Temperature, °F (°C)	2400 (1316)	2600 (1430)	2400 (1316)	2600 (1430)
Air Flow, lb/sec (kg/sec)	1376 (625)	1510 (685)	953 (433)	1230 (558)
Pressure Ratio	15	23	15	23
NGCC Net Output, MW	391	480	263	400
Net Efficiency, % (LHV basis)	56.7	60	56	60
NO_x (ppmvd at 15% O_2)	25	25	9	9

9.3. IGCC systems with ITM oxygen production

In this section, a model to simulate the performance and cost of IGCC systems with ITM process is set up based on a simple ITM operation model.

9.3.1. ITM performance model

The performance of an ITM process can be simulated on the basis of a set of operating equations [Air Products, 2002]. As shown in Figure 9-3, compressed, hot air at

an absolute pressure P, temperature T, and with the composition x_{feed} , passes through an ITM vessel. Oxygen permeates the membrane at the oxygen low pressure side and is collected in the permeate stream at almost 100% purity and an absolute pressure, P_{perm} . The oxygen-depleted non-permeate stream emerges out of the ITM unit with the composition, x_{np} , and at an essentially unchanged pressure P. The device operates isothermally at a temperature T.



Figure 9 - 3 Simplified schematic process of an ITM unit

The oxygen-depleted non-permeate gas stream composition (x_{np}) can be calculated from an overall mass balance on the ITM unit. The overall recovery (R) is defined as the available fraction of oxygen recovered from the feed stream, which is shown as the following equation,

$$R = \frac{F_{perm}}{x_{feed} \cdot F}$$
(9-1)

where R= the overall recovery of the ITM unit

 F_{perm} = the molar flow rate of permeated oxygen (mole/hr)

F = the molar flow rate of air fed into the ITM unit (mole/hr)

 x_{feed} = the molar concentration of oxygen in the air

Theoretically, the overall recovery is ultimately limited by the driving force for oxygen flux. The driving force is due to the partial pressure difference of oxygen on both sides of the membrane. As the feed gas passes across the ITM device, the partial pressure of oxygen decreases since the gas is depleted of oxygen. The theoretical overall recovery is achieved when the oxygen partial pressure in the air falls as low as that in the permeate stream. The theoretical overall recovery can be calculated as,

$$R_{T} = 1 - \frac{(1 - x_{feed})P_{perm}}{x_{feed}(P - P_{perm})}$$
(9-2)

where R_T = the theoretical overall recovery of the ITM unit

 P_{perm} = the pressure of the permeated oxygen stream (psia)

P = the pressure of the air stream fed into the ITM unit (psia)

$$x_{feed}$$
 = the molar concentration of oxygen in the air feed

Consistent with many industrial separation processes, from an economical point of view, a commercial ITM separation process is best operated at 25% -85% of theoretical recovery. Hence, the practical overall recovery of an ITM unit is,

$$\mathbf{R} = \boldsymbol{\eta} \cdot \mathbf{R}_{\mathrm{T}} \tag{9-3}$$

where R = the practical overall recovery of an ITM unit

 η = percentage of the theoretical recovery

According to experimental data from the Air Products, a useful heuristic for calculating the separation performance of ITM is that the oxygen partial pressures in the permeate and in the feed streams are related as,

$$P_{O_2} \cong 7P_{\text{perm}} \tag{9-4}$$

where P_{O_2} = the partial pressure of oxygen in the air at the inlet of the ITM unit

 P_{perm} = the partial pressure of oxygen in the permeate side

According to the mass balance, the molar concentration of oxygen in the nonpermeate gas stream is given by,

$$\mathbf{x}_{O_2,np} = \frac{\mathbf{x}_{\text{feed}} \cdot \mathbf{F} - \mathbf{F}_{\text{perm}}}{\mathbf{F} - \mathbf{F}_{\text{perm}}}$$
(9-5)

Substituting Eq. 9-1 into the above equation, the molar concentration of oxygen in the non-permeate gas stream depending on recovery is given by,

$$x_{O_2,np} = \frac{x_{feed}(1-R)}{(1-R \cdot x_{feed})}$$
(9-6)

Air Products has recommended the operating conditions of ITM units which are summarized in Table 9-2.

Table 9 - 2Recommended operating parameters for ITM oxygen process design
[Air Product, 2002]

Recommended operating parameters for ITM design	Low	High
T (°C)	800	900
Feed pressure, P (psia)	100	1000

Permeate pressure, P _{perm} (psia)	1.9	100
Feed O_2 fraction	0.1	0.21
Percentage of theoretical recovery	25%	85%

As a new technology in the developing stage, there is no practical data available to build up its capital cost model. Air products estimated that ITM would be 32% cheaper than the conventional cryogenic technology [Air Products, 2004]. Hence the capital cost of an ITM unit is estimated as 68% of the capital cost of a cryogenic ASU with the same capacity.

9.3.2. IGCC designs with ITM air separation

Figure 9-4 represents a schematic of an IGCC system integrated with the ITM oxygen production. In this design, the oxygen production process is fully integrated with the gas turbine [Air Products, 2003]. Compressed air extracted from the gas turbine compressor is heated by the oxygen-depleted non-permeated air from the ITM unit. Then the compressed air is further heated by burning a portion of clean syngas to reach the operating temperature of the ITM unit. The ITM unit is exothermically operated at an essentially unchanged pressure. The oxygen stream from the ITM unit at a low pressure and a high temperature is cooled down to produce steam for steam cycle, and then compressed to a pressure suitable for gasifier operation. The heat from the hot oxygen-depleted non-permeate air is used to pre-heat the inlet air of the ITM unit. The cooled oxygen-depleted non-permeate air is then fed into the gas turbine combustor.

Figure 9-5 shows a revised design of IGCC system with ITM oxygen production. In this design, the air fed into the ITM unit comes from a standalone compressor. This

design requires one more air compressor, but offers more flexible operation options. The two designs are implemented in Aspen simulation model, and the design parameters of the two cases are given in Table 9-3.

Table 9-3 gives the ITM operating conditions of two IGCC systems with the ITM process. These two cases are used to investigate the influence of ITM on the performance and cost of IGCC systems. Other technical and economic assumptions for these case studies are the same as those given in Table 6-1 and 6-2.

 Table 9 - 3
 Design parameters of ITM units in the Aspen simulation models

Case number	ITM operating temperature (F)	ITM air feed pressure (psia)	ITM percentage of theoretical recovery	Notes
ITM-A	1500	250	80%	Integrated with GT
ITM-B	1500	200	80%	Standalone ITM



Figure 9 - 4 Overview of an IGCC system fully integrating with the ITM oxygen production



Figure 9 - 5 Overview of an IGCC system with standalone ITM oxygen production

Table 9-4 shows the performance and cost of the two ITM cases and the corresponding cryogenic ASU case. Comparing to the cryogenic ASU case, ITM cases show significant improvement on the performance and cost of IGCC systems. For example, the net efficiency of the IGCC plants with ITM increases approximately 2% percentage points, an improvement of about 5.7% over the cryogenic case. The net power output increases by 37 MW. The total capital requirement per kW reduces from \$1311 to \$1240, and the cost of electricity also reduces by about 2.6 \$/MWh. Corresponding to the improvement of the net efficiency due to the ITM technology, the CO₂ emission also reduces by 5.5%.

 Table 9 - 4
 Performance and cost comparison of IGCC reference plants with ITM

Case	Effic- iency (HHV)	Net power (MWe)	ASU TCR (M\$)	TCR (M\$)	TCR (\$/kW)	COE (\$/MWh)	CO ₂ emission (kg/kWh)	O ₂ flowrate (TPD)	ITM TCR (k\$/TPD)
ITM-A	39.3	574.2	59.7	712.3	1240	45.71	0.78	3623.8	16.5
ITM-B	39.4	575.8	59.9	716.9	1245	45.84	0.78	3644.0	16.4
Cryo.	37.1	537.9	89.3	705.7	1311	48.40	0.83	3650.2	24.5

Next, the WGS reactor and Selexol process for CO_2 capture are incorporated into the Case ITM-B to study the effect of adopting ITM technology on the performance and cost of the capture plant. Table 9-5 shows the performance improvement of IGCC power plant with CO_2 due to the ITM technology.

	ITM	Cryo.	% Change
Net efficiency	33.6	32.0	5.2
Net power output (MW)	538.1	502.2	7.1
Plant TCR (\$/kW)	1631	1714	-4.8
COE (\$/MWh)	66.38	69.91	-5.0
CO ₂ emission (kg/kWh)	0.092	0.096	-4.8

Table 9 - 5Performance and cost comparison of IGCC capture plant with ITM

9.4. IGCC systems using GE H turbine

Different from GE F gas turbine, GE H gas turbine uses a closed steam cooling system, which requires a different gas turbine cooling and heat recovery system. Figure 9-6 shows an overview of the power block of an IGCC system using GE H turbine, which is a three-pressure, reheat steam cycle and its integration with the gas turbine cooling system. Gas turbine cooling steam is extracted from the high pressure steam turbine exhaust to the closed circuit system that is used to cool the gas turbine stage 1 and 2 nozzles and buckets. The steam in the cooling circuit system is heated to approximately the reheat temperature of the steam cycle, and returns to the immediate pressure steam turbine.

A syngas heating system utilizes low grade energy from the HRSG to improve combined-cycle thermal efficiency. Water extracted from the discharge of the HRSG IP economizer is supplied to the syngas heater to pre-heat and saturate the syngas before it is supplied to the combustion system. The water leaving the fuel heater is returned to the cycle through the condensate receiver to the condenser.



Figure 9 - 6 Combined power block cycle with GE H turbine [Matta, 2000]

Due to the higher exhaust temperature from GE H turbine, the steam cycle can use higher pressure and temperature to achieve better energy efficiency. The parameters of this three-pressure, reheat steam cycle used in the simulation model are given in the following table. The other general technical and economic parameters these case studies are the same as those given in Table 6-1 and 6-2.

Table 9 - 6Steam cycle parameters of the IGCC using GE H turbine

Parameter	HP throttle	Hot reheat	LP admission
Pressure (psig/Bar)	2400/165	345/23.8	31/2.2
Temperature (°F/°C)	1050/565	1050/565	530/277

The capital cost of the H turbine is estimated with the gas turbine cost model developed in Chapter 3, which is a function of the net power output the gas turbine. Simulation results show that GE H gas turbines can greatly improve the performance of IGCC power plants. As shown in Table 9-7, comparing to an IGCC with GE F turbine,

the utilization of GE H turbine can increase the net efficiency of IGCC systems by more than 4 percentage points, reduce the capital cost per kilowatt by more than 4%, lower the cost of electricity by more than 7%, and reduce the CO_2 emission rate by about 10%. The effects of the utilization of GE H turbine on the performance of CO_2 capture plant are given in Table 9-8.

Parameter	H turbine	F turbine	% Change
Net efficiency	41.4	37.1	11.4
Net power (MW)	860.3	537.9	59.9
TCR (\$/kW)	1256	1311	-4.2
COE (\$/MWh)	44.82	48.40	-7.4
CO ₂ emission (kg/kWh)	0.74	0.83	-10.4

Table 9 - 7Performance and cost comparison of IGCC reference plants using
different gas turbines

<i>Table 9 - 8</i>	Performance and cost comparison of IGCC capture plants using
	different gas turbines

Parameter	H turbine	F turbine	% Change
Net efficiency	36.2	32	13.1
Net power (MW)	814.6	502.2	62.2
TCR (\$/kW)	1573	1714	-8.2
COE (\$/MWh)	62.51	69.91	-10.6
CO ₂ emission (kg/kWh)	0.085	0.096	-11.7

9.5. IGCC system with ITM and H turbine

IGCC systems adopting ITM oxygen production and GE 7H turbine with and without CO_2 capture are also modeled and simulated. Table 9-9 presents the performance of an IGCC employing both ITM and GE 7H gas turbine. This next generation IGCC system with advanced technology expected to be available in the next decade, can

achieve a thermal efficiency as high as 42.3% on a higher heating value basis. The total capital requirement of the IGCC plant with advanced technologies lowers to 1184 \$/kW, which is comparable to or lower than the capital cost of current PC power plants. Due to the lower capital cost and higher energy efficiency, the cost of electricity of the next generation IGCC power plant also is estimated to be slightly lower than that of the current supercritical PC plants.

Parameter	ITM-H turbine	Cryo-F turbine	% Change
Net efficiency (%, HHV)	42.3	37.1	13.9
Net power (MW)	929.5	537.9	72.8
TCR (\$/kW)	1184	1311	-9.7
COE (\$/MWh)	42.30	48.40	-12.6
CO ₂ emission (kg/kWh)	0.73	0.83	-12.3

Table 9 - 9Performance and cost improvement of the IGCC reference plant using
ITM and GE 7H turbine

From Table 9-10, it is noticed that the energy efficiency of the next generation IGCC system with CO_2 capture is approximately 38.2%, which is higher than that of current IGCC systems without CO_2 capture. The total capital cost of the next generation IGCC systems with CO_2 capture is 1470 \$/kW, which is only about 10% higher than that of the current IGCC systems without CO_2 capture.

The preliminary performance and cost analysis of IGCC systems with emerging advanced technologies for oxygen production and gas turbine power generation are expected to greatly improve the performance of IGCC systems. Simulation results show that these new technologies will almost offset the influence of CO_2 capture on the performance and cost of current IGCC plants.
Parameter	ITM-H turbine	Cryo-F turbine	% Change
Net efficiency (HHV)	38.2	32.0	19.5
Net power (MW)	902.9	502.2	79.8
TCR (\$/kW)	1471	1714	-14.2
COE (\$/MWh)	58.6	69.9	-16.2
CO ₂ emission (kg/kWh)	0.081	0.096	-15.6

Table 9 - 10Performance and cost improvement of the IGCC capture plant using
ITM and GE 7H turbine

As with all cases involving advanced technologies, full-scale demonstrations and commercialization are needed to verify the performance and cost assumptions employed in this analysis. For example, while the H-class gas turbine is already offered commercially, its design performance and cost when fired with syngas or hydrogen-rich fuel gas (rather than natural gas) remain to be demonstrated and determined reliably. Similarly, the scale-up and application of the ITM process to an IGCC also remains to be demonstrated. Hence, the uncertainty characteristics discussed earlier in Chapter 8 apply equally well to the advanced technologies discussed here.

REFERENCES (CHAPTER 9)

- 1. Air Products, 2002: Method for Predicting Performance of an Ion Transport Membrane Unit-Operation, Advanced Gas Separation Technology, Allentown, Pennsylvania
- 2. Air Products, 2003: The Development of ITM Oxygen Technology for Integration in IGCC and Other Advanced Power Generation Systems
- 3. Air Products, 2004: ITM Oxygen for Gasification presented at: Gasification Technologies 2004 Washington, D.C. 3-6 October 2004
- 4. Brdar R.D. and Jones R.M., 2000, GE IGCC Technology and Experience with Advanced Gas Turbines, GER-4207
- 5. Matta R.K., Mercer G.D., and Tuthill R.S., 2000: Power Systems for the 21st Century –"H" Gas Turbine Combined-Cycles, GER-3935B
- 6. Mathieu P., 2002: Private communication
- O'Brien J.N., Blau J., Rose M., 2004: An Analysis of the Institutional Challenges to Commercialization and Deployment of IGCC Technology in the U.S. Electric Industry: Recommended Policy, Regulatory, Executive and Legislative Initiatives, Final Report prepared for U.S. Department of Energy, National Energy Technology Laboratory, Gasification Technologies Program and National Association of Regulatory Utility Commissioners
- 8. Prasas R., Chen, J., Hassel, B., San, 2002: OTM-an advanced oxygen technology for IGCC, Francisco, Oct 30, 2002, Gasification conference
- Richards, R.E., 2001, Development of ITM Oxygen Technology for Integration in IGCC & Other Advanced Power Generation Systems (ITM Oxygen), Technical Progress Report for the period January – March 2001 for DOE
- 10. Stiegel, G.J., 2005: Overview of Gasification Technologies Global Climate and Energy Project, Advanced Coal Workshop, March 15, 2005
- 11. Smith R.W., Polukort P., Maslak C.E., Jones C.M., and Gardiner B.D., 2001: Advanced Technology Combined Cycles, GER-3936A
- 12. Todd D.M., 2002, The Future of IGCC, Gasification 5, Noordwijk, The Netherlands

Chapter 10. CONCLUSION

The main objective of this research has been to perform a technical and economic assessment of IGCC systems with and without CO_2 capture. This chapter summarizes the key points presented in this thesis.

10.1. Model development

Detailed engineering models of IGCC systems with oxygen-blown Texaco quench gasifiers were developed in the Aspen Plus simulation software environment. A previous cost model of this IGCC system developed at Carnegie Mellon was updated with more recent data and coupled to the system performance model. To simulate CO₂ capture, performance models of the water gas shift (WGS) reaction system and the Selexol-based CO₂ capture process were derived using detailed chemical simulations, theoretical analysis, and regression analysis. The cost models of the WGS reaction system and the Selexol process for CO₂ capture were coupled to (dependent upon) the input and output parameters of the corresponding performance models. The CO₂ capture system was incorporated into the IGCC model in Aspen Plus with re-design of heat integration of the whole plant. Since the cost models of IGCC systems with and without CO₂ capture were also linked with the plant performance model, all economic assessments reflected plant design assumptions as well as economic and financial parameters.

The IGCC plant and the Selexol-based CO_2 capture process models were also implemented (in reduced form) in a general power generation modeling framework— IECM. The probabilistic capability of the IECM was then applied to models of IGCC systems with and without CO_2 capture in IECM. Therefore, risk and uncertainty analyses, which are important aspects of technical assessment and policy analysis, could be performed. Thus, these models (performance and cost models in Aspen Plus and IECM) provide an analytic environment and tools for technical and economic assessment of gasification—based energy conversion systems with various CO₂ capture options on a systematic and consistent basis.

10.2. Model applications

As a developing technology, IGCC systems have shown advantages over traditional combustion-based energy conversion technologies in terms of energy efficiency, environmental emissions, and greenhouse gas control. First, the models developed in this thesis were used to investigate the factors influencing the technical and economic performance of greenfield IGCC systems with various CO₂ capture options. Then the technical feasibility and economic cost of repowering old PC power plants by IGCC technology were investigated. The uncertainties associated with IGCC systems and with the CO₂ capture process were also studied. Finally, the models were extended to include some emerging novel technologies, and used to assess the potential performance and economic improvement of advanced IGCC systems in the near future.

10.2.1. Greenfield IGCC plants

Through case studies of an IGCC plant using an oxygen-blown Texaco quench gasifier, the effect of factors including CO_2 capture efficiency, coal type, plant size, and capital structure were studied. Four coals, representing bituminous, sub-bituminous, and lignite coals, were used to investigate the effects of coal quality on the performance and cost of IGCC systems with and without CO_2 capture. Although the Texaco gasifier modeled in this study is able to gasify all coals regardless of coal rank, caking characteristics, or amount of coal fines, the coal rank was found to significantly influence

the gasification efficiency, the thermal efficiency and capital cost of an IGCC power plant. The water slurry feeding mechanism used in this plant design resulted in high (nonoptimal) total water input when utilizing low rank coal, like lignite. The relatively high feed rates of coal and high oxygen requirements to maintain gasifier temperatures resulted in increased capital cost and auxiliary power consumption relative to the nominal plant design using bituminous coal.

The effect of different CO_2 capture efficiencies on the power consumption, thermal efficiency, capital cost, cost of electricity, and CO_2 avoidance cost were studied. It was found that the avoidance cost reaches the lowest point when the total CO_2 removal efficiency is in the range of 85%-90%. This indicates that the optimal CO_2 capture efficiency is also in this range.

Generally, the size of a plant will influence its performance and cost because a relatively large plant will benefit from an economy of scale and higher efficiency. Three IGCC systems with one, two, and three GE 7FA gas turbines, respectively, were studied to show the influence of plant size on IGCC systems and CO₂ capture. The plant size has notable influence on the total capital requirement. The capital requirement of the biggest plant is about 280 \$/kW less than that of the smallest one. Thermal efficiency also improves slightly with an increase of the plant size. The efficiency of the biggest plant is about 0.5 percentage points higher than that of the smallest one. The CO₂ avoidance cost decreases slightly with an increase of the plant size. The avoidance cost of the biggest plant is approximately \$2/tonne lower than that of the smallest one.

One of the major issues hindering the application of IGCC is difficulty with plant financing, since IGCC is perceived to be a riskier technology for power generation than conventional combustion-based systems. Six different capital structures were studied to investigate the influence of capital structures on the economic competitiveness of IGCC power plants. Simulation results showed that the total capital requirement of the IGCC plant reduces with an increase of the debt-to-equity ratio. Further study showed that without an incentive financing approach, the IGCC power plant without CO_2 capture is less competitive than PC and NGCC power plants in terms of both the total capital requirement and the cost of electricity (COE). An incentive financing policy for IGCC power plants, like the 3-Party Covenant analyzed in this thesis, can improve the competitive ability of IGCC power plants and accelerate their commercialization. Due to the advantages of IGCC for CO_2 capture, additional analysis showed that if CO_2 capture is required for power generation processes, IGCC plants without an incentive capital structure would be competitive with PC and NGCC plants.

10.2.2. IGCC Repowering

In this thesis, two simulation models were set up in Aspen Plus to evaluate the repowering cases. One model simulated the most restrictive condition, replacing the existing boiler with a gasifier, a gas turbine and HRSG, and no modification to the steam turbine and the feedwater system. Another model simulated the most favorable condition, in which the steam turbine had sufficient design margins so that it could be incorporated into a three-pressure reheat cycle with no feedwater heating. The two models were further revised to incorporate the CO_2 capture function. Simulation results showed that IGCC repowering is less capital intensive and has a shorter construction period, but the

feasibility of repowering is very site specific. Under suitable conditions, IGCC repowering with or without CO_2 capture may be an economically attractive option for existing steam power units. Repowering also provides an option for introducing new power generation technology with lower risk to utilities.

10.2.3. Uncertainty analysis of IGCC systems

The uncertainty and variability in IGCC systems with CO2 capture come from the limited experience in producing, constructing and operating IGCC plants with CO2 capture. This study investigated the influence of uncertainties and variability associated with plants and process designs on the capital cost, cost of electricity and CO2 avoided avoidance. After preliminary screening, this thesis focused on investigating the uncertainties of two key parameters, capacity factor and fuel price. Considering the effect of uncertainties of capacity factor and coal price, there was a significant possibility that the COE of an IGCC capture plant could be higher than that of the deterministic estimates found in many recent studies.

10.2.4. IGCC with advanced technologies

There are substantial R&D programs in the U.S. to improve the efficiency and cost effectiveness of IGCC technology. This thesis studied the effects on performance and cost of IGCC systems for two emerging advanced technologies—the ion transport membrane (ITM) for oxygen production and the GE H-class gas turbine for power generation.

The net efficiency of an IGCC system using ITM increased approximately 2 percentage points, while the total capital requirement of the IGCC plant fell from \$1311

to \$1240. It was also found that GE H-class gas turbines could significantly improve the performance of IGCC power plants. Compared to IGCC plants using GE F-class turbines, the utilization of GE H turbines can increase the net efficiency by more than 4 percentage points, reduce the capital cost per kilowatt by more than 4%, and reduce the cost of electricity by more than 7%.

This preliminary analysis found that an IGCC plant employing both ITM and GE 7H gas turbines could achieve a thermal efficiency as high as 42.3% on a higher heating value basis, and that the total capital requirement and the cost of electricity of such an IGCC plant would be slightly lower than those of current PC power plants. It was also predicted that the energy efficiency of such an IGCC system with CO₂ capture is approximately 38.2%, which is higher than that of current IGCC systems without CO₂ capture. The total capital cost of the next generation IGCC systems with CO₂ capture would be about 1470 \$/kW, which would be able to compete with current IGCC systems without CO₂ capture.

10.3. Some considerations about future work

This work may be furthered in several directions. First, the IGCC plant models could be developed with greater consideration of optimal system design, especially when the CO_2 capture process is incorporated into IGCC plants. Studies of the optimal heat recovery scheme, the optimal operation pressure and temperature of the gasifier and HRSG, the optimal integration of oxygen production and the gas turbine system all would provide a better understanding of the potential advantages of IGCC systems with CO_2 capture.

Second, the models for the water gas shift reaction system and Selexol-based CO_2 capture process could be refined with the availability of additional data. Another direction of future work might be extending the CO_2 capture process with different commercial solvents, such as Rectisol and Ucarsol.

The IGCC models also could be extended by incorporating more technology options. For instance, the models of IGCC systems with different gasifier types, such as the Shell gasifier and E-Gas system. Different syngas cleanup processes, such as high temperature cleanup processes, advanced CO_2 capture systems, and options for combined capture and sequestration of CO_2 and H_2S could be studied to provide a more comprehensive set of options for technical and economic assessment of CO_2 capture from IGCC systems.

Finally, additional analyses of uncertainties and their effect on performance and cost estimates could be carried out, especially for many of the advanced technologies currently under development.

APPENDIX A. CO CONVERSION EFFICIENCY OF THE WGS REACTION

Recall the WGS reaction equation given by:

$$CO + H_2O \Leftrightarrow CO_2 + H_2$$

Using the definition of the chemical equilibrium constant, the chemical equilibrium constant (K) for the water gas shift reaction can be given by,

$$\mathbf{K} = \frac{[\mathrm{CO}_2][\mathrm{H}_2]}{[\mathrm{CO}][\mathrm{H}_2\mathrm{O}]} \tag{A-1}$$

where [i] = the molar concentration of species i at the equilibrium state.

Substituting Eq. 4-3, 4-4, 4-5, and 4-6 into Eq. A-1, then the equilibrium constant in the high temperature reactor (K_h) can be expressed as,

$$K_{h} = \frac{([CO_{2}]_{0} + \xi_{h}[CO]_{0})([H_{2}]_{0} + \xi_{h}[CO]_{0})}{([CO]_{0} - \xi_{h}[CO]_{0})([H_{2}O]_{0} - \xi_{h}[CO]_{0})}$$
(A-2)

Treating the CO conversion in the high temperature reactor (ξ_h) as an unknown parameter, noting that the above equation is parabolic in ξ_h , and then solving Eq. A-2, the CO conversion at the high temperature reactor can be obtained and given by:

$$\xi_{\rm h} = \frac{u_1 - \sqrt{u_1^2 - 4w_1 v_1}}{2w_1} \tag{A-3}$$

where,

$$u_1 = K_h([CO]_0 + [H_2O]_0) + ([CO_2]_0 + [H_2]_0)$$

$$v_{1} = K_{h} ([CO]_{0} [H_{2}O]_{0}) - ([CO_{2}]_{0} [H_{2}]_{0})$$

$$w_{1} = K_{h} - 1$$
(A-4)

In a similar way, the total CO conversion (ξ_{total}) in the two reactors is given by

$$\xi_{tot} = \frac{u_2 - \sqrt{u_2^2 - 4w_2 v_2}}{2w_2} \tag{A-5}$$

where,

$$u_{2} = K_{l}([CO]_{0} + [H_{2}O]_{0}) + ([CO_{2}]_{0} + [H_{2}]_{0})$$

$$v_{2} = K_{l}([CO]_{0}[H_{2}O]_{0}) - ([CO_{2}]_{0}[H_{2}]_{0})$$

$$(A-6)$$

$$w_{2} = K_{l} - 1$$

Furthermore, the equilibrium constants in the high temperature and low temperature reactors can be calculated as a function of temperature as follows (Davis, 1980):

$$K_h = \exp(\frac{8240}{T_h + dT_h + 459.67} - 4.33)$$
(A-7)

$$K_{l} = \exp(\frac{8240}{T_{l} + dT_{l} + 459.67} - 4.33)$$
(A-8)

Where K_h and K_l are the equilibrium constants of shift reaction in the high and low temperature reactors with taking into account of the approach temperatures; dT_h and dT_l are the Aspen approach temperatures (F) for the high and low temperature reactors, respectively. T_h and T_l are the reaction equilibrium temperatures (F), which are the final temperatures when the WGS reaction reaches equilibrium states, at the high and low temperature reactors, respectively. The two temperatures, T_h and T_l , can be calculated using the following regression equations based on ASPEN Plus simulations,

$$T_{h} = 0.0122 P_{0} + 0.8668 T_{0} + 3297 .049 [CO_{2}]_{0} - 21.634 [H_{2}O]_{0} + 356 .234 [H_{2}]_{0} + 401 .392 [N_{2}]_{0} + 2290 .608 [CO]_{0} [H_{2}O]_{0}$$

(R2=0.99) (A-9)

$$T_{1} = -0.00136 P_{0} + 0.1031 T_{0} + 16608 .87[CO_{2}]_{0} + 404.098[H_{2}O]_{0}$$

+ 331.976[H_{2}]_{0} + 258.772[N_{2}]_{0} - 1198.036[CO]_{0}[H_{2}O]_{0} - 2105.116[CO_{2}]_{0}[H_{2}]_{0}

Here T_0 and P_0 are the temperature (F) and pressure (psia) of the syngas fed into the high temperature reactor. $[i]_0$ is the molar fraction of syngas composition i before fed into the high temperature reactor.

According to the definition of CO conversion, the CO conversion in the low temperature reactor is calculated from the CO conversion in the high temperature (ξ_h) and the total CO conversion (ξ_{tot}) as in the following equation.

$$\xi_1 = 1 - \frac{1 - \xi_{\text{total}}}{1 - \xi_h} \tag{A-11}$$

APPENDIX B. METHODOLOGY FOR CALCULATING THE CATALYST VOLUME OF THE WGS REACTION

The volumes of catalyst, either clean shift catalysts or sour tolerance shift catalysts, can be calculated as in the following steps.

The catalyst volume (V_{cat}, ft^3) can be given by,

$$V_{cat.} = \frac{VF}{SV}$$
(B-1)

where SV is the space velocity (1/hr); VF is the volumetric flow rate of syngas (ft^3/hr) .

For a well mixed, constant volume batch reactor, the space velocity is related to the fraction conversion of reactant (x) and the reaction rate (r) by the following equation (Polyanin, 2002):

$$SV^{-1} = \int_{0}^{x} \frac{dx}{r}$$
(B-2)

where x is the fraction conversion of CO to CO_2 .

The reaction rate of the WGS reaction can be given by (Doctor, 1994),

$$r = k \left[([CO]_0 - x)([H_2O]_0 - x) - \frac{([CO_2]_0 + x)([H_2]_0 + x)}{K} \right]$$
(B-3)

where k is the reaction rate constant of the water gas shift reaction; $[CO]_0$, $[H_2O]_0$, $[CO_2]_0$, $[H_2]_0$ are the inlet molar concentration of CO, H₂O, CO₂ and H₂, respectively; K is the equilibrium constant of the water gas shift reaction at the equilibrium temperature.

Equation (B-3) can be substituted into Equation (B-2) to produce the following equation,

$$\frac{k}{SV} = \frac{2K}{\sqrt{q}} \left\{ \ln\left[\frac{2wx - u - \sqrt{q}}{2wx - u + \sqrt{q}}\right] - \ln\left[\frac{-u - \sqrt{q}}{-u + \sqrt{q}}\right] \right\}$$
(B-4)

where: w = K - 1

$$\sqrt{q} = \sqrt{u^2 - 4wv}$$

$$u = K([CO]_0 + [H_2O]_0) + ([CO_2]_0 + [H_2]_0)$$

$$v = K([CO]_0[H_2O]_0) - ([CO_2]_0[H_2]_0)$$
(B-5)

Using the above equations, the volume of high temperature catalyst $(V_{cat,h})$ is given by:

$$V_{\text{cat.,h}} = \frac{VF_{\text{h}}}{SV_{\text{h}}}$$
(B-6)

where

$$\frac{k_{h}}{SV_{h}} = \frac{K_{h,real}}{\sqrt{q_{1}}} \left\{ ln[\frac{2w_{1}x_{1} - u_{1} - \sqrt{q_{1}}}{2w_{1}x_{1} - u_{1} + \sqrt{q_{1}}}] - ln[\frac{-u_{1} - \sqrt{q_{1}}}{-u_{1} + \sqrt{q_{1}}}] \right\}$$
(B-7)

$$\sqrt{\mathbf{q}_{1}} = \sqrt{\mathbf{u}_{1}^{2} - 4\mathbf{w}_{1}\mathbf{v}_{1}}$$
$$u_{1} = K_{h,real}([CO]_{0} + [H_{2}O]_{0}) + ([CO_{2}]_{0} + [H_{2}]_{0})$$

$$v_{1} = K_{h,real} ([CO]_{0}[H_{2}O]_{0}) - ([CO_{2}]_{0}[H_{2}]_{0})$$

$$w_{1} = K_{h,real} - 1$$
(B-8)

Recalling the equilibrium constant of the water gas shift reaction in Eq. (A-7), then the equilibrium constant $K_{h,real}$ calculated based on the equilibrium reaction temperature in the high temperature reactor can be given by,

$$K_{h,real} = \exp(\frac{8240}{T_h + 459.67} - 4.33)$$
(B-9)

In a similar way, the volume of the low temperature catalyst ($V_{cat,l}$) is given by,

$$V_{cat,l} = \frac{VF_l}{SV_l} \tag{B-10}$$

where

$$\frac{k_{l}}{SV_{l}} = \frac{K_{l,real}}{\sqrt{q_{2}}} \left\{ \ln\left[\frac{2w_{2}x_{2} - u_{2} - \sqrt{q_{2}}}{2w_{2}x_{2} - u_{2} + \sqrt{q_{2}}} - \ln\left[\frac{-u_{2} - \sqrt{q_{2}}}{-u_{2} + \sqrt{q_{2}}}\right] \right\}$$

$$\sqrt{q_{2}} = \sqrt{u_{2}^{2} - 4w_{2}v_{2}}$$

$$u_{2} = K_{l,real} ([CO]_{0} + [H_{2}O]_{0} - 2x_{1}) + ([CO_{2}]_{0} + [H_{2}]_{0} + 2x_{1})$$

$$v_{2} = K_{l,real} [([CO]_{0} - x_{1})([H_{2}O]_{0} - x_{1})] - [([CO_{2}]_{0} + x_{1})([H_{2}]_{0} + x_{2})]$$
(B-11)

$$w_2 = K_{l,real} - 1 \tag{B-12}$$

Here, and the equilibrium constant $K_{l,real}$ is calculated based on the equilibrium reaction temperature in the low temperature reactor as:

$$K_{l,real} = \exp(\frac{8240}{T_l + 459.67} - 4.33)$$
(B-13)

For the iron-based catalyst, the reaction rate constant (k) is given by [Doctor, 1994],

$$\log(\frac{k}{A_p}) = 6.947 - \frac{3830}{0.85T_h + 0.15T_0 + 459.67}$$
(B-14)

where T_h is the equilibrium temperature in the high temperature reactor (F); T_0 is the syngas temperature at the inlet of the high temperature reactor (F).

For the copper-based catalyst, the reaction rate constant (k) is given by [Campbell, 1970]:

$$\log(\frac{k}{A_p}) = 6.91 - \frac{3062}{0.85 \cdot T_l + 0.15 \cdot T_{l,i} + 459.67}$$
(B-15)

where T_l is the equilibrium temperature in the low temperature reactor (F); T_{li} is the syngas temperature at the inlet of the low temperature reactor (F).

Here A_p is pressure-dependent activity factors, which can be given by [Doctor 1994],

$$P \le 400 \, psig \ , A_p = 4 \cdot 10^{-8} \, P^3 + 10^{-5} \, P^2 + 0.0092P + 0.9984$$
$$P \succ 400 \, psig \ , A_p = 4 \tag{B-16}$$

For a cobalt-based catalyst, the catalyst reaction rate constant is given by the following equation, which is regressed using the published data [Park, 2000]:

$$R\ln k = 119.75 + \frac{10^7}{T^2} - \frac{95702}{T}$$
(R²=0.996) (B-17)

where R is the gas constant; T is the reaction temperature in the reactor.

REFERENCES (APPENDIX B)

- 1. Campbell, J. S., "Influences of Catalyst Formulation and Poisoning on the Activity and Die-Off of Low Temperature Shift Catalysts", Ind. Eng. Chem. Proc. Des. Dev., 9(4), 1970, pp. 588-595.
- 2. A. D. Polyanin, A. M. Kutepov, et al., Hydrodynamics, Mass and Heat Transfer in Chemical Engineering, Taylor & Francis, London, 2002.
- Doctor R.D., Molburg, J.C. Thimmapuram, P., Berry, G.F., and Livengood, D.C. "Gasification Combined Cycle: Carbon Dioxide Recovery, Transport Disposal," ANL/ESD-24, Argonne National Laboratory (Sept. 1994)
- Park J.N., Kim, J. H., Lee, H. I., A study on the Sulfur-Resistant Catalysts for Water Gas Shift Reaction IV. Modification of CoMo/-Al₂O₃ Catalyst with K, Korean Chem. Soc. Vol.21, No.12 1239, (Oct. 2000)

APPENDIX C. CALCULATION PROCESS OF THE WGS REACTION SYSTEM

This section illustrates the calculation process of the performance and cost model of the WGS reaction system discussed in Chapter 4 through a case study. The input parameters are as follows.

Syngas to high temperature shift reactor	Molar concentration	Flow rate (lb-mol/hr)	Volume (cft/hr)
H_2S	0.000	0.029	0.468
CH ₄	0.033	9.849	160.594
Ar	0.343	101.302	1656.494
COS	0.000	0.001	0.016
NH ₃	0.000	0.043	0.699
N ₂	0.429	126.627	2070.685
СО	19.841	5852.992	95708.543
Н2О	60.179	17752.435	255451.049
CO ₂	4.197	1238.133	20003.158
H ₂	14.976	4417.884	71715.234
total	100.000	29499.294	446766.941
total flow rate (lb-mol/hr)	29499.295		
Temperature(F)	450		
Pressure(psia)	610		

Table C - 1Input parameters of the WGS model

The first step is to calculate the CO conversion efficiency in the high and low temperature reactor, which are given in Table C-2, and C-3.

	1	
High temperature reactor	Value	Equations used
Equilibrium temperature at HT reactor (F)	851.454	(A-9)
Equilibrium constant at HT reactor temperature	7.062	(A-7)
Equilibrium constant at HT reactor temperature taking into account approach temperature	6.278	(B-9)
		• /
Middle variable u1	5.278	
Middle variable v1	-5.216	
Middle variable w1	0.743	(A-4)
CO conversion efficiency in HT reactor	0.870	(A-3)
CO molar concentration change in HT reactor	0.173	

Table C - 2Calculation of the CO conversion efficiency in the high temperature
reactor

Table C - 3	Calculation of the CO conversion efficiency in the low temperature
	reactor

Low temperature reactor	Value	Equations used
Equilibrium temperature at LT reactor (F)	575.040	(A-10)
Equilibrium constant at LT reactor temperature	37.848	(A-8)
Equilibrium constant at LT reactor temperature taking into account approach temperature	33.777	(B-13)
Middle variable u2	32.777	(A-6)
Middle variable v2	-27.220	
Middle variable w2	4.027	
Total CO conversion efficiency	0.971	(A-5)

With the CO conversion efficiency, the compositions of the syngas from the high and low temperature reactors are given by Table C-4, and C-5.

Syngas from HT reactor	Molar concentration	Flow rate (lb-mol/hr)	Volume (cft/hr)	Equation used
H ₂ S	0.000	0.029	0.677	
CH ₄	0.033	9.849	232.447	
Ar	0.343	101.302	2395.548	
COS	0.000	0.001	0.023	
NH ₃	0.000	0.043	1.011	
N ₂	0.429	126.627	2994.502	
СО	2.570	758.105	17927.399	(4-3)
H ₂ O	42.908	12657.548	274474.925	(4-6)
CO_2	21.468	6333.020	149086.154	(4-5)
H ₂	32.247	9512.770	223837.105	(4-4)
total	100.000	29499.294	670949.793	
total flow rate (lb- mol/hr)	29499.295			
Temperature(F)	851			(A-9)
Pressure(psia)	604			

Table C - 4Syngas compositions from the high temperature reactor

Syngas from HT reactor	Molar concentration	Flow rate (lb-mol/hr)	Volume (cft/hr)	Equation used
H ₂ S	0.000	0.029	0.547	
CH ₄	0.033	9.849	187.747	
Ar	0.343	101.302	1935.779	
COS	0.000	0.001	0.019	
NH ₃	0.000	0.043	0.817	
N ₂	0.429	126.627	2419.791	
СО	0.581	171.405	3275.385	(4-7)
H ₂ O	40.919	12070.848	206971.530	(4-10)
CO ₂	23.457	6919.720	131153.762	(4-9)
H ₂	34.236	10099.471	191794.246	(4-8)
total	100.000	29499.294	537739.622	
total flow rate (lb- mol/hr)	29499.295			
Temperature(F)	575.040			(A-10)
Pressure(psia)	591.882			

Table C - 5Syngas compositions from the low temperature reactor

Table C-6 and C-7 give the calculation steps of the catalyst volumes and the process facility costs of the shift reactors.

Parameter	Value	Equation used
Middle variable q1^0.5	3.721	(B-8)
Middle variable u1	5.843	
Middle variable v1	0.837	
Middle variable w1	6.062	
Pressure-dependent activity		
factors	4.000	(B-16)
Reaction reat in HT reactor	12584.716	(B-14)
SV in HT reactor	1625.458	(B-7)
Volume of HT catalyst	412.776	(B-6)
Volume of HT reactor	495.331	
PFC of HT reactor	698.167	(4-23)
Initial HT catalyst cost	20.639	

Table C - 6Calculation of catalyst volume and PFC of the high temperature reactor

Parameter	Value	Equation used
Middle variable q2^0.5	16.240	(B-12)
Middle variable u2	17.750	
Middle variable v2	0.348	
Middle variable w2	36.848	
Pressure-dependent activity factors	4.000	(B-16)
Reaction reat in LT reactor	23039.292	(B-15)
SV in LT reactor	2829.039	(B-11)
Volume of LT catalyst	67.795	(B-10)
Volume of LT reactor	81.354	
PFC of LT reactor	277.444	(4-23)
Initial LT catalyst cost	16.949	

The final step is to calculate the process facility costs of the heat exchangers, which is given by Table C-8.

Parameter	Value	Equation used
Heat exchanger 1		
Hot gas inlet T, F	851.454	(A-9)
Hot gas outlet T	450.000	Design value
Cold fluid inlet T	351.463	(4-20)
Cold fluid outlet T	580.153	(4-11)
Heat released by syngas, Btu/hr	53917730.708	(4-12), (4-13)
Heat released by syngas, kW	15797.895	
Log mean temperature difference, C	94.768	
PFC of heat exchanger 1	489.104	(4-24)
Heat exchanger 2		
Hot gas inlet T, F	575.040	(A-10)
Hot gas outlet T	381.463	(4-17)
Cold fluid inlet T	57.000	Design value
Cold fluid outlet T	400.000	Design value
Heat released by syngas, Btu/hr	1.047E+08	(4-15), (4-16)
Heat released by syngas, kW	30687.447	
Log mean temperature difference, C	134.508	
PFC of heat exchanger 2	4605.725	(4-25)
Heat exchanger 3		
Hot gas inlet T, F	381.463	(4-17)
Hot gas outlet T	100.000	Design value
Cold fluid inlet T	57.000	Design value
Cold fluid outlet T	351.463	(4-20)
Heat released by syngas, Btu/hr	2.416E+08	(4-18), (4-19)
Heat released by syngas, kW	70801.745	
Log mean temperature difference, C	20.062	
PFC of heat exchanger 3	3878.719	(4-26)

Table C - 8Process facility costs of the heat exchangers

APPENDIX D. CALCULATION PROCESS OF THE SELEXOL SYSTEM FOR CO_2 CAPTURE

This section illustrates the calculation process of the performance and cost model for the Selexol system discussed in Chapter 5 through a case study. The input parameters are as follows.

Table D - 1 Selexol properties for calculation Specific Specific Vapor Specific CO_2 Number pressure heat volume Viscosity gravity Mole solubility of @25C,cp @25C, weight @25C, @25C SCF/US commerci (gallon/lbkg/m^3 mmHg Btu/lb F gal @25C al plants mol) 5.8 1030 280 0.00073 0.49 0.485 32 32.574146

	1	20		. 0					
Gas	CO ₂	H ₂	CH ₄	СО	H_2S	COS	Ar	NH ₃	N ₂
Relativly Solubility	1.000	0.013	0.067	0.028	8.930	2.330	0.000	4.870	0.000
(scf/lb-mol)	377.05	379.50	378.46	379.17	376.08	374.53	379.01	375.88	379.23
Cp	0.199	3.425	0.593	0.248	0.245		0.125	0.520	0.249
(Btu/lb F)									
Cv		2.440	0.450	0.172				0.399	0.176
(Btu/lb F)	0.152								
K=Cp/Cv	1.310	1.400	1.320	1.410	1.310	1.310	1.600	1.310	1.400
Mole weight	44.010	2.016	16.043	28.010	34.082	60.074	39.948	17.031	28.013
Solution heat (Btu/lb)	160.000		75.000		190.000				

Table D - 2Properties of gases in syngas

Spacias	Syngas fed	into heater exc	hanger	Syngas out heater exchanger		
Species	Molar conc.	flow rate (lb-mol/hr)	Volume (cft/hr)	Molar conc.	flow rate (lb-mol/hr)	Volume (cft/hr)
H_2S	0.02	2.77	29.49	0.02	2.77	28.68
CH ₄	0.06	7.83	83.31	0.06	7.83	81.02
Ar	0.41	56.58	604.45	0.41	56.58	587.89
COS	0.01	1.39	14.81	0.01	1.39	14.40
NH ₃	0.00	0.00	0.00	0.00	0.00	0.00
N2	0.68	94.25	1006.95	0.68	94.25	979.36
СО	0.98	136.31	1456.26	0.98	136.31	1416.37
H ₂ O	0.19	26.33	229.67	0.19	26.33	221.96
CO ₂	33.33	4619.95	47871.78	33.33	4619.95	46479.79
H ₂	64.32	8914.59	94165.06	64.32	8914.59	91555.39
total	100	13860	145462	100	13860	141365
Temperature (F)	70	70	70	55	55	55
Pressure (psia)	550	550	550	545	545	545

Table D - 3Composition of syngas before CO2 capture

Table D - 4Assumption for power consumption calculation

Efficiency of power recovery turibne	Efficiency of compressor for Selexol compression	Efficiency of recycle gas compressor	Efficiency of CO ₂ compressor	Efficiency of three stage CO ₂ compressor	Evaporation temperature of refrigeration(F)
0.70		0.02	0.02	0.02	10
0.78	0.8	0.82	0.82	0.82	10

Table D - 5	CO ₂ capture	efficiency an	d flashing tanl	k pressures
-------------	-------------------------	---------------	-----------------	-------------

CO ₂ capture efficiency	Flashing tank 1	Flashing tank	Flashing tank3
	pressure (psia)	pressure 2 (psia)	pressure (psia)
0.97	60	14.7	4

The calculation processes and output parameters of the Selexol system are as follows.

extra selexol ratio	1.5	Eq. (5-11)
CO ₂ capture in absorber (SCF/hr)	1689702.3	
CO ₂ solution heat(Btu/hr)	31555552.2	Eq. (5-7)
Heat released from syngas	1594624.8	Eq. (5-6)

Table D - 6CO2 capture amount required by capture efficiency

Table D - 7Calculating the flow rate of Selexol solvent

Estimated Selexol flow rate(lb-mol/hr)	Estimated Selexol temperature increase due to CO ₂ solution heat	Estimated Selexol temperature increase due to syngas cooling	Estimated CO_2 in the lean Selexol at the last stage (SFC Selexol	
10578.96	21.74	1.10	112203.81	
Adjusted Selexol flow rate(lb- mol/hr)	Adjusted Selexol temperature increase due to CO ₂ solution heat	Adjusted Selexol temperature increase due to syngas cooling	Estimated CO_2 in the lean Selexol at the last stage (SFC Selexol	
9486.24	24.25	1.23	100457.65	
Adjusted Selexol flow rate(lb- mol/hr)	Adjusted Selexol temperature increase due to CO ₂ solution heat	Adjusted Selexol temperature increase due to syngas cooling	Estimated CO_2 in the lean Selexol at the last stage (SFC Selexol	
9859.17	23.33	1.18	104466.45	
Adjusted Selexol flow rate(lb- mol/hr)	Adjusted Selexol temperature increase due to CO ₂ solution heat	Adjusted Selexol temperature increase due to syngas cooling	Estimated CO_2 in the lean Selexol at the last stage (SFC Selexol	
9717.09	23.67	1.20	102939.17	
Adjusted Selexol flow rate(lb- mol/hr)	Adjusted Selexol temperature increase due to CO ₂ solution heat	Adjusted Selexol temperature increase due to syngas cooling	Estimated CO ₂ in the lean Selexol at the last stage (SFC Selexol	Eq (5-3) ~
9769.18	23.54	1.19	103499.12	Eq. (5-12)

Table D – 7 continued

Adjusted Selexol flow rate(lb- mol/hr)	Adjusted Selexol temperature increase due to CO ₂ solution heat	Adjusted Selexol temperature increase due to syngas cooling	Estimated CO_2 in the lean Selexol at the last stage (SFC Selexol	
9749.80	23.59	1.19	103290.82	Eq (5-3) ~
Final Selexol flow rate(lb- mol/hr)	Final Selexol temperature increase due to CO_2 solution heat	Final Selexol temperature increase due to Syngas cooling	Estimated CO_2 in the lean Selexol at the last stage (SFC Selexol	Eq. (5-12)
9756.97	23.57	1.19	103367.90	

Table D - 8Calculation of the operating pressure of sump tank

Operating pressure of sump tank (psia)	180.322	
H2 in recycle gas (SCF/hr)	29179.53	E_{a} (5.14)
CO_2 in recycle gas	27737.39	$\sim Eq.(5-17)$

Table D - 9	Calculation the temperature change of Selexol due to CO ₂ release from
	flash tank

	Adjusted amount of CO ₂ captured	Estimated heat(Btu) absorbed due to CO ₂ released from Selexol	Selexol temperature decrease
SFC/hr	1074956.316	17036816	12.7268
lb-mol/hr	2850.952396		
	Adjusted amount of CO ₂ captured	Estimated heat(Btu) absorbed due to CO ₂ released from Selexol	Selexol temperature decrease
SFC/hr	1033634.983	13410930	10.0182
lb-mol/hr	2741.361752		
	Adjusted amount of CO ₂ captured	Estimated heat(Btu) absorbed due to CO ₂ released from Selexol	Selexol temperature decrease
SFC/hr	1042429.258	14182615	10.59466
lb-mol/hr	2764.685547		

Table D – 9 continued

	Adjusted amount of CO ₂ captured	Estimated heat(Btu) absorbed due to CO ₂ released from Selexol	Selexol temperature decrease
SFC/hr	1040557.603	14018380	10.47197
lb-mol/hr	2759.721625		
	Final amount of CO_2 captured	Final heat(Btu) abosrbed due to CO ₂ released from Selexol	Final Selexol temperature decrease
SFC/hr	1040955.941	14053333	10.49808
lb-mol/hr	2760.778079		

 Table D - 10
 Gases retained in the solvent at the flash tank 1

Amount of gases captured in Selexol in first flash tank	CO ₂	H ₂	CH ₄	СО	H_2S
SFC/hr	1040955.941	674.496	14.690	47.031	951.506
lb-mol/hr	2760.778	1.777	0.039	0.124	2.530

Table D - 11Gases released from the solvent at the flash tank 1

Amount of gases released from first flash tank	CO ₂	H ₂	CH ₄	СО	H_2S
SFC/hr	752114.26	44303.92	188.06	1434.28	90.98
lb-mol/hr	1994.72	116.74	0.49	3.78	0.24
volume flow(cubic feet/hr)	180304.10	10551.23	44.93	342.25	21.87

Table D - 12Gases retained in the solvent at the flash tank 2

Amount of gases captured in Selexol in second flash tank	CO ₂	H ₂	CH ₄	СО	H_2S
SFC/hr	293983.24	84.01	1.92	5.94	586.35
lb-mol/hr	779.68	0.22	0.005	0.015	1.55

Amount of gases released from second flash tank	CO ₂	H ₂	CH ₄	СО	H_2S
SFC/hr	746972.69	590.48	12.76	41.08	365.14
lb-mol/hr	1981.08	1.55	0.03	0.11	0.97
volume flow(cubic feet/hr)	714322.91	561.06	12.159	39.07	350.10

Table D - 13Gases released in the solvent at the flash tank 2

Table D - 14Gases retained in the solvent at the flash tank 3

Amount of gases captured in Selexol in third flash tank	CO ₂	H ₂	CH ₄	СО	H_2S
SFC/hr	103367.89	25.38	1.71	5.27	555.45
lb-mol/hr	274.15	0.07	0.004	0.01	1.47

Table D - 15Gases released in the solvent at the flash tank 3

Amount of gases released from third flash tank	CO ₂	H ₂	CH_4	СО	H_2S
SFC/hr	190615.34	58.62	0.22	0.66	30.90
lb-mol/hr	505.54	0.15	0.0006	0.002	0.08
volume flow(cubic feet/hr)	492358.34	162.83	0.60	1.85	86.59

Table D - 16Final product of CO2 from Selexol

	CO ₂	H ₂	CH ₄	СО	H_2S
SFC/hr	1689702.30	44953.03	201.04	1476.03	487.03
lb-mol/hr	4481.35	118.45	0.53	3.89	1.29
volume flow(cubic feet/hr)	433670.46	11463.08	51.43	377.10	125.39

Power recovery in turbine 1(hp)	594.74	Eq. (5-17)
Power recovery in turbine 2(hp)	586.43	Eq. (5-17)
Power consumption of solvent compression(hp)	2105.39	Eq. (5-20)
Selexol temperature increase due to compression	3.74	Eq. (5-21)
Power consumer of refrigeration(hp)	885.54	Eq. (5-22)
Calculation power of recycle gas compressor		
Average k=Cp/Cv	1.39	
Power of recycle gas compressor(hp)	2965.07	Eq. (5-19)
Calcuation of power of CO ₂ compressor in flash tank 2		
Average k=Cp/Cv	1.31	
Power of compressor(hp)	1555.53	Eq. (5-19)
Calcuation of power of CO ₂ compressor in flash tank 3		
Average k=Cp/Cv	1.31	
Power of compressor(hp)	738.34	Eq. (5-19)
Calculation of power of CO ₂ product compressor		
Cold temperature(F) of CO ₂ product	Design value	
Average k=Cp/Cv	1.31	
Power of compressor(hp)	6317.88	Eq. (5-19)
Total energy consumption of Selexol (hp)	13386.61	
total energy consumption (Kw)	9979.72	

Table D - 17Power consumption calculation

Cost model of Selexol		
Cost of absorber per train(k\$)	750.43	Eq. (5-23)
Cost of sump tank (k\$)	155.24	Eq. (5-25)
Cost of recycle compressor (k\$)	2246.19	Eq. (5-26)
Cost of power recovery turbine 1(k\$)	275.75	Eq. (5-24)
Cost of power recovery turbine 2 (k\$)	266.87	Eq. (5-24)
Cost of flashing tank 1(k\$)	107.33	Eq. (5-31)
Cost of flashing tank 2 (k\$)	107.33	Eq. (5-31)
Cost of flashing tank 3 (k\$)	107.33	Eq. (5-31)
Cost of CO ₂ compressor in flash tank 2 (k\$)	1017.81	Eq. (5-28)
Cost of CO ₂ compressor in flash tank 3 (k\$)	614.62	Eq. (5-28)
Cost of CO ₂ product compressor per stage (k\$)	5263.78	Eq. (5-29)
Cost of refrigeration 3 (k\$)	661.31	Eq. (5-30)
Cost of Selexol pump (k\$)	295.29	Eq. (5-27)
Heat exchanger (k\$)	1103.99	Eq. (5-32)

Table D - 18Process facility cost of Selexol process

APPENDIX E. PRELIMINARY DISTRIBUTIONS OF UNCERTAIN PARAMETERS

The basis of the probability distribution of model parameters (Table E-1, Table E-2, and Table E-3) for the preliminary uncertainty screening in Chapter 8 is briefly explained here. As mentioned in Chapter 8, these distributions take into account the data reported in literatures, modeling approximations, and expert's technical judgments. The parameters and their distributions are given in Table E-1, E-2, and E-3.

Table E - 1Distribution functions assigned to the parameters of the IGCC process
(the distribution functions in this table, except the distribution of the
fixed charge factor, mainly come from reference [1] and were updated
with data from reference [2] and [3])

Parameter	Unit	Determinis tic value	Distribution function
Facility cost paramet	ers		
Fixed charged	%	14.8	Triangular(7.1, 14.8, 17.4)
Engineering and home office fee	% of TPC	10	Triangular(7,10,12)
Indirection construction cost factor	% of TPC	20	Triangular(15,20,20)
Project uncertainty	% of TPC	12.5	Uniform(10,15)
General facilities	% of TPC	15	Triangular(10,15,25)
Process contingency			
Oxidant feed	% of PFC	5	Uniform(0,10)
Gasification	% of PFC	10	Triangular(0,10,15)
Selexol	% of PFC	10	Triangular(0,10,20)
Low temperature gas cleanup	% of PFC	0	Triangular(-5,0,5)
Claus plant	% of PFC	5	Triangular(0,5,10)
Beavon-Stretford	% of PFC	10	Triangular(0,10,20)

Tab<u>le E – 1 continued</u>

Parameter	Unit	Determinis tic value	Distribution function
Process condensate treatment	% of PFC	30	Triangular(0,30,30)
Gas turbine	% of PFC	12.5	Triangular(0,12.5,25)
Heat recovery steam generator	% of PFC	2.5	Triangular(0,2.5,5)
Steam turbine	% of PFC	2.5	Triangular(0,2.5,5)
General facilities	% of PFC	5	Triangular(0,5,10)
Maintenance costs			
Gasification	% of TPC	4.5	Triangular(3,4.5,6)
Selexol for sulfur remove	% of TPC	2	Triangular(1.5,2,4)
Low temperature gas cleanup	% of TPC	3	Triangular(2,3,4)
Claus plant	% of TPC	2	Triangular(1.5,2,2.5)
Boiler feed water	% of TPC	2	Triangular (1.5, 2, 4)
Process condensate treatment	% of TPC	2	Triangular(1.5,2,4)
Gas turbine	% of TPC	1.5	Triangular(1.5,1.5,2.5)
Heat recovery steam generator	% of TPC	2	Triangular (1.5, 2, 4)
Steam turbine	% of TPC	2	Triangular(1.5,2,2.5)
Other fixed operating	g cost parameters		
Labor rate	\$/hr	25	Triangle (17,25,32)
Variable operating co	ost parameters		
Ash disposal	\$/ton	10	Triangular(10,10,25)
Sulfur byproduct	\$/ton	75	Triangular(60,75,125)

		Nominal	
Performance parameter	Unit	value	Distribution function
Mole weight of Selexol	lb/mole	280	Triangular(265,280,285)
Pressure at flash tank 1	Psia	60	Uniform(40,75)
Pressure at flash tank 2	Psia	20	Uniform(14.7,25)
Pressure at flash tank 3	Psia	7	Uniform(4,11)
Power recovery turbine efficiency	%	75	Uniform(70,80)
Selexol pump efficiency	%	75	Uniform(70,80)
Recycle gas compressor efficiency	%	75	Uniform(70,80)
CO ₂ compressor efficiency	%	79	Triangular(75,79,85)
Cost parameter	Unit	Value	Distribution function
WGS catalyst cost	\$/ft^3	250	Triangular(220,250,290)
Selexol solvent cost	\$/lb	1.96	Triangular(1.32,1.96,2.9)
Process contingency of WGS system	% of PFC	5	Triangular(2,5,10)
Process contingency of Selexol system	% of PFC	10	Triangular(5,10,20)
Maintenance cost of WGS system	% of PFC	2	Triangular (1, 2, 5)
Maintenance cost of Selexol system	% of PFC	5	Triangular(2,5,10)

 Table E - 2
 Distribution functions assigned to Selexol-based CO2 capture process

Table E - 3Distribution functions assigned to the fuel cost and capacity

Parameter	Unit	Deterministic	Distribution function
Fuel price	\$/MBtu	1.3	Triangular(1.0,1.3,2.1)
Capacity factor	%	75	Triangular(35, 75, 90)

The data sources of the parameters in the above tables are given in the following:

1. Fixed charge factor: Several studies reports different values, half of which range from 14.5% to 15%. Here a triangular distribution with a default mode value of 14.8% is

assigned to this parameter. Values of the fixed charge factor and their sources are given in the following table.

Fixed charge factor (%)	References
7.1	4
11.9	5
14.5~15	6, 7, 8
17.4	9

Fuel price: Fuel cost is the major variable operation cost of an IGCC power plant. The lowest fuel price reported is 1.03 \$/MBtu, and the highest price reported is 2.11 \$/MBtu. For the uncertainty screening purpose, a rough triangular distribution is assigned to the fuel price.

Fuel price (\$/MBtu)	References
1.03	8
1.3~1.5	2, 3, 5, 10
1.58	7
1.66	6
1.79	9
2.11	4

3. Capacity factor: For the uncertainty screening purpose, a triangular distribution with a mode value of 75% is assigned to the capacity factor.

Capacity factor (%)	References
35	11
51	12, 13
60~65	5, 6, 11, 12
65~70	11, 12, 14
70~75	11, 12, 15,
75~80	8, 11, 15
80~85	3, 9, 15
90	7, 16
4. Mole weight of Selexol: Selexol solvent is a mixture of dimethyl ethers of polyethylene glycol with the formulation of CH₃(CH₂CH₂0)_nCH₃, where n is between 3 and 9. Mole weight of Selexol influences the calculation of the flow rate of solvent. Different mole weights are reported. A triangular distribution is assigned to this parameter.

Mole weight of Selexol (lb/mole)	References
265	17
280	18, 19
285	20

5. Pressures of flush tanks: Three flush tanks are used to release CO₂ captured in the solvent at reduced pressures. The flash pressure in each flash tank is a design parameter, which influences the power consumption of CO₂ compression and CO₂ capture efficiency. Due to limited data points, uniform distributions are assigned to the three pressure parameters.

Flash pressure (psia)	References
40 (flash tank 3)	21
75 (flash tank 3)	20
14.7 (flash tank 2)	21
25 (flash tank 2)	20
4 (flash tank 3)	21
11 (flash tank 3)	20

- 6. Power recovery turbine efficiency, Selexol pump and recycle compressor efficiency: Power recovery turbine and pump are common mechanical devices used in industrial fields. The efficiency of such devices may vary depending on the flow type and the operating conditions. The efficiencies, typically, would be around 70-80%. Here a uniform distribution (uniform(70,80)) is employed [22].
- CO₂ compressor efficiency: CO₂ compressor is used to compress CO₂ product to desirable pressures. From the reported data, the compression efficiencies varies from 75% to 88%. Here a triangular is used for this parameter.

CO ₂ compressor efficiency (%)	References
75	23
75~85	24
85	25
88	26

8. Selexol solvent cost: The cost of the sorbent depends on various market forces. Here a triangle distribution is given to the solvent cost.

Selexol cost (\$/lb)	References
1.4	20
1.96	27
2.9	3

- 9. WGS catalyst cost: the deterministic value of the catalyst cost from reference [20]. Based on expert's judgment, the range of the price fluctuation is around $\pm 20\%$, so a triangle distribution is given to the catalyst cost.
- 10. Process contingency of WGS and Selexol system: The deterministic value of the process contingency came from reference [3]. The value range of the contingency came from the recommendation of reference [28].
- 11. Maintenance cost: the maintenance cost is usually expressed as a fraction of the total plant cost. This parameter depends on some design parameters as well as the operating conditions [22]. Based on the recommended range of reference [28], as well as expert's judgments, triangular distributions are assigned to the maintenance costs of WGS and Selexol processes.

REFERENCES (APPENDIX E)

- Frey, H.C., and N. Akunuri, "Probabilistic Modeling and Evaluation of the Performance, Emissions, and Cost of Texaco Gasifier-Based Integrated Gasification Combined Cycle Systems Using ASPEN," Prepared by North Carolina State University for Carnegie Mellon University and U.S. Department of Energy, Pittsburgh, PA, January 2001.
- 2. Shelton W., Lyons J., 2000: Texaco Gasifier IGCC Base Cases, Process Engineering Division, NETL, PED-IGCC-98-001
- Foster Wheeler Ltd, 2003; Potential for Improvement in Gasification Combined Cycle Power Generation with CO₂ Capture," IEA Greenhouse Gas R&D Programme, Report Number PH4/19, May 2003
- 4. Hendriks, C. A., 1994: Carbon Dioxide Removal from Coal-Fired Power Plants. Kluwer Academic Publishers, Dordrecht, the Netherlands
- Condorelli P., Smelser, S. C., McCleary G. J., 1991 Engineering and Economic Evaluation of CO₂ Removal from Fossil-Fuel-Fired Power Plants. Volume 2: Coal Gasification Combined-Cycle Power Plants. Electric Power Research Institute; Report # IE-7365
- 6. Doctor R. D.; Molburg J. C., Thimmapuram P. R., 1997: Oxygen-Blown Gasification Combined Cycle: Carbon Dioxide Recovery, Transport, and Disposal. Energy Convers. Mgmt. Vol. 38, Suppl.
- IEA Greenhouse Gas R&D Programme and Stork Engineering Consultancy B.V., 1999: Assessment of Leading Technology Options for Abatement of CO₂ Emissions; December 1999.
- Simbeck D., 1998: A Portfolio Selection Approach for Power Plant CO₂ Capture, Separation and R&D Options. Fourth International Conference on Greenhouse Gas Control Technologies, Interlaken, Switzerland; 1998.
- Chiesa P., Consonni S, 1998: Comparative Analysis of IGCCs With CO₂ Sequestration. Fourth International Conference on Greenhouse Gas Control Technologies, Interlaken, Switzerland; 1998.
- 10. IECM User manual, 2005, www.iecm-online.com
- 11. McDaniel J.E., Hornick M., 2003: Polk Power Station ICGG 7th year of commercial operation, Gasification technologies, San Francisco, California
- 12. Marco Kanaar, Carlo Wolters, 2004: Fuel Flexibility, the European Gasification Conference, May 2004

- 13. IgnacioMéndez-Vigo, 2003: Elcogas Puertollano IGCC Update, Gasification Technologies, San Francisco, California
- 14. Yamaguchi M., 2004: First Year Operation Experience with the Negishi IGCC, Gasification Technologies 2004, Washington DC
- 15. Keeler C.G., 2003: Operating Experience at the Wabash River Repowering Project, Gasification technologies, San Francisco, California
- 16. Daslav C., BrkicBrkic C., 2003: The Experience of Snamprogetti's Four Gasification Projects, Gasification Technologies, San Francisco, California
- Epps R., 1994: Use of Selexol Solvent for Hydrocarbon Dewpoint Control and Dehydration of Natural Gas, presented at the Laurance Reid Gas Conditioning Conference, Norman, OK, February 27-March 2, 1994
- Sciamanna S. and Lynn S., 1998: Solubility of hydrogen sulfide, sulfur dioxide, carbon dioxide, propane, and n-butane in poly(glycol ethers), Ind. Eng., Chem. Res., 27, 1988
- 19. UOP, 2000: Use of SELEXOL Process in Coke Gasification to Ammonia Project, UOP report
- 20. Doctor R.D., etc., 1996: KRW oxygen-blown gasification combined cycle carbon dioxide recovery, transport, and disposal, ANL/ESD-34, 1996
- Chiesa P. and Consonni S., 1999: Shift Reactors and Physical Absorption for Low-CO₂ Emission IGCCs, Journal of Engineering for Gas Turbines and Power, Vol. 121
- 22. Rao A.B., 2003, A technical, environmental, and economic assessment of Aminebased carbon capture systems for greenhouse gas control, Ph.D thsis, Carnegie Mellon University
- 23. Price, B.C., 1984: Processing high CO₂ gas. Energy Progress, 1984, 4(3)
- 24. Bolland, O. and H. Undrum, 1998: Removal of CO₂ from gas turbine power plants: evaluation of pre- and post-combustion methods. Fourth International Conference on Greenhouse Gas Control Technologies, 30 Aug-2 Sep., 1998, Interlaken, Switzerland
- 25. Hendriks, C., 1994: Carbon dioxide capture from power stations, a report published by IEA Greenhouse Gas R&D Programme, UK
- 26. Alston, 2001: Engineering feasibility and economics of CO₂ capture on an existing coal-fired power plant. Final report prepared by Alstom Power Inc. Report no. PPL-01-CT-09

- 27. Personal communication with UOP.
- 28. EPRI Technical Assessment Guide, 1993