Development and Application of Optimal Design Capability for Coal Gasification Systems: Performance, Emissions, and Cost of Texaco Gasifier-Based Systems Using ASPEN

Technical Report

Work Performed Under Contract No. DE-AC21-92MC29094

Report Submitted to
U.S. Department of Energy
National Energy Technology Laboratory
626 Cochrans Mill Road, P.O. Box 10940
Pittsburgh, Pennsylvania 15236

By
Center for Energy and Environmental Studies
Carnegie Mellon University
Pittsburgh, Pennsylvania 15213

Prepared Under Subcontract By
H. Christopher Frey
Naveen Akunuri
North Carolina State University
Raleigh, North Carolina 27695

January, 2001

LIST OF TABLES

LIST OF FIGURES

1.0 INTRODUCTION

This study deals with the development and application of new systems models for estimating the performance, emissions, and cost of selected gasification-based power generation systems, including characterization of uncertainty in the estimates. Gasification technologies and their commercial status are briefly reviewed with a focus on gasification of coal. The study focuses on modeling and assessment of two Texaco gasifier-based Integrated Gasification Combined Cycle (IGCC) systems using ASPEN. ASPEN is a steady-state chemical process simulator.

The systems models enable the evaluation of the interactions among various process areas within the IGCC systems, as well as the performance and cost of alternative system designs based upon different gas cooling approaches. The technical bases for the models are briefly presented. For each of the systems modeled detailed information is given regarding the process performance, auxiliary power, net plant output, plant efficiency, emissions, capital cost, annual cost, and levelized cost calculations.

A deterministic case study of each of the system models is presented to illustrate the typical performance, emissions, and cost of each system. The uncertainty in the point estimates assumed in the case studies are analyzed for each of the models to characterize uncertainty in model predictions, such as for net plant efficiency, net power output, air pollutant emissions, and capital, annual, and levelized costs. The key uncertainties with respect to plant efficiency and cost are identified. The Texaco gasifier-based IGCC models are intended for use as benchmarks in comparisons with other coal/fuel-based power generation systems, models for many of which have been developed in previous work (Frey and Rubin, 1990; Frey and Rubin, 1991; Frey and Rubin, 1992a; Frey and Rubin, 1992b; Frey, 1994; Frey and Williams, 1995; Frey *et al*, 1994; Agarwal and Frey, 1995; Agrawal and Frey, 1997; Bharvirkar and Frey, 1998). Thus, the models presented here are several of a set of complimentary models that enable comparisons of competing systems for strategic planning purposes.

1.1 Overview of Gasification Systems

Gasification systems are a promising approach for clean and efficient power generation as well as for polygeneration of a variety of products, such as steam, sulfur, hydrogen, methanol, ammonia, and others (Philcox and Fenner, 1996). As of 1996, there were 354 gasifiers located at 113 facilities worldwide. The gasifiers use natural gas, petroleum residuals, petroleum coke, refinery wastes, coal, and other fuels as inputs, and produce a synthesis gas containing carbon monoxide (CO), hydrogen (H_2), and other components. The syngas can be processed to produce liquid and gaseous fuels, chemicals, and electric power. In recent years, gasification has received increasing attention as an option for repowering at oil refineries, where there is currently a lack of markets for low-value liquid residues and coke (Simbeck, 1996).

A general category of gasification-based systems is Integrated Gasification Combined Cycle (IGCC) systems. IGCC is an advanced power generation concept with the flexibility to use coal, heavy oils, petroleum coke, biomass, and waste fuels to produce electric power as a primary product. IGCC systems typically produce sulfur as a byproduct. Systems that produce many co-products are referred to as "polygeneration" systems. IGCC systems are characterized by high thermal efficiencies and lower environmental emissions than conventional pulverized coal fired plants (Bjorge, 1996).

A generic IGCC system is illustrated schematically in Figure 1.1. In an IGCC power plant, the feedstock to the gasifier is converted to a syngas, composed mainly of hydrogen and carbon monoxide, using a gasification process. After passing through a gas cleanup system, in which particles and soluble gases are removed via wet scrubbing and in which sulfur is removed and recovered via a selective removal process, the syngas is utilized in a combined cycle power plant. Different variations of IGCC systems exist based upon the type of coal gasifier technology, oxidant (e.g., oxygen or air), and gas cleanup system employed.

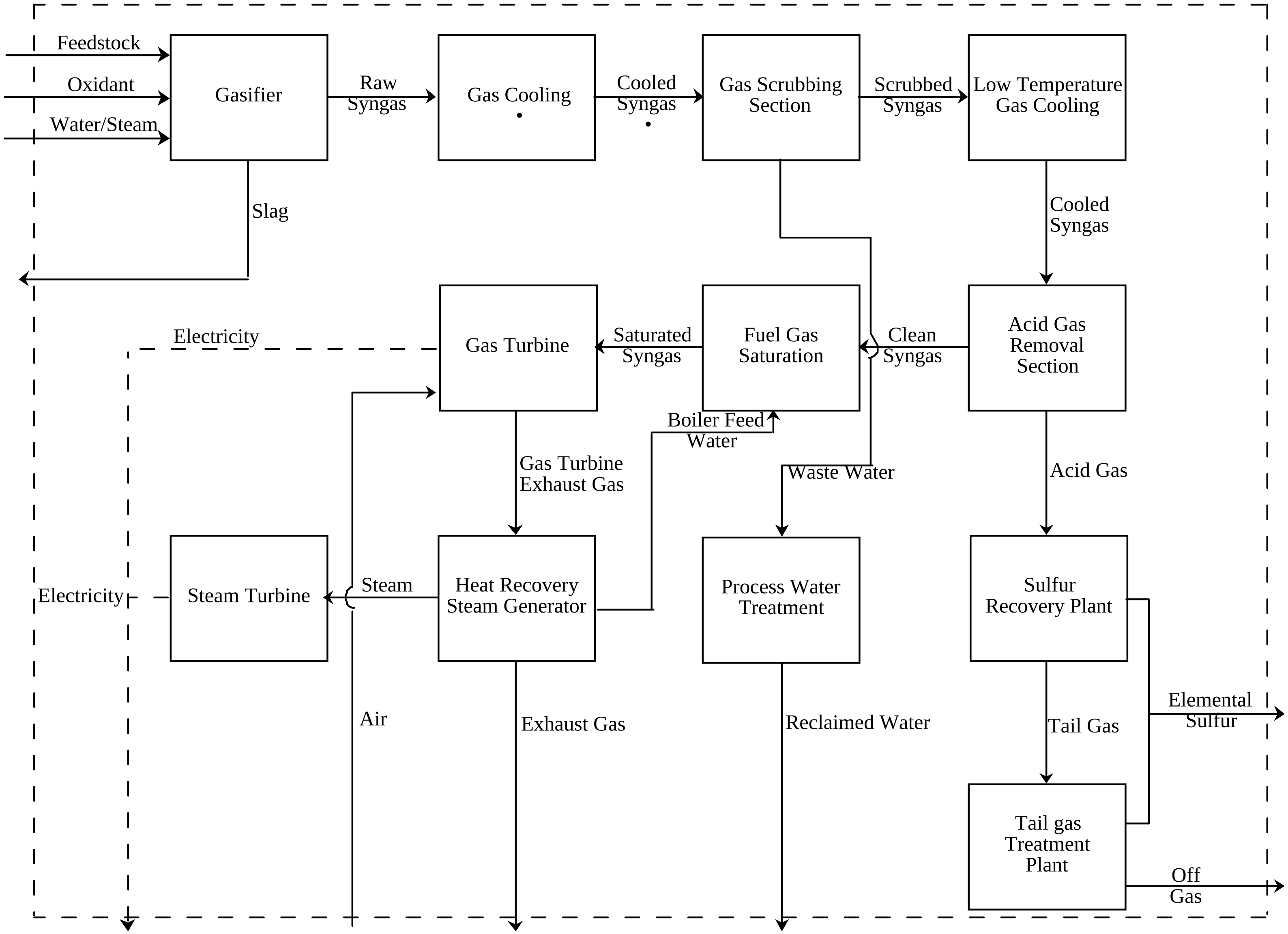

Figure 1.1 Conceptual diagram of an IGCC System

A typical IGCC system includes process sections of Fuel Handling, Gasification, High-Temperature Gas Cooling, Low Temperature Gas Cooling and Gas Scrubbing, Acid Gas Separation, Fuel Gas Saturation, Gas Turbine, Heat Recovery Steam Generator, Steam Turbine, and Sulfur Byproduct Recovery. The specific design of each of the process sections such as gasification and high-temperature gas cooling varies in different IGCC systems.

1.1.1 Gasification

Three generic designs of gasification are typically employed in IGCC systems, each of which are described below. In all types of reactors, the feedstock fuel is converted to syngas in reactors with an oxidant and either steam or water. The oxidant is required to partially oxidize the fuel. The exothermic oxidation process provides heat for the endothermic gasification reactions. Water or steam is used as a source of hydrolysis in the gasification reactions. The type of reactor used is the primary basis for classifying different types of gasifiers.

1.1.1.1 Moving-Bed or Counter-Current Reactors

Moving bed reactors feature counter-current flow of fuel with respect to both the oxidant and the steam. For example, in the case of coal gasification, coal particles of approximately 4 mm to 30 mm (Simbeck *et al*., 1983) in diameter are introduced at the top of the reactor, and move downward. Oxidant is introduced at the bottom of the reactor. A combustion zone at the bottom of the reactor produces thermal energy required for gasification reactions, which occur primarily in the central zone of the reactor. Steam is also introduced near the bottom of the gasifier. As the hot gases from combustion and gasification move upward, they come into contact with the fuel introduced at the top. The heating of the fuel at the top of the reactor results in devolatilization, in which lighter hydrocarbon compounds are driven off and exit as part of the syngas. Because the gases leaving the gasifier contact the relatively cool fuel entering the gasifier, the exit syngas temperature is relatively low compared to other types of reactors. The counter-current flow of fuel with the oxidant and steam can result in efficient utilization of the fuel, as long as the residence time of the fuel is long enough for even the larger particles to be fully consumed. Ash and unconverted fuel exit the bottom of the gasifier via a rotating grate.

A typical syngas exit temperature for a moving bed gasifier is approximately 1,100 ^{o}F. At this temperature, some of the heavier volatilized hydrocarbon compounds, such as tars and oils, will not be cracked and can easily condense in downstream syngas cooling equipment. Because fuel is introduced at the top of the gasifier where the syngas is exiting, this type of gasifier cannot handle fine fuel particles. Such particles would be entrained with the exiting syngas and would not be converted to syngas in the reactor bed. Cyclones are typically used to capture fine particles in the syngas, which are often sent to a briquetting facility to form larger particles and then recycled to the gasifier for another attempt at conversion.

An overall measure of gasifier performance is the cold gas efficiency. The cold gas efficiency is the ratio of the heating value of "cold" syngas, at standard temperature, to the heating value of the amount of fuel consumed required to produce the syngas. The cold gas efficiency does not take into account recovery of energy in the gasifier such as through steam generation or associated with sensible heat of the syngas at high temperatures. Moving bed gasifiers tend to have very high cold gas efficiencies, with values in the range of 80 to 90 percent.

Typical examples of such reactors are Lurgi dry bottom gasifiers and the British Gas/Lurgi slagging gasifiers.

1.1.1.2 Fluidized-Bed Gasifiers

Fluidized bed reactors feature rapid mixing of fuel particles in a 0.1 mm to 10 mm size range with both oxidant and steam in a fluidized bed. The feedstock fuel, oxidant and steam are introduced at the bottom of the reactor. In these reactors, backmixing of incoming feedstock fuel, oxidant, steam, and the fuel gas takes place resulting in a uniform distribution of solids and gases in the reactors. The gasification takes place in the central zone of the reactor. The coal bed is fluidized as the fuel gas flow rate increases and becomes turbulent when the minimum fluidizing velocity is exceeded.

The reactors have a narrow temperature range of 1800 ^{o}F to 1900 ^{o}F. The fluidized bed is maintained at a nearly constant temperature, which is well below the initial ash fusion temperature to avoid clinker formation and possible defluidization of the bed. Unconverted coal in the form of char is entrained from the bed and leaves the gasifier with the hot raw gas. This char is separated from the raw gas in the cyclones and is recycled to the hot ash agglomerating zone at the bottom of the gasifier. The temperature in that zone is high enough to gasify the char and reach the softening temperature for some of the eutectics in the ash. The ash particles stick together, grow in size and become dense until they are separated from the char particles, and then fall to the base of the gasifier, where they are removed.

The processes in these reactors are restricted to reactive, non-caking coals to facilitate easy gasification of the unconverted char entering the hot ash zone and for uniform backmixing of coal and fuel gas. The cold gas efficiency is approximately 80 percent (Supp, 1990). These reactors have been used for Winkler gasification process and High-temperature Winkler gasification process. A key example of fluidized gasification design is the KRW gasifier.

1.1.1.3 Entrained-Flow Reactors

The entrained-flow process features a plug type reactor where the fine feedstock fuel particles (less than 0.1 mm) flow co-currently and react with oxidant and/or steam. The feedstock, oxidant and steam are introduced at the top of the reactor. The gasification takes place rapidly at temperatures in excess of 2300 ^{o}F. The feedstock is converted primarily to H_2, CO, and CO_2 with no liquid hydrocarbons being found in the syngas. The raw gas leaves from the bottom of the reactor at high temperatures of 2300 ^{o}F and greater. The raw gas has low amounts of methane and no other hydrocarbons due to the high syngas exit temperatures.

The entrained flow gasifiers typically use oxygen as the oxidant and operate at high temperatures well above ash slagging conditions in order to assure reasonable carbon conversion and to provide a mechanism for slag removal (Simbeck *et al.*, 1983). Entrained-flow gasification has the advantage over the other gasification designs in that it can gasify almost all types of coals regardless of coal rank, caking characteristics, or the amount of coal fines. This is because of the relatively high temperatures which enable gasification of even relatively unreactive feedstocks that might be unsuitable for the lower temperature moving bed or fluidized bed reactors. However, because of the high temperatures, entrained-flow gasifiers use more oxidant than the other designs. The cold gas efficiency is approximately 80 percent (Supp, 1990). Typical examples of such reactors are Texaco Gasifiers and Destec Gasifiers.

The advantage of adopting entrained flow gasification over the above mentioned reactors is the high yield of synthesis gas containing insignificant amounts of methanol and other hydrocarbons as a result of the high temperatures in the entrained-flow reactors.

Texaco gasification is a specialized form of entrained flow gasification in which coal is fed to the gasifier in a water slurry. Because of the water in the slurry, which acts as heat moderator, the gasifier can be operated at higher pressures than other types of entrained-flow gasifiers. Higher operating pressure leads to increased gas production capability per gasifier of a given size (Simbeck *et al*., 1983)

In this study, we focus on modeling assessment of entrained flow gasification. Assessments of moving bed and fluidized bed gasifier based systems have been done in previous work (Frey and Rubin, 1992a, 1992b, Frey et al., 1994, Frey, 1998).

1.1.2 High Temperature Gas Cooling

The design of the high temperature syngas cooling process area depends on the type of gasifier used. The gas cooling requirements for entrained flow gasification systems are more demanding than for other gasification systems as the former produce syngas at higher temperatures. Typically, the gas cooling process for systems employing entrained flow gasification systems either use heat exchangers to recover thermal energy and generate steam or use water quenching. The former design can be radiant and convective or radiant only, while the latter is known as total quench high temperature gas cooling. The former is more efficient as it can produce high temperature and pressure steam, whereas the latter is much less expensive (Doering and Mahagaokar, 1992).

1.1.2.1 Radiant and Convective Syngas Cooling Design

The design of a radiant and convective gasification system is shown in Figure 1.2. Each gasifier has one radiant cooler and one convective cooler. The hot syngas is initially cooled in an radiant heat transfer type of heat exchanger. High pressure steam is generated in tubes built into the heat transfer surface at the perimeter of the cylindrical gas flow zone. The molten slag drops into a slag quench chamber at the bottom of the radiant gas cooler where it is cooled and removed for disposal. The gas leaves the radiant cooler at a temperature of approximately 1500 ^{o}F.

The syngas from the radiant heat exchanger flows into a convection type of heat exchanger. In the convective heat exchanger, the syngas flows across the boiler tube banks. These tubes help remove the entrained particles in the syngas that are too fine to drop out in the bottom of the radiant cooler. High pressure steam is generated in these tubes. The cooled gas leaves the convective chamber at a temperature of approximately 650 ^{o}F.

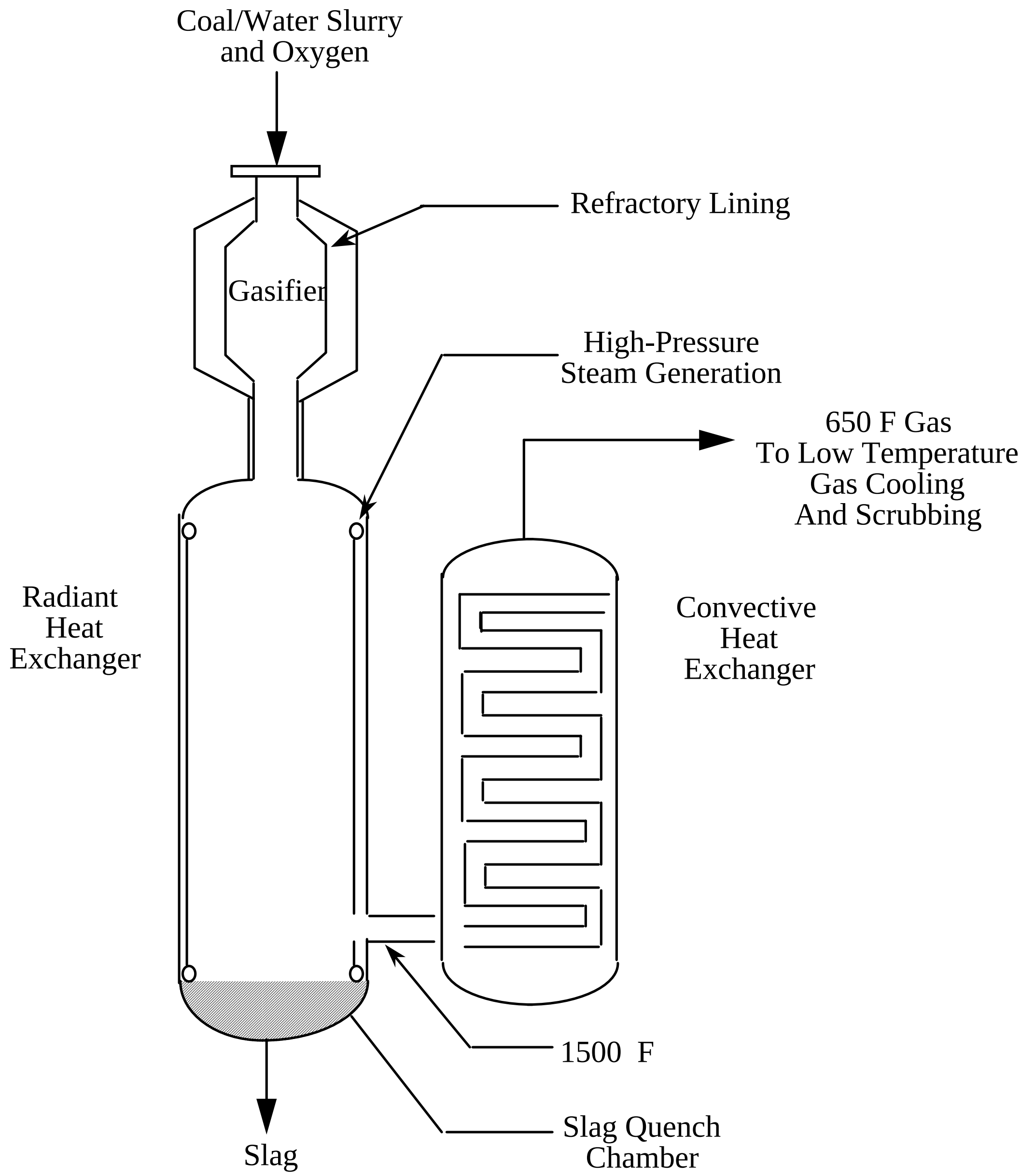

Figure 1.2 Radiant and Convective High Temperature Syngas Cooling Design

1.1.2.2 Radiant Only Syngas Cooling Design

The hot syngas is cooled initially in the radiant cooler and high pressure steam is generated as in the radiant and convective design. However, in this case both the molten slag and the raw gas are quenched in the water pool at the bottom of the radiant cooler. The cooled slag is removed from the cooler for disposal. The raw gas, saturated with moisture, flows out of the radiant cooler at a temperature of approximately 400 ^{o}F.

1.1.2.3 Total Quench Design

The total quench design is depicted in Figure 1.3. In this design, the hot syngas and the molten slag particles flow downward through a water spray chamber and a slag quench bath. Water is sprayed just beneath the partial oxidation chamber to cool the hot syngas. The entrained slag is separated from the syngas in the slag quench bath (Nowacki, 1981). There is no high pressure steam generation in this method as in the previous two designs since there is no heat recovery. The raw gas saturated with moisture flows to the gas scrubbing unit at a temperature of 430 ^{o}F.

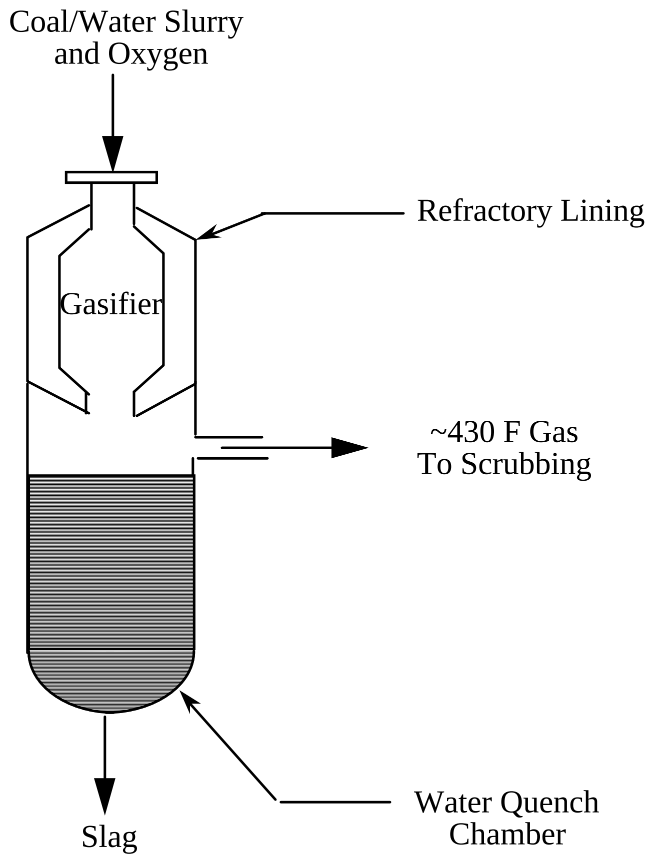

Figure 1.3 Total Quench High Temperature Syngas Cooling Design

In this study, both the radiant and convective and the total quench high temperature syngas cooling designs are evaluated. The radiant and convective design has the advantage over total quench syngas cooling of a higher plant efficiency. However, the cost of the radiant and convective design is higher than that of the total quench design. The total quench design results in increased moisturization of syngas, which can prove effective in terms of preventing NO_x formation in the gas turbine combustor and in terms of augmenting power production from the gas turbine. In a water quench system, large quantities of water are used and thus contaminated by the slag, requiring complex primary and secondary treatment facilities. Hence total quench design has additional operating problems such as those caused due to increased water treating facilities, increased discharge water permitting issues, and added operating and maintenance costs when compared to radiant and convective design (Doering and Mahagaokar, 1992).

1.2 Commercial Status of Coal and Heavy Residual Oil-Fueled Gasification Systems

The IGCC concept has been demonstrated commercially. Table 1.1 lists the IGCC plants currently in operation or undergoing construction. The Texaco coal gasification process has been successfully used in a number of chemical plants since the early 1980s for the production of synthesis gas from coal. A Texaco-based 95 MW IGCC power plant was operated successfully from 1984 to 1988 in California (Simbeck *et al*., 1996). API Energia, a joint venture of Asea Brown Boveri and API, adopted Texaco gasification to gasify visbreaker residue from an API refinery to produce steam and power. Tampa Electric Company’s Polk Power station also utilizes Texaco gasification, gasifying about 2,000 tons of coal per day to produce 250 MW of power. The El Dorado gasification project demonstrates that hazardous waste streams can be converted by gasification to valuable products. (Farina *et al*., 1998).

A Destec gasifier-based IGCC power plant at Wabash River Station is currently under operation (Simbeck *et al*., 1996). A 335 MW IGCC demonstration plant for European electricity companies is operating at Puertollano, Spain (Mendez-vigo *et al*., 1998). The Texaco gasifier-based El Dorado plant, the Shell-Pernis plant in Netherlands, and the Sarlux plant in Italy using low pressure (38 barg) Texaco gasification to produce hydrogen and/or steam along with power (Bjorge *et al*., 1996).

Table 1.1 IGCC Projects Under Operation or Construction

Project	Location	Start-up Date	Plant Size	Products	Gasifier	Fuel
Cool Water IGCC	Barstow, California	1984	120 MW	Power	Texaco	Coal
PSI Wabash River	Terre Haute, Indiana	1996	262 MW	Power	Destec	Coal
Tampa Electric	Polk, Florida	1996	250 MW	Power	Texaco	Coal
Pinon Pine Sierra Pacific	Sparks, Nevada	1996	100 MW	Power	KRW	Coal
Texaco El Dorado	El Dorado, Kansas	1996	40 MW	Co-generation Steam and H_2	Texaco	Pet Coke
Shell Pernis	Netherlands	1997	120 MW	Co-generation H_2	Shell/Lurgi	Oil
Sarlux	Sarroch, Italy	1998	550 MW	Co-generation Steam	Texaco	Oil
API Energia	Falconara Marittima	1999	234 MW	Power	Texaco	Oil
Puertallano		1997	335 MW	Power	Prenflo	Coal

1.3 Motivating Questions

In order to study the benefit and risks of a new process technology, there is a need for the development of a systematic approach for technology assessment. The performance, emissions and costs of individual IGCC systems need to be characterized as a basis for comparison with conventional and with other advanced alternatives. There is also a need to develop a baseline case study of an existing commercial IGCC technology for comparison with other more advanced (less commercial) IGCC systems in future technology studies. The present study deals with the study of an existing commercial IGCC technology and has the following motivating questions.

1. What are the thermal efficiencies, emissions, and costs of selected entrained-flow gasification-based IGCC systems when fueled by coal?
2. How does the design of the high temperature gas cooling system of a coal-fueled IGCC system affect the performance, emissions, and costs?
3. What are the uncertainties in the point estimates assumed for the IGCC systems?
4. What are the key sources of uncertainty in the performance, emissions, and costs of the technologies?

1.4 Objectives

The objectives of the current work are:

1. To develop new systems models based upon the best available information regarding process performance, emissions and cost for the following configurations:
   - (a) Oxygen-blown coal-fueled Texaco gasifier-based IGCC system with radiant and convective high temperature syngas cooling; and
   - (b) Oxygen-blown coal-fueled Texaco gasifier-based IGCC system with total quench high temperature syngas cooling.
2. To verify the models;
3. To compare the high temperature syngas cooling designs;

4. To characterize uncertainty in the performance, emissions, and costs of these systems to provide insight into the potential pay-offs and downside risks of these technologies.

1.5 General Methodological Approach

This section describes the methodologies adopted for the development of performance, emissions and costs of two IGCC systems and the integration of the performance and cost models. The requirement for a probabilistic analysis of the models developed is also discussed.

1.5.1 Performance and Cost Model Development of the IGCC System

The Federal Energy Technology Center (FETC) of the U.S. Department of Energy has developed a number of performance simulations of IGCC systems in the ASPEN modeling environment. A number of these models have been refined by Frey and others (Frey and Rubin, 1991, Frey *et al*., 1994, Frey, 1998) in order to calculate mass and energy balances for IGCC systems, conduct sensitivity analyses of performance parameters, track environmental species, and evaluate design modifications. Subroutines that calculate capital, annual, and levelized costs have also been developed and incorporated with the refined performance models.

The Texaco gasifier-based IGCC system models developed in this study are based primarily on the general configuration and design basis of a study sponsored by the Electric Power Research Institute (EPRI) (Matchak *et al*., 1984). K.R. Stone developed a process simulation model based on the radiant and convective high temperature gas cooling design in 1985 at FETC. This FETC model has been substantially refined for this study.

The IGCC simulation models of radiant and convective gasifier design and total quench high temperature gas cooling design developed in the present study are intended to predict the output values of process performance measures (e.g., plant thermal efficiency) for a given set of input assumptions. The key refinements to the earlier FETC model, which are also incorporated into the new model of the total quench based system, include complete replacement of the gas turbine flowsheet with a more detailed model, implementation of a more detailed fuel gas saturation model, incorporation of NO_x emissions as a model output, refinement and more comprehensive inclusion of auxiliary power demand estimates, and implementation of a capital, annual, and levelized cost model. The key improvements to the original FETC model of the radiant and convective based system are described in more detail for the gas turbine and the cost model in Chapters 3.0 and 5.0 and the auxiliary power consumption models are elaborated upon in Chapter 4.0.

1.5.2 Modeling Process Flowsheets in ASPEN

The performance model of the Texaco-based IGCC was developed as an ASPEN input file. ASPEN is a FORTRAN-based deterministic steady-state chemical process simulator developed by the Massachusetts Institute of Technology (MIT) for DOE to evaluate synthetic fuel technologies (MIT, 1987). The ASPEN framework includes a number of generalized unit operation "blocks", which are models of specific process operations or equipment (e.g., chemical reactions, pumps). By specifying configurations of unit operations and the flow of material, heat, and work streams, it is possible to represent a process plant in ASPEN. In addition to a varied set of unit operations blocks, ASPEN also contains an extensive physical property database and convergence algorithms for calculating results in closed loop systems, all of which make ASPEN a powerful tool for process simulation.

ASPEN uses a sequential-modular approach to flowsheet convergence. In this approach, mass and energy balances for individual unit operation blocks are computed sequentially, often in the same order as the sequencing of mass flows through the system being modeled. However, when there are recycle loops in an ASPEN flowsheet, stream and block variables have to be manipulated iteratively in order to converge upon the mass and energy balance. ASPEN has a capability for converging recycle loops using a feature known as "tear streams."

In addition to calculations involving unit operations, there are other types of blocks used in ASPEN to allow for iterative calculations or incorporation of user-created code. These include design specifications and FORTRAN blocks.

A design specification is used for feedback control. The user can set any flowsheet variable or function of flowsheet variables to a particular design value. A feed stream variable or block input variable is designated to be manipulated in order to achieve the design value. FORTRAN statements can be used within the design specification block to compute design specification function values.

FORTRAN blocks are used for feedforward control. Any FORTRAN operation can be carried out on flowsheet variables by using in-line FORTRAN statements that operate on these variables. FORTRAN blocks are one method for incorporating user code into the model. It is also possible to call any user-provided subroutine from either a design specification or FORTRAN block.

1.5.3 Modeling Methodology for Cost Estimation

The variety of approaches available to developing cost estimates for process plants differ in the level of detail with which costs are separated, as well as in the simplicity or complexity of analytic relationships used to estimate line item costs. The level of detail appropriate for the cost estimate depends on: (1) the state of technology development for the process of interest; and (2) the intended use of the cost estimates.

The models developed here are intended to estimate the costs of innovative coal-to-electricity systems for the purpose of evaluating the comparative economics of alternative process configurations. The models are intended to be used only for preliminary or "study grade" estimates using representative (generic) plant designs and parameters.

In the electric utility and chemical process industries, there are generally accepted guidelines regarding the approach to developing cost estimates. The Electric Power Research Institute has defined four types of cost estimates: simplified, preliminary, detailed, and finalized. The cost estimates developed in this work are "preliminary" (EPRI, 1986). Preliminary cost estimates are appropriate for the purposes of evaluating alternative technologies, and for research planning. These cost estimates are sensitive to the performance and design parameters that are most influential in affecting costs (Frey and Rubin, 1990).

One of the major constraints on the development of the cost model is the availability of data from which to develop cost versus performance relationships for specific process area or for major equipment items. Data from published studies can be used to develop cost models for specific process areas using regression analysis (Frey and Rubin, 1990).

The new cost models developed for each of the three technologies evaluated in this work include capital, annual, and levelized costs. The models estimate the direct capital costs of each major plant section as a function of key performance and design parameters. The total capital cost is calculated based on direct and indirect capital costs. The total direct cost is a summation of the plant section direct costs and general facilities cost. The total indirect cost is the sum of indirect construction costs, engineering and home office fees, sales tax, and environmental permitting costs. The latest process contingency factors have been incorporated in the cost model and are included in the total capital cost.

The annual cost model includes both fixed and variable operating costs. Fixed operating costs include operating labor, maintenance labor and materials, and overhead costs associated with administrative and support labor. The latest maintenance cost factors have been included in the cost model in order to calculate process area annual maintenance cost. Variable operating costs include fuel, consumables, ash disposal, and byproduct credits. The operating costs are estimated based on 31 cost parameters such as unit prices and costs (Frey and Rubin, 1990).

1.5.4 Integration of Performance and Cost Models

The cost model has been developed as a FORTRAN subroutine, which is linked to the ASPEN simulation model. The cost model obtains approximately 50 to 60 process variables from the ASPEN performance model for use in both the capital and annual cost calculations. Newly developed regression models are used to calculate the auxiliary power requirements for many of the process areas. The overall plant efficiency is calculated in the cost model subroutine taking into account the gross gas turbine and steam turbine output and the auxiliary power demands.

1.5.5 Probabilistic Analysis

The complexity of gasification systems implies that it is difficult to evaluate all possible combinations of gasification components based upon the relatively small population of demonstration and commercial plants. For each of the major components of a typical gasification system (e.g., fuel feed, gasification, syngas cooling, syngas cleanup, power generation, byproduct recovery), there are many possible options. Limited performance and cost data for first generation systems, coupled with uncertainties associated with a large number of alternative process configurations, motivates a systematic approach to evaluating the risks and potential pay-offs of alternative concepts.

Technology assessment models are typically developed for the purpose of providing a point-estimate which may be intended to serve as an accurate and precise prediction of some quantity (e.g., thermal efficiency, total capital cost). The purpose of such analyses are to provide decision makers with a best-estimate that can be used in comparison with other assessments or to develop design targets or budgetary cost estimates. However, quantitative measures of the accuracy and precision of model predictions are usually not developed, because no information on model or input uncertainty is accounted for quantitatively. Deterministic estimates for the performance and cost of new process technologies are often significantly biased toward optimistic outcomes (Merrow *et al*., 1981). Such biases can lead to serious misallocation of resources if decisions are made to pursue research and development on a technology whose risks were not properly quantified.

To explicitly represent uncertainties in gasification systems and other process technologies, a probabilistic modeling approach has been developed and applied. This approach features: (1) development of sufficiently detailed engineering models of performance, emissions, and cost; (2) implementation of the models in a probabilistic modeling environment; (3) development of quantitative representations of uncertainties in specific model parameters based on literature review, data analysis, and elicitation of technical judgments from experts; (4) identification of key uncertainties in the model input variables; and (5) modeling applications for cost estimating, risk assessment, and research planning. The methods have been applied to previous case studies of gasification and other advanced power generation and environmental control systems (e.g., Frey and Rubin 1992; Frey *et al*., 1994).

1.6 Overview of the Report

The organization of the report is as per the following order. Chapter 2 provides a technical background for Texaco gasifier-based IGCC systems. Chapter 3 elaborates on the development of the performance model of a coal-fueled IGCC system with Texaco gasifier with radiant and convective design. Chapter 4 documents the auxiliary power consumption models of the IGCC plant developed in Chapter 3. The direct capital costs of the IGCC system with the radiant and convective high temperature gas cooling design are modeled in Chapter 5. The model developed in Chapters 3 to 5 is applied to a deterministic case study in Chapter 6. Chapters 7 discusses the development of the performance, emissions, and costs of a coal-fueled Texaco gasifier-based IGCC system with total quench high temperature gas cooling. Chapter 8 provides the results of applying the model developed in Chapter 7 to a deterministic case study. Chapter 9 discusses the uncertainty analysis performed on the three IGCC models developed in the present study. Chapter 10 presents the conclusions obtained from the current study.

2.0 TECHNICAL BACKGROUND FOR INTEGRATED GASIFICATON COMBINED CYCLE SYSTEMS

This study describes performance and cost models for two Texaco gasifier-based IGCC systems: (1) radiant and convective high temperature syngas cooling using coal; and (2) total quench high temperature syngas cooling using coal. IGCC systems for radiant and convective model and total quench model are illustrated schematically in Figure 1.1. The fuel is fed to the gasifier in a slurry in the case of coal being used as feedstock. Oxygen is used to combust only a portion of the feedstock in order to provide thermal energy needed by endothermic gasification reactions. The raw syngas leaves the gasifier at approximately 2400 ^oF and cooled either by a series of radiant and convective heat exchangers to a temperature of 650 ^oF or by contact with water to a temperature of 433 ^oF. The syngas passes through a wet scrubbing system to remove particulate matter and water soluble gases such as NH_3.

The scrubbed gas is further cooled to 101 ^oF prior to entering a Selexol acid gas separation unit. H_2S and COS are removed from the syngas in the Selexol unit and sent to a Claus plant and a Beavon-Stretford tail gas treatment unit for sulfur recovery. The clean gas is reheated and saturated with moisture prior to firing in a gas turbine. The saturation helps prevent formation of thermal NO_x during combustion. The hot gas turbine exhaust passes through a Heat Recovery Steam Generator (HRSG) to provide energy input to a steam turbine bottoming cycle. Both the gas turbine and the steam turbine generate power.

The details of the major process areas are briefly described.

2.1 Gasification

Texaco gasification can handle a wide variety of feedstocks including coal, heavy oils, and petroleum coke (Preston, 1996). The current study focuses on IGCC systems using coal feed. The feed coal is crushed and slurried in wet rod mills. The coal slurry containing about 66.5 weight percent solids is fed into the gasifier, which is a open, refractory-lined chamber, together with a feed stream of oxidant. The slurry is transferred to the gasifier at high pressure through charge pumps. The water in the coal slurry acts as a temperature moderator and also as a source of hydrogen in gasification (Simbeck *et al.*, 1983). Oxygen is assumed as the oxidant for the IGCC systems evaluated in this study. The oxidant stream contains 95 percent pure oxygen. The oxygen is compressed to a pressure sufficient for introduction into the burner of the Texaco gasifier (Matchak *et al.*, 1984).

Gasification takes place rapidly at temperatures exceeding 2300 ^{o}F. Coal is partially oxidized at high temperature and pressure. Figure 2.1 demonstrates the temperature variation across the gasifier (Simbeck *et al.*, 1983). The combustion zone is near the top of the reactor, where the temperature in the gasifier changes from approximately 250 to 2500 ^{o}F. As a result, a raw gas composed mainly of carbon dioxide, carbon monoxide, hydrogen, and water vapor is produced. The syngas contains soot particles. The syngas leaves the gasifier at temperatures in the range of 2300 ^{o}F to 2700

^{o}F. Because of the high temperatures characteristic of entrained-flow gasifiers, the syngas contains smaller amounts of methane than other types of gasifiers and is free of tars and other hydrocarbons (Simbeck *et al*., 1983).

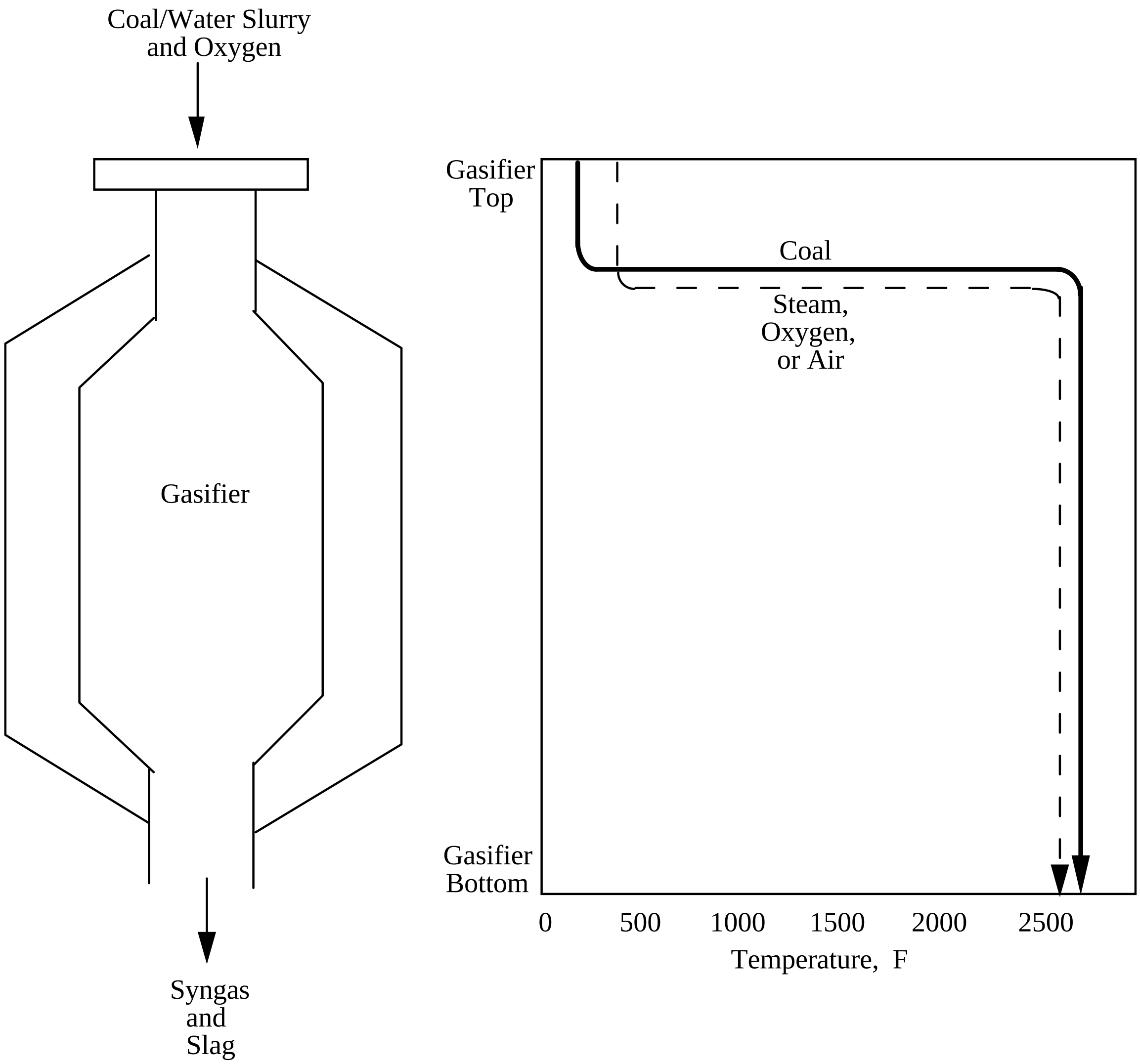

Figure 2.1 Temperature Variation in an Entrained Gasifier

(Based on Simbeck *et. al*., 1983)

2.2 High-Temperature Gas Cooling

In the case of radiant and convective (RC) based system model, the hot gas from the gasifier is initially cooled in a radiant heat exchanger. High pressure steam is generated in tubes built into the heat transfer surface at the perimeter of the cylindrical gas flow zone. Molten slag entrained in the raw gas drops into a water quench pool at the bottom of the radiant gas cooler, where it is cooled and removed for disposal. The gas leaves the radiant cooler at a temperature of approximately 1500 ^{o}F, and enters a convective heat exchanger. In the convective gas cooler, the gas flows across boiler tube banks, where high pressure steam is generated. The syngas leaves the convective cooler at a temperature of approximately 650 ^{o}F, and flows to the gas scrubbing unit.

In the total quench case, the hot gas is cooled in a water spray chamber and then directly quenched in a quench pool at the bottom of the gasifier and is cooled to a temperature of 433^{o}F before it flows to the gas scrubbing unit.

2.3 Gas Scrubbing Process and Low Temperature Gas Cooling

The cooled syngas from the high temperature gas cooling section enters the gas scrubbing unit, where it is washed with water to remove fine particles. The particle-laden water is sent to a water treatment plant and used again as quench water. The scrubbed gas enters various heat exchangers in the low temperature gas cooling section. The heat removed from the syngas is utilized to generate low-pressure steam to heat feed water or as a source of heat for fuel gas saturation.

2.4. Sulfur Removal Process

The syngas from the low temperature gas cooling section enters the acid gas removal section of the plant. The acid gas removal system employs the Selexol process for selective removal of hydrogen sulfide (H_2S) and carbonyl sulfide (COS). Usually COS is present in much smaller quantities than H_2S. In this unit, most of the entering H_2S is removed by absorption in the Selexol solvent, with a typical removal efficiency of 95 to 98 percent (Simbeck *et al*., 1983). Typically only about one third of COS in the syngas will be absorbed. H_2S and COS stripped from the Selexol solvent, along with sour gas from the process water treatment unit is sent to the Claus sulfur plant for recovery of elemental sulfur.

2.5 Fuel Gas Saturation and Combustion

Thermal NO_X constitutes a major portion of the total NO_x emissions from a gas turbine combustor fired on syngas. To control the formation of thermal NO_x, water vapor must be introduced along with the cleaned gas into the combustors of gas turbines. The water vapor lowers the peak flame temperatures. The formation of NO from nitrogen and oxygen in the inlet air is highly temperature sensitive. Lowering the peak flame temperature in the combustor by introducing water vapor results in less formation of thermal NO and hence, lower NO emissions.

Another advantage of fuel gas moisturization is to increase the net power output of the gas turbine. The introduction of moisture into the syngas lowers the syngas heating value and requires an increase in fuel mass flow in order to deliver the same amount of total heating value to the gas turbine engine. Because the mass flow of combustor gases is constrained by choked flow conditions at the turbine inlet nozzle, the inlet air flow has to be reduced to compensate for the increased fuel flow. This results in less power consumption of power by the gas turbine compressors, resulting in an increase in the net gas turbine output.

The saturation of fuel gas takes place in a saturator vessel, which is adiabatic. The clean gas from the acid gas removal system enters the saturator from the bottom while hot water, which is at a higher temperature than that of the syngas, is sprayed from the top of the vessel, as shown in Figure 2.2. The typical temperature of the hot water is 380 ^{o}F, while that of the syngas is 85 ^{o}F before saturation. The saturated gas is heated to a temperature of approximately 350 ^{o}F and exits from the saturator from the top of the vessel while the hot water gets cooled and exits from the bottom of the vessel. The heat needed for heating the water is transferred from low temperature gas cooling units and the heat recovery steam generators to the fuel gas saturation unit as shown in Figure 2.3. A portion of the cold water leaving the fuel gas saturator is sent to heat exchangers in low temperature gas cooling section, where it get heated while cooling the hot syngas from the gas scrubbing section. The remaining portion of cold water is heated by heat exchange with boiler feedwater from the heat recovery steam generation system. Both the portions of heated water are combined to form the hot water spraying from the top of the saturator vessel. The clean, medium BTU gas from the fuel gas saturator is combusted in the gas turbine combustors.

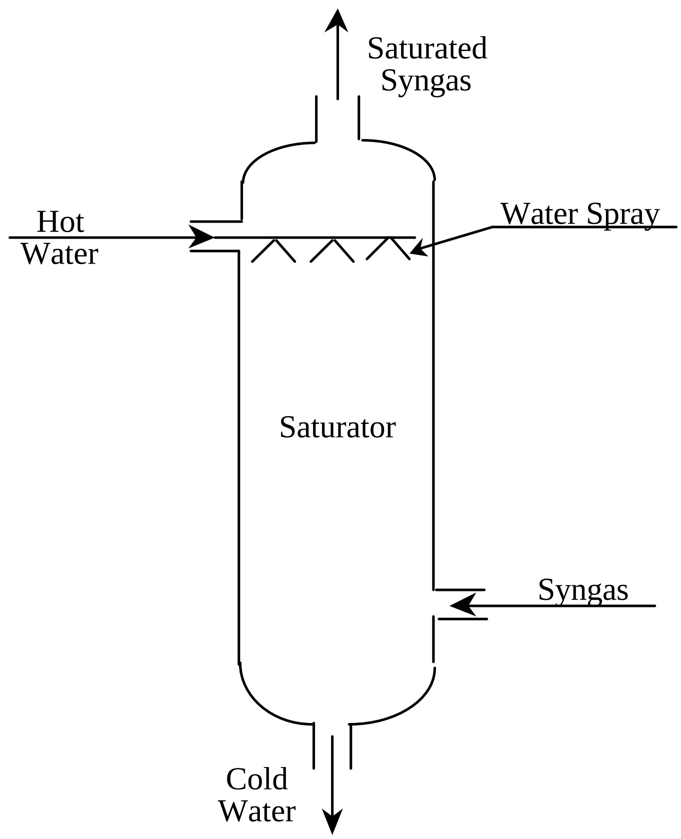

Figure 2.2 Fuel Gas Saturator

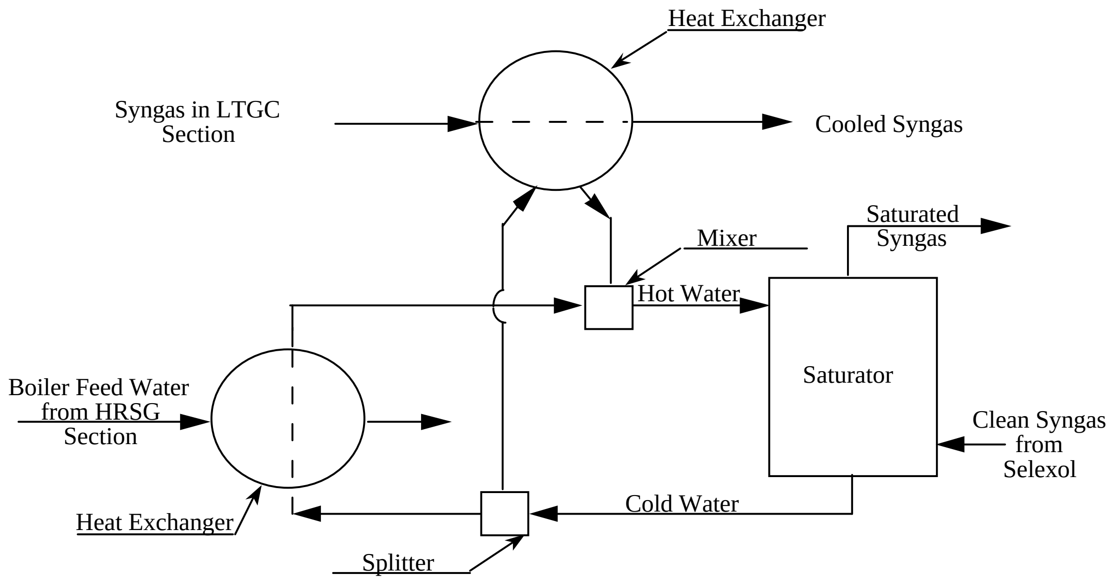

Figure 2.3 Simplified Schematic of Fuel Gas Saturation

2.6 Combined Cycle

A combined cycle system is composed of a gas turbine and a bottoming steam cycle. Both the gas turbine and the steam turbine provide shaft energy to a generator for production of electricity. The gas turbine primarily consists of a compressor, a combustor, and an expander. The compressor supplies required air to the combustor. The combustor is divided into a section for stoichiometric adiabatic combustion of the fuel gas and a subsequent section for quench of the primary combustion products with secondary air. The gases exiting the quench stage of the combustor are at the turbine inlet temperature. The hot exhaust gases from the gas turbine combustors are at a temperature of 2350 ^{o}F. The hot gases are sent to the gas turbine expanders, which in turn drive the generators.

If the gas turbine design is used for syngas as well as for natural gas, then the total mass flow through the turbine is more or less equal in both the cases. However, the heating value of natural gas is higher than the heating value of syngas. Therefore, the fuel flow rate for syngas is significantly larger than that for the natural gas. Typically, the mass flow at the turbine inlet nozzle is limited by choking. Therefore, an increase in the fuel mass flow rate must be compensated by a reduction in the compressor air flow rate, for a given pressure ratio and firing temperature. This causes a net reduction in the power consumed by the compressors leading to a net increase in the gas turbine output.

The hot gas turbine exhaust gases enter the heat recovery steam generator (HRSG) process area. The heat recovery steam generator system has gas-gas heat exchangers that recover the sensible heat from the hot exhaust gases. The HRSG consists of a superheat system including reheaters, high pressure evaporators, and boilers. High pressure steam is generated in the superheat steam system using the heat recovered from the hot turbine exhaust gases. This unit also superheats the high pressure saturated steam generated in the high temperature gas cooling unit in the radiant and convective cooling process. The exhaust gases that have been cooled flow out of the heat recovery steam generators at temperatures in the range of 250 ^{o}F to 300 ^{o}F. Most of the steam generated in the HRSGs is sent to the steam turbines where it is expanded and more electric power is generated. A portion of steam is sent to the fuel gas saturation unit.

3.0 DOCUMENTATION OF THE PLANT PERFORMANCE SIMULATION MODEL IN ASPEN OF THE COAL-FUELED TEXACO-GASIFIER BASED IGCC SYSTEM WITH RADIANT AND CONVECTIVE HIGH TEMPERATURE GAS COOLING

This chapter presents the ASPEN simulation model of the performance of an IGCC system using a Texaco gasifier with radiant and convective high temperature gas cooling. The details involving the modeling of the mass and energy balances of the major process sections of the system are described. Tables and figures describing the components of the process sections of the IGCC system model are listed in detail. The convergence sequence, which specifies the calculation sequence of the simulation model, is presented. Also given is a list of FORTRAN blocks and design specifications required for the simulation model. Finally, the methods used for modeling the air pollutant emissions from the IGCC system are discussed.

3.1 Process Description

The model of the Texaco gasifier-based IGCC system with radiant and convective high temperature gas cooling is based primarily on the findings of a study sponsored by the Electric Power Research Institute (Matchak *et al.*, 1984). This study provides extensive information on the mass flows, temperatures, and pressures of streams, power production and consumption, and costs associated with each process section of the plant. Thus it provides comprehensive and internally consistent information for use in model development.

The model presented in this study is based upon a previous model developed by K.R.Stone in 1985 for FETC. The modifications that were made to the previous model include incorporation of a new and more detailed model of the gas turbine, implementation of a fuel gas saturation model, modeling of NO_X emissions, incorporation of refined auxiliary power consumption estimates, and implementation of a capital, annual, and levelized cost model.

The present model consists of slurry preparation units, a gasification unit, high temperature gas cooling, particulate removal and ash removal, low temperature gas cooling unit, fuel gas saturator and acid gas removal section, byproduct sulfur production, and combined-cycle power system as shown in Figure 3.1. In addition to these units, the model also incorporates auxiliary support facilities such as those that collect and treat utility waste water.

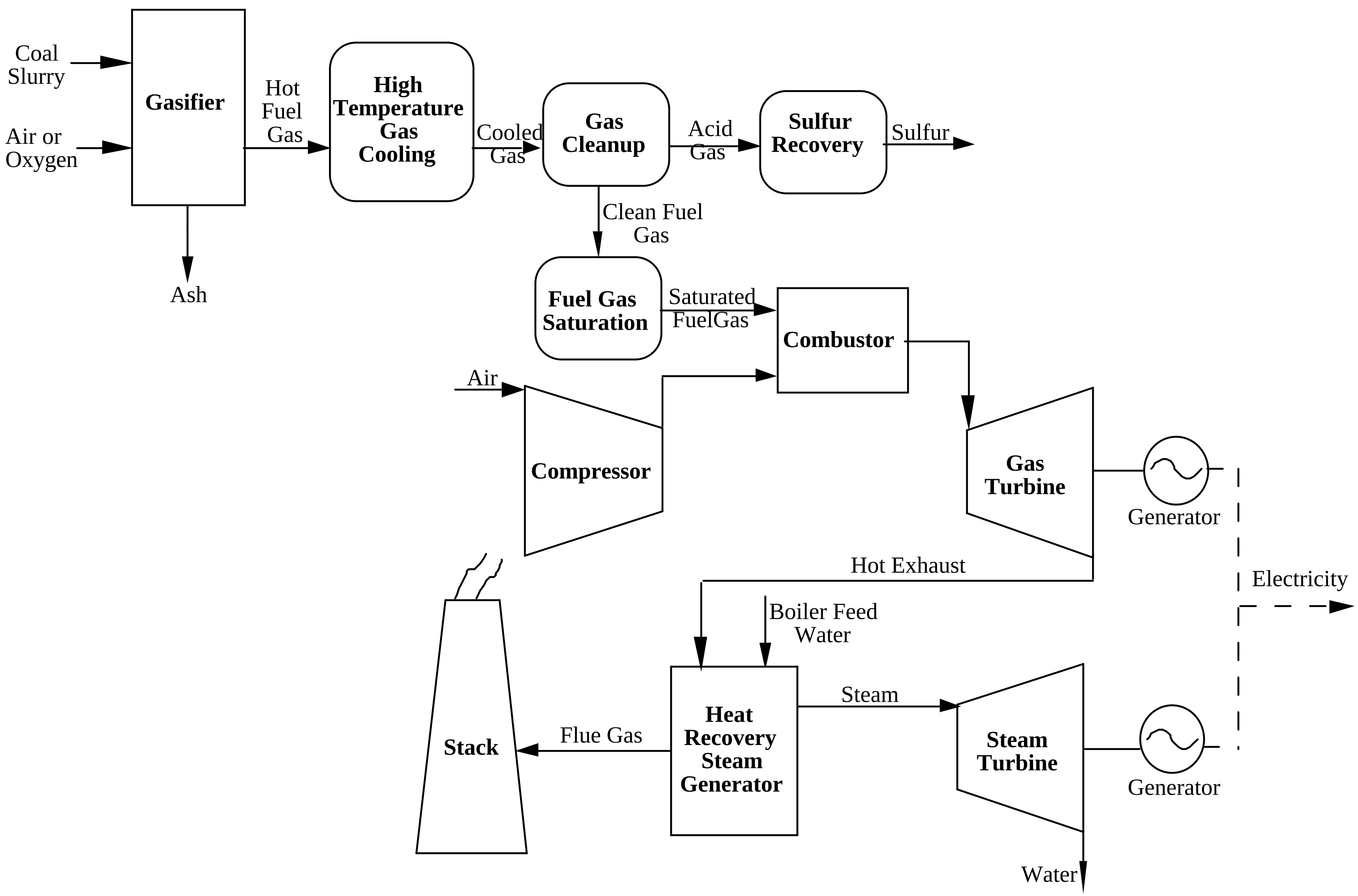

Figure 3.1 IGCC System

3.2 Major Process Sections in the Radiant and Convective IGCC Model

The major flowsheet sections in the process are described below. Each major process section is referred to as a flowsheet. Within each flowsheet, unit operation models represent specific components of that process area. There are user-specified inputs regarding key design assumptions for each unit operation model. The numerical values of the design assumptions are presented in this chapter. However, a user could substitute other values for these to reflect other design alternatives.

3.2.1 Coal Slurry and Oxidant Feed to the Gasifier

In this section the approach used to model slurry and oxidant feed to the gasifier is described. The ASPEN performance simulation model accepts user input regarding the characteristics of the coal assumed as a gasifier feedstock. The base case assumption regarding the coal composition is given in Table 3.1 for a typical Illinois No. 6 bituminous coal. The coal is modeled as part of a coal-water slurry, such that the slurry contains 66.5 weight percent solids.

Figure 3.2 illustrates the mass flows in the gasification process area, while Table 3.2 describes the unit operations that are modeled in this process area. The coal slurry flows through a pump, modeled as a unit operation of the type "PUMP" with a block identification of "SLURPUMP", to a user-defined unit operation identified as "COALCONV". The slurry pump serves to raise the pressure of the slurry to 650 psia, which is high enough for introduction into the gasifier, which operates at 615 psia in the base case scenario. The COALCONV block serves to decompose the coal into its constituent elements. The portions of the coal that represent soot and slag are modeled as being removed from the coal by the blocks "MAKESLAG" and "MAKESOOT". MAKESLAG calculates the heat required to convert a portion of the coal to slag and MAKESOOT calculates the heat required to convert a portion of the coal to soot. Both of the heat streams are directed to the gasifier main reactor modeled by the block "GASIFIER". The equations used in MAKESOOT and MAKESLAG are, respectively,

$$0.012\ C + 0.852\ ASH \rightarrow SOOT \tag{3-1}$$

$$0.0007\ C + 0.992\ ASH \rightarrow SLAG \tag{3-2}$$

The oxidant feed is modeled to consist of 95 percent pure oxygen at 250 ^{o}F and 734 psia. The mass flow rate of oxidant is modeled by a design specification, SETOXYG. SETOXYG varies the flow of oxidant such that the heat loss from the gasifier is less than one percent of the total heat input to the gasifier. Thus, the ASPEN model calculates the oxygen flow required to obtain the user specified gasifier outlet temperature and to overcome this heat loss. The coal slurry and oxidant feed are mixed in the unit operation block GASIFMIX and sent to the gasification unit modeled by the equilibrium reactor unit operation block GASIFIER.

Table 3.1 Proximate and Ultimate Analysis of the Base Case Illinois No.6 Coal

Proximate Analysis	Wt-%, run-of-mine basis
Moisture	10.00
Fixed Carbon	48.87
Volatile Matter	32.22
Ash	8.91
Ultimate Analysis	Wt-%, dry basis
Carbon	69.62
Hydrogen	5.33
Nitrogen	1.25
Sulfur	3.87
Oxygen	10.03
Ash	9.90
Ash Fusion Temperature, ^{o}F	2,300
Higher Heating Value, BTU/lb	12,774

3.2.2 Gasification

The coal slurry and oxidant feed are delivered to the gasifier burners where gasification takes place. The gasifier is modeled to operate at a design pressure of 615 psia and a design temperature of 2400 ^{o}F. The operating temperature is sufficiently higher than the ash fusion temperature of 2300 ^{o}F of the Illinois No. 6 coal to cause the ash to become molten and separate out easily from the raw gas. The unit operation block

GASIFIER simulates the gasification process. A portion of the coal feed burns, providing heat for the endothermic gasification reactions that result in the formation of CO, CO_2, H_2, CH_4, and H_2S. The chemical reactions modeled in the equilibrium gasifier reactor model are:

$$C + 2H_2 \rightarrow CH_4 \tag{3-3}$$

$$C + H_2O \rightarrow CO + H_2 \tag{3-4}$$

$$CO + H_2O \rightarrow CO_2 + H_2 \tag{3-5}$$

$$CH_4 + 1.5\,O_2 \rightarrow CO + 2H_2O \tag{3-6}$$

$$2CO + O_2 \rightarrow 2CO_2 \tag{3-7}$$

$$S + H_2 \rightarrow H_2S \tag{3-8}$$

$$N_2 + 3H_2 \rightarrow 2NH_3 \tag{3-9}$$

$$CO + H_2S \rightarrow COS + H_2 \tag{3-10}$$

$$Ar \rightarrow Ar \tag{3-11}$$

Equations (3-3), (3-4), and (3-5) are the primary gasification reactions. Equation (3-3) is an exothermic reaction and is known as methanation. The formation of methane increases the heating value of the product gas. Equations (3-4) and (3-5) are endothermic reactions and are known as watergas and water gas shift reactions respectively, leading to the formation of hydrogen. Equation (3-6), in series with Equation (3-3), represents the partial combustion of coal and Equation (3-7) in sequence with Equations (3-3) and (3-4), models the complete oxidation of coal.

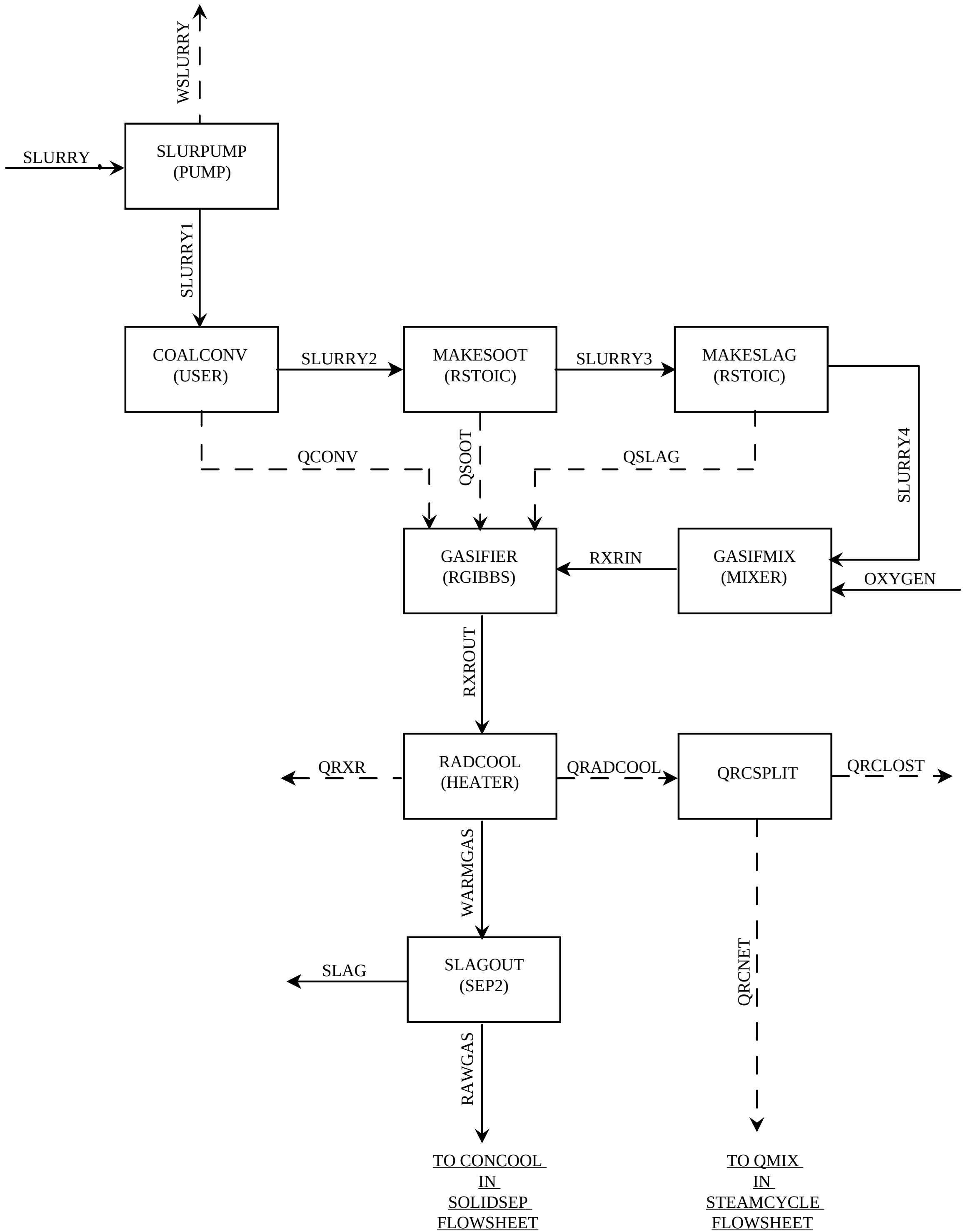

Figure 3.2 Gasification Flowsheet

Table 3.2 Gasification Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	SLURPUMP (PUMP)	TYPE=2 Pressure = 650 psia Efficiency = 0.65	This block simulates Coal-Water Slurry Pump, which delivers slurry to the gasifier burners.
2	COALCONV (USER)		This block decomposes coal into its elements using the subroutine USRDEC
3	MAKESOOT (RSTOIC)	Temperature = 59 ^oF Pressure drop = 0 psia	Simulates the stoichiometric reaction, which produces soot based on the coal's ultimate analysis.
4	MAKESLAG (RSTOIC)	Temperature = 59 ^oF Pressure drop = 0 psia	Simulates the stoichiometric reaction, which produces slag based on the coal's ultimate analysis.
5	GASIFMIX (MIXER)		Represents a Mixer that mixes the coal slurry and the oxidant feed.
6	GASIFIER (RGIBBS)	Temperature = 2400 ^oF Pressure = 615 psia NAT = 6 NPHS = 1 NPX = 2 NR = 9 IDELT = 1	This block simulates the stoichiometric reactions associated with the Gasifier Reactor.
7	RADCOOL (HEATER)	Temperature = 1500 ^oF Pressure = 613 psia	Simulates a Radiant Cooler which lowers the temperature of the syngas to 1500 ^oF from 2400 ^oF
8	QRCSPLIT (FSPLIT)	FRAC QRCLOST = 0.06 RFRAC QRCNET = 1.0	This block is used to indicate that some amount of heat is lost from the Radiant Cooler.

(continued on next page)

Table 3.2. Continued

9	SLAGOUT (SEP2)	COMP FRAC COAL = 1.0 ASH = 1.0 SLAG = 1.0 SOOT = 0.0	This block places slag into the Gasifier bottoms stream.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

3.2.3 High-Temperature Gas Cooling and Particulate Removal

The crude gas leaving the gasificaton unit is at a temperature of 2400 ^{o}F. This hot gas enters the radiant syngas coolers, simulated by the block "RADCOOL", where it is cooled by generating high pressure (1545 psia) saturated steam through recovery of high-level sensible heat. In the actual system, molten slag entrained in the hot gas from the gasifier drops into a water quench pool at the bottom of the radiant gas coolers, where it is cooled and removed for disposal. The blocks and streams modeled in this section are shown in Table 3.2 and Figure 3.2. RADCOOL simulates cooling of the syngas to a temperature of 1500 ^{o}F. The cooled syngas flows to the SLAGOUT block, which simulates the separation of slag from the raw gas. The block QRCSPLIT is used to model sensible heat lost due to radiation. A default assumption is that six percent of the total heat gained by cooling the syngas from the gasifier to 1500 ^{o}F is lost to the surroundings due to radiative heat transfer from the hot walls of the heat exchanger. QRCPLIT splits the heat stream QRADCOOL from RADCOOL into heat streams QRCNET and QRCLOST. QRCLOST is set to six percent of QRADCOOL.

The cooled raw gas from the radiant gas coolers is sent through a separating block, SLAGOUT, which separates the slag from the rawgas. Carbon conversion indicates the amount of carbon in the fuel that is in the syngas. The carbon loss refers to the carbon in the slag, and it is specified as one of the parameters of SLAGOUT. The raw gas, removed of slag, is further cooled to 650 ^{o}F in the vertical convective syngas coolers, simulated by block CONCOOL as shown in Table 3.3 and Figure 3.3. The heat stream leaving CONCOOL is modeled by QCONCOOL. QCONCOOL is a heat stream, obtained by transferring heat for the cooled syngas, which exits CONCOOL at 650 ^{o}F and an output pressure of 603 psia. QCONCOOL is used for generating additional high pressure (1545 psia) saturated steam to be used in the steam cycle. The cooled raw gas from the convective coolers, modeled by CONGAS, is further cooled to 403 ^{o}F by a gas-gas heat exchanger, simulated by the GASCOOL block. QGASCOOL models the heat stream leaving the GASCOOL block. QGASCOOL is used for simulating the reheating of the saturated fuel gas, which enters the gas turbine combustor.

Figure 3.3 and Table 3.3 illustrate the particulate scrubbing sections of the model. The cooled raw gas, which contains particulate matter, enters the particulate scrubbing unit, modeled by the unit operation block PARTSCRB. The solids in the raw gas are removed through contact with recycled condensate, modeled by the stream CONDSATE, from the low-temperature gas cooling section and makeup water. The scrubbed gas, modeled by the NH3FREE stream, then enters the low-temperature gas cooling section.

The solids flow to the ash dewatering unit, simulated by block WWSEP, where the stream MIXEDWW is filtered to yield an ash cake and water.

A design specification is used to set the flowrate of CONDSATE equal to the flowrate of the stream ALLCOND. ALLCOND is the water reclaimed from the low temperature gas cooling section and sent to the particulate matter scrubbing unit. CONDSATE is the name of the water stream that is an input to the particulate matter scrubbing unit. The water stream reclaimed from the scrubbing unit, PURGEH2O, is recycled to the slag quench pool, particulate scrubbing unit and to the process water treatment unit where the recycled water purges to avoid build up of particles such as chlorides and soot particles in the treatment unit. The purged water is treated in the treatment unit. The treated water is used as slurry water and cooling tower makeup.

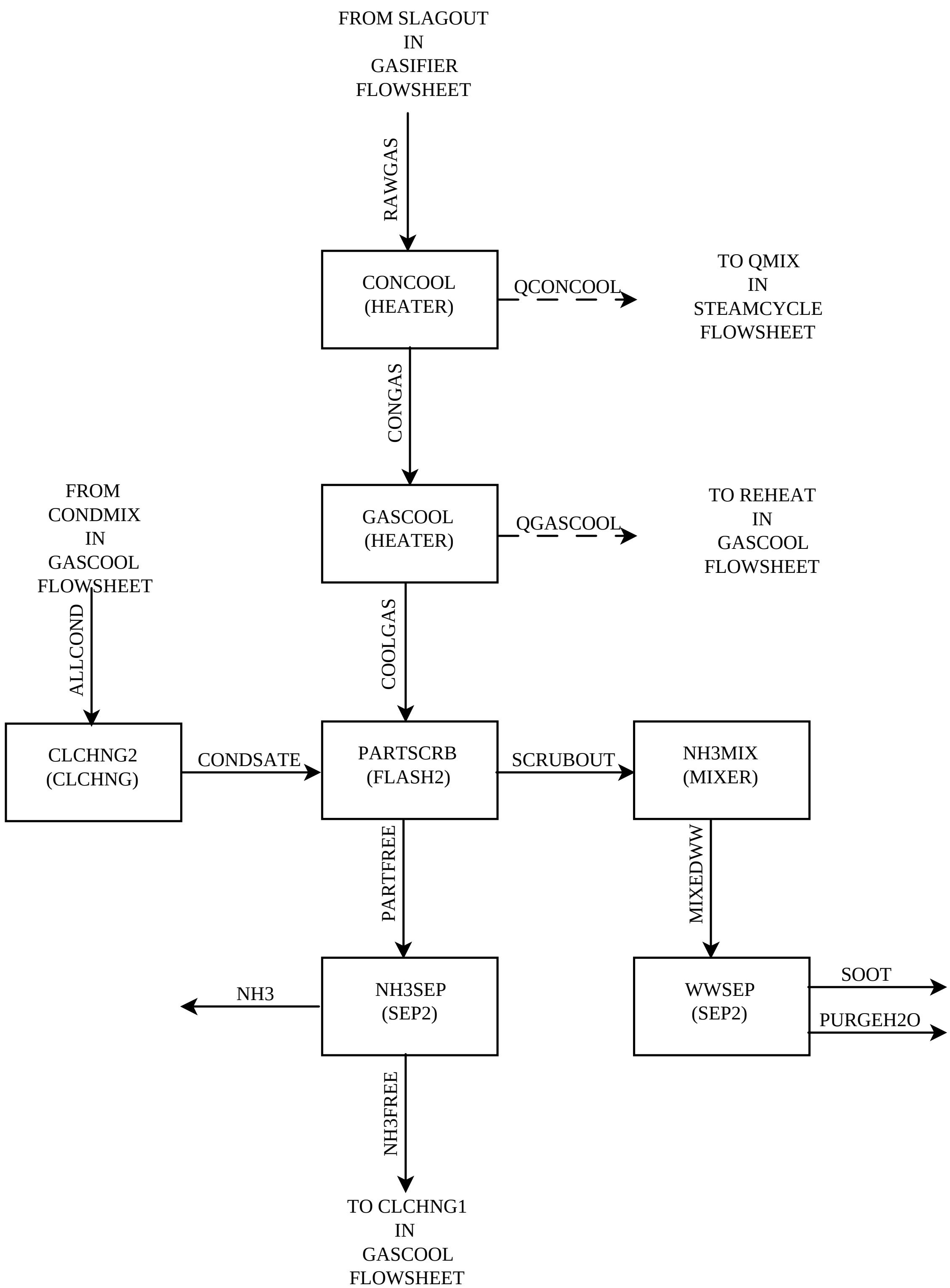

Figure 3.3 Solids Separation Flowsheet

Table 3.3 High-Temperature Gas Cooling (Solids Separation) Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	CONCOOL (HEATER)	Temperature = 650 ^oF Pressure = 603 psia	Simulates a Convective Syngas Cooler.
2	GASCOOL (HEATER)	Temperature = 403 ^oF Pressure = 598 psia	Simulates a Fuel Gas Reheater - Hot Side.
3	CLCHNG2 (CLCHNG)		
4	PARTSCRB (FLASH2)	Q = 0 Pressure = 572	This block, which simulates a Particulate Scrubber, removes soot from gas stream.
5	NH3MIX (MIXER)		This block takes the scrubbed bottoms of the particulate scrubber and mixes it.
6	NH3SEP (SEP2)		Simulates the absorption of ammonia in the syngas into scrubber water.
7	WWSEP (SEP2)		This block separates soot and water from the mixed water from the NH3MIX block.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

3.2.4 Low-Temperature Gas Cooling and Fuel Gas Saturation

The low temperature gas cooling section is shown in Figure 3.4 and Table 3.4. The scrubbed gas, NH3FREE, is cooled by circulating saturator water in a heat exchanger, simulated by block COOL1. The gas is further cooled to 130 ^oF by a heating vacuum condensate, which is simulated by heater block COOL2. Block COOL3 models the cooling of the raw gas from 130 ^oF to 101 ^oF in the trim cooler by heat exchange with cooling water. The condensate from the heat exchangers is collected in the condensate collection drum, the latter of which is simulated by mixer block CONDMIX. The cooled gas, stream COLDGAS, is sent to the Selexol acid gas removal unit.

The Selexol unit separates the stream COLDGAS into streams CLEANGAS, ACIDGAS, and FLASHGAS. ACIDGAS, containing 97.6 percent of H_2S by volume is sent to the mixer, CLAUSMIX, in the Claus plant. The stream FLASHGAS is sent to the mixer, BSMIX, in the Beavon-Stretford tail gas treatment plant.

The clean gas from the Selexol process, modeled by the stream CLEANGAS, enters the saturation unit at 85 ^{o}F and 429 psia. The details of the modeling of the fuel gas saturation unit are shown in Figure 3.5 and Table 3.4. The required amount of water to be added to clean gas to make its moisture content 28.2 % by weight is calculated by a FORTRAN block SATURH2O. SATURH2O obtains the mass flow of clean gas entering the saturator block and calculates the required saturated fuel gas mass flow, modeled by the stream SATGAS. SATGAS is required to be at a temperature of 347 ^{o}F which is achieved by using a design specification SETSATR. SETSATR calculates the required amount of hot water entering the saturation unit through the block FAKESPLT. This block splits the hot water stream, HOTH2O into HOTH21 and SATCOM streams. HOTH21 is cooled by a heat exchanger, FAKECOOL, to a temperature of 235 ^{o}F. SATCOM and CLEANGAS enter FAKEMIX, which simulates a mixer. The saturated fuel gas from FAKEMIX, SATGAS1, is heated to the required temperature of 347 ^{o}F in the block FAKEHEAT by QHEATS, the heat stream leaving FAKECOOL. As shown in Figure 2.3, the required amount of circulating water to the saturation unit is maintained by heating the circulating water in heat exchanger COOL1. A slip stream of high pressure boiler feed water (BFW) is used to supply the necessary heat to the circulating water coming out of GASCOOL. The slipstream BFW, the circulating water from GASCOOL and COOL1 combine to form the hot water, HOTH2O which enters the block FAKESPLT. The fuel gas exits the saturator at 347 ^{o}F with a moisture content of 28.2 weight percent and is reheated to 570 ^{o}F in the block REHEAT with the help of the heat stream QGASCOOL from the high temperature gas cooling section. The reheated fuel gas, the stream GTFUEL, flows to the gas turbine combustors.

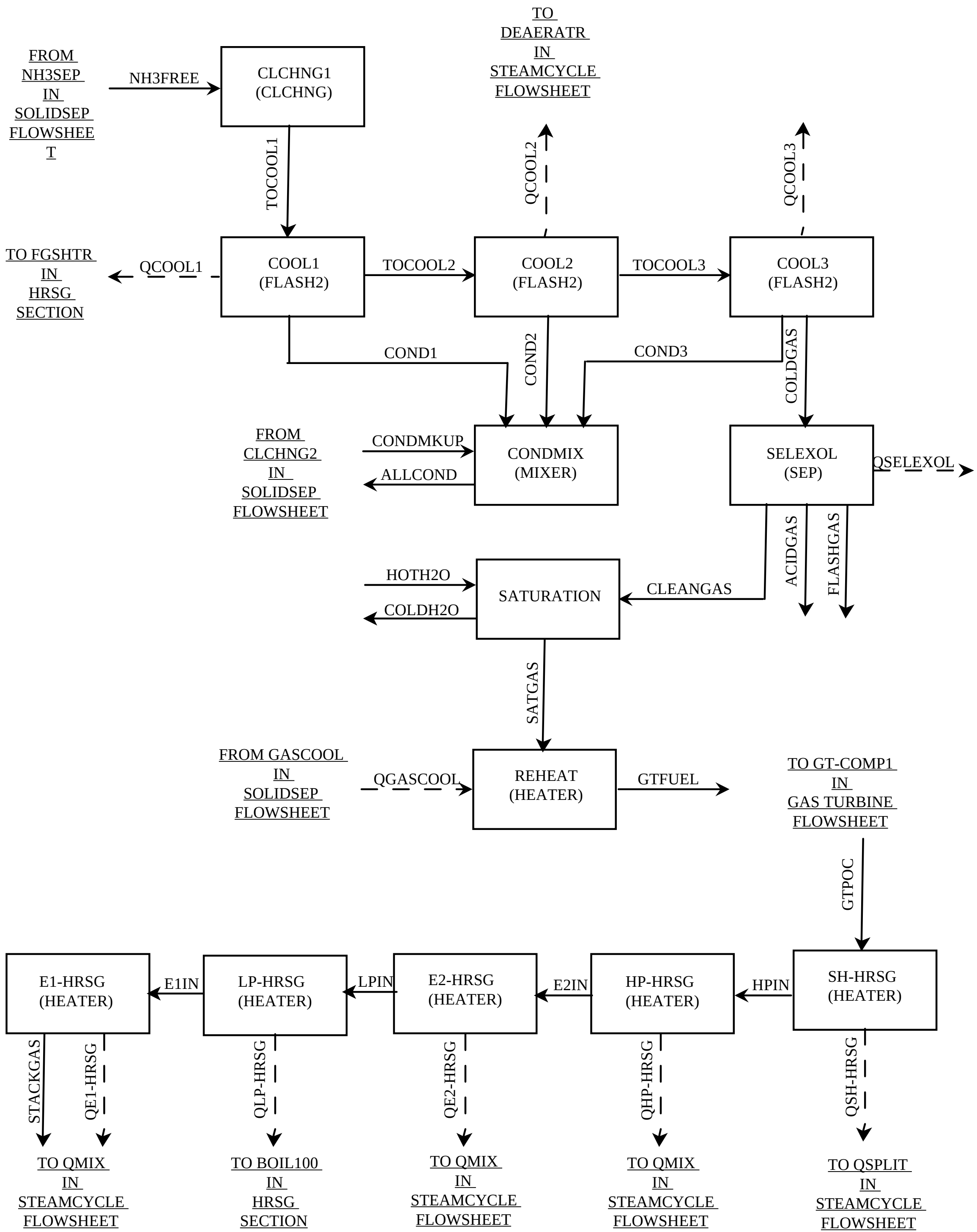

Figure 3.4 Low Temperature Gas Cooling Flowsheet

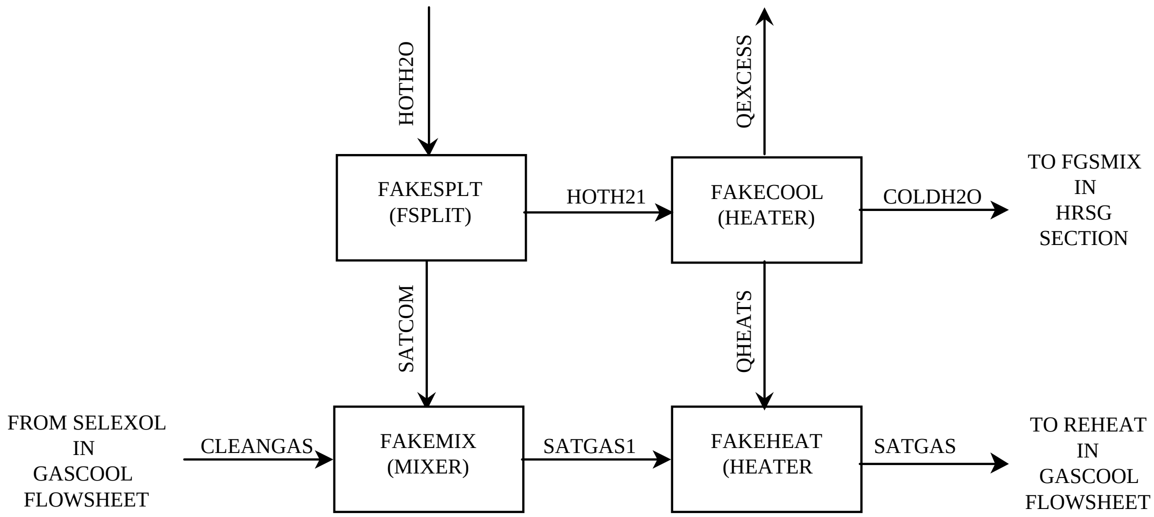

Figure 3.5 Fuel Gas Saturation Flowsheet

Table 3.4 Low Temperature Gas Cooling Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	CLCHNG1 (CLCHNG)		This block changes stream class from MIXCINC to Conventional.
2	COOL1 (FLASH2)	Temperature = 262 °F Pressure = 567 psia	This block simulates a heat exchanger, which reduces the temperature of the syngas to 262 °F from 323 °F across a pressure drop of 5 psia.
3	COOL2 (FLASH2)	Temperature = 130 °F Pressure = 562 psia	This block simulates a heat exchanger, which reduces the temperature of the syngas to 130 °F from 262 °F across a pressure drop of 5 psia.

(continued on next page)

Table 3.4. Continued

4	COOL3 (FLASH2)	Temperature = 101 °F Pressure = 557 psia	This block simulates a heat exchanger, which reduces the temperature of the syngas to 101 °F from 130 °F across a pressure drop of 5 psia.
5	CONDMIX (MIXER)		This block simulates the mixing of all condensates in this section.
6	SELEXOL (SEP)	CLEANGAS T = 85 °F P = 429 psia ACID GAS T = 120 °F P=22 psia FLASH GAS T = 58 °F P = 115 psia	This block separates the syngas into Acid Gas, Flash Gas, and Clean Gas.
7	FAKESPLT (FSPLIT)		This block splits the HOTH2O required for saturation of fuel gas to 28.2 wt % moisture. The split is set by the FORTRAN block SATURH2O.
8	FAKECOOL (HEATER)	Temperature = 235 °F Pressure = 429 psia	Simulates the cooling of the hot BFW.
9	FAKEMIX (MIXER)		Simulates the mixing of the CLEANGAS and SATCOM.
10	FAKEHEAT	Temperature = 347 °F Pressure = 419 psia	Simulates the heating of the saturated gas such that the fuel gas temperature before entering REHEAT is 347 °F.
11	REHEAT (HEATER)	Pressure = 414 psia	Simulates a Fuel Gas Reheater - Cold Side.
12	SH-HRSG (HEATER)	Temperature = 743 °F Pressure drop = 0 psia	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.

(continued on next page)

Table 3.4. Continued

12	HP-HRSG (HEATER)	Temperature = 641 ^oF Pressure drop = 0 psia	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
13	E2-HRSG (HEATER)	Temperature = 401 ^oF Pressure drop = 0 psia	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
14	LP-HRSG (HEATER)	Temperature = 366 ^oF Pressure drop = 0 psia	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
15	E1-HRSG (HEATER)	Temperature = 271 ^oF Pressure drop = 0 psia	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

3.2.5 Gas Turbine

The gas turbines represented in the model are assumed to be heavy duty "F" class systems similar to a General Electric MS7001F. The model developed here was designed to include appropriate details regarding the cooling air loss, the size of the gas turbine, and NO_x emission estimation in comparison to the original FETC model. The gas turbine has a multi-staged compressor, which compresses the air required for combustion, and increases the temperature and pressure of air. The compressor usually has several extraction points, from which some amount of compressed air is removed and is injected into the blades and vanes of the hottest turbine stages in order to cool the blades and vanes. The gas turbine combustor receives the syngas and the compressed air and combusts them. The hot exhaust gases are expanded in the turbine in several stages, represented in the model by three expanders.

3.2.5.1 Compression

Ambient conditions of the atmospheric air entering the gas turbine compressor are assumed to be 59 ^{o}F, 14.7 psia, and 60 percent relative humidity. These values are taken as defaults and can be changed by the user. The default compressor ratio is assumed to be 15.5, which is typical of heavy duty gas turbines (Farmer, 1997), resulting in a compressor outlet pressure of 227.85 psia. Figure 3.6 and Table 3.5 present the gas turbine model in detail.

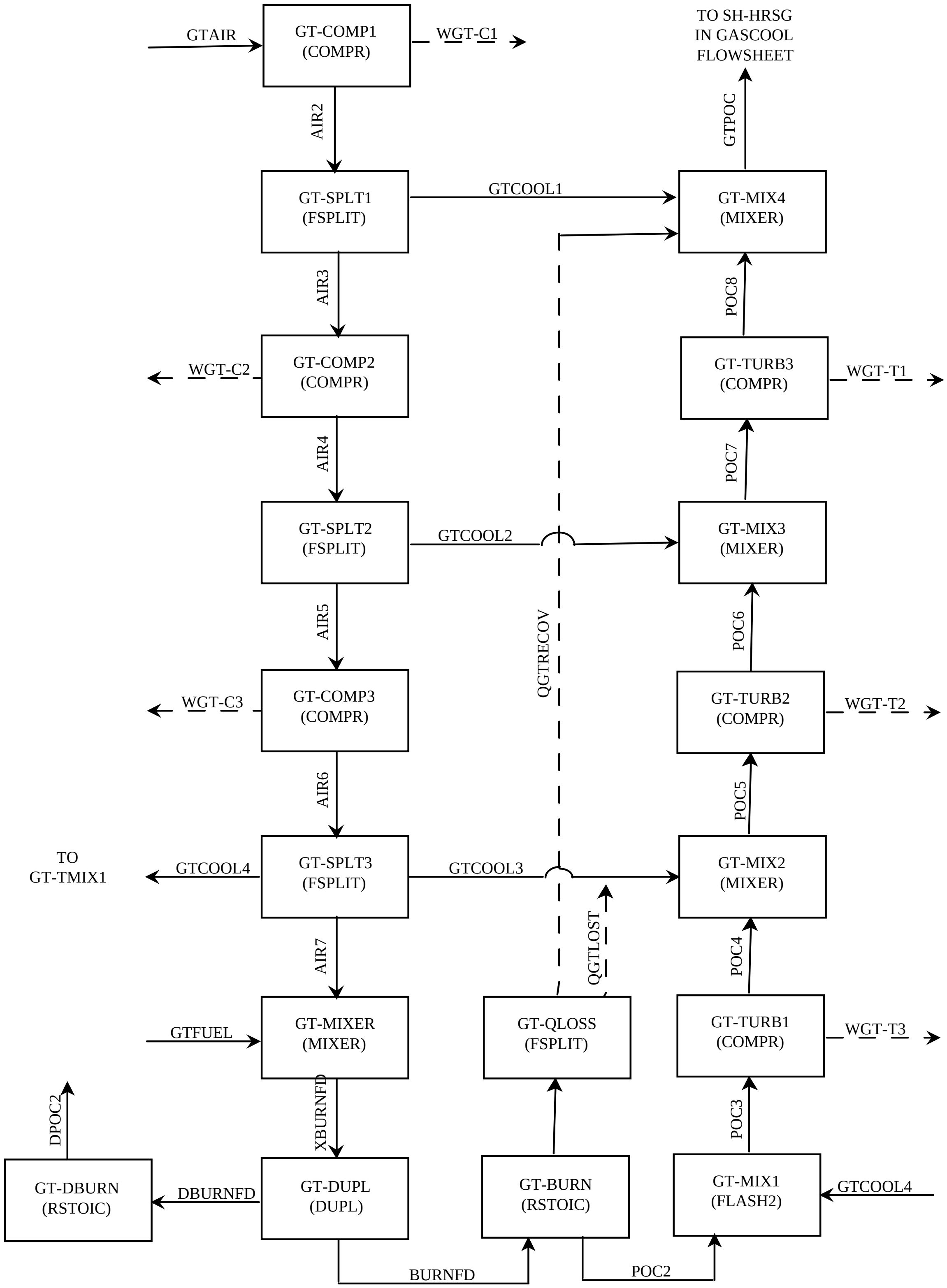

Figure 3.6 Gas Turbine Flowsheet

Table 3.5 Gas Turbine Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	GT-COMP1 (COMPR)	TYPE = 3 Pressure = 34.77 psia Isentropic Efficiency = 0.88	This block simulates a compressor, which compresses the air entering the Gas Turbine. The pressure and isentropic efficiency are set up by FORTRAN block STCTAIL.
2	GT-SPLT1 (FSPLIT)		This block splits the compressed air coming out of the block GT-COMP1 and directs one stream is used to cool the products of combustion of the Gas Turbine. This accounts for cooling the leakages and blockages.
3	GT-COMP2 (COMPR)	TYPE = 3 Pressure = 83.07 psia Isotropic Efficiency = 0.88	Similar to GT-COMP1
4	GT-SPLT2 (FSPLIT)		Similar to GT-SPLT1. This corresponds to 1st stage Rotor and 2nd stage Vane Cooling.
5	GT-COMP3 (COMPR)	TYPE = 3 Pressure = 227.85 psia Isentropic Efficiency = 0.88	Similar to GT-COMP1
6	GT-SPLT3 (FSPLIT)		Similar to GT-SPLT1. This corresponds to 1st stage Vane Cooling.
7	GT-MIXER (MIXER)	NPK = 1	This block simulates the mixing of the compressed air and expanded fuel gas.

(continued on next page)

Table 3.5. Continued

8	GT-DUPL (DUPL)		This block makes a copy of the mixed fuel+air inlet stream. It is used for calculating actual fuel heating value.
9	GT-BURN (RSTOIC)	Temperature = 2350 ^{o}F Pressure = 218.74 psia	Simulates the stoichiometric reactions that take place in Gas Turbine combustor.
10	GT-DBURN (RSTOIC)	Temperature = 2350 ^{o}F Pressure = 218.74 psia	Simulates the stoichiometric reactions that take place in a dummy Gas Turbine combustor.
11	GT-QLOSS (FSPLIT)	FRAC QGTLOST = 0.5 Frac QGTRECOV = 0.5	Simulates the loss of heat from the Gas Turbine combustor.
12	GT-MIX1 (FLASH2)	Temperature = 2350 ^{o}F Pressure = 218.74 psia ENT = 1.0	Simulates the mixing of cool air with the hot products of combustion.
13	GT-TURB1 (COMPR)	TYPE = 3 Pressure = 83.07 psia Isoentropic Efficiency = 0.89	Simulates a compressor for the expansion and subsequent cooling of the mixing of products of combustion and cool air.
14	GT-MIX2 (MIXER)	Pressure = 83.07 psia	Simulates the mixing of cool air with the hot products of combustion.
15	GT-TURB2 (COMPR)	TYPE = 3 Pressure = 34.77 psia Isoentropic Efficiency = 0.89	Simulates a compressor for the expansion and subsequent cooling of the mixing of products of combustion and cool air.
16	GT-MIX3 (MIXER)	Pressure = 34.77 psia	Simulates the mixing of cool air with the hot products of combustion.
17	GT-TURB3 (COMPR)	TYPE = 3 Pressure = 15.2 psia Isoentropic Efficiency = 0.89	Simulates a compressor for the expansion and subsequent cooling of the mixing of products of combustion and cool air.

(continued on next page)

Table 3.5. Continued

18	GT-MIX4 (MIXER)		Simulates the mixing of cool air with the hot products of combustion.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

The outlet pressure at the last compressor stage is estimated in the FORTRAN block STCTAIL based on the inlet pressure of the first stage compressor and the user specified pressure ratio, which is 15.5 in this case. The individual compressor stage outlets for the first, second, and third stages are estimated by the following relationships, respectively:

$$P_{c,1,o} = P_{ambient}\ PR^{\ 0.33} \tag{3-12}$$

$$P_{c,2,o} = P_{ambient}\ PR^{\ 0.67} \tag{3-13}$$

$$P_{c,3,o} = P_{ambient}\ PR \tag{3-14}$$

where,

PR = pressure ratio =15.5

$P_{ambient}$ = 14.7 psia

The compressors were modeled by three unit operation blocks, GT-COMP1, GT-COMP2, and GT-COMP3 with outlet pressures specified as 36.41 psia, 91.08 psia, and 227.85 psia allowing for some pressure loss. The isentropic efficiencies of each of the compressors is 0.81 as discussed in Section 3.2.5.7 Based upon these default assumptions, the discharge temperature of outlet air entering the gas turbine combustor is found to be 838 ^{o}F based upon simulation results from the ASPEN model. After each stage of compression, the compressed air is split into two or more streams. One stream undergoes further compression and the other streams represented by GT-COOL1, GT-COOL2, GT-COOL3, and GT-COOL4 are used for cooling the turbine blades after each expansion stage of the gas turbine.

The reheated fuel gas, GTFUEL and the compressed air, AIR7 enter the combustor modeled by the stoichiometric reactor block GT-BURN. The following chemical reactions are used in the block GT-BURN to simulate the combustion.

$$2\,CO + O_2 \rightarrow 2\,CO_2 \tag{3-15}$$

$$2\,H_2 + O_2 \rightarrow 2\,H_2O \tag{3-16}$$

$$CH_4 + 1.5\,O_2 \rightarrow CO + 2\,H_2O \tag{3-17}$$

$$2\,H_2S + 3\,O_2 \rightarrow 2\,H_2O + 2\,SO_2 \tag{3-18}$$

$$COS + 1.5\,O_2 \rightarrow CO_2 + SO_2 \tag{3-19}$$

$$2\,NH_3 + 2.445\,O_2 \rightarrow 0.1\,N_2 + 1.71\,NO + 0.09\,NO_2 + 3\,H_2O \tag{3-20}$$

$$N_2 + 1.05\,O_2 \rightarrow 1.9\,NO + 0.1\,NO_2 \tag{3-21}$$

These reactions represent the oxidation of the syngas components CO, H_2, CH_4, H_2S, COS, and NH_3. In addition, Equation (3-21) is used to model the formation of thermal NO and NO_2, while Equation (3-20) is used to model the formation of fuel-bound NO and NO_2 from NH3 in the syngas.

The firing temperature of a gas turbine is the temperature at which combustion process takes place in the combustor. The higher the firing temperature, the higher the temperature of the hot gases entering the turbine. The firing temperature of the gas turbine is constrained by the requirement that the turbine exhaust gas, GTPOC, has a temperature of 1120 ^{o}F or less to prevent damage to the turbine blades (Farmer, 1997). This constraint is met using a design specification, SETHRST, which is described Section 3.2.5.6.

The expansion of the hot products of combustion, stream POC2, leaving the combustor is modeled in three stages. Each of the three stages consist of a turbine, which are modeled by GT-TURB1, GT-TURB2, and GT-TURB3 and a mixer, which are modeled by GT-MIX1, GT-MIX2, GT-MIX3. In each of these stages, the hot gases are mixed with the cooler air coming from one of the blocks GT-SPLT1, GT-SPLT2, or GT-SPLT3 and then expanded in the turbine. The first, second, and third turbines have an outlet pressures of 91.08 psia, 36.41 psia, and 15.42 psia, respectively, and each has an isentropic efficiency of 0.919. The exhaust gases, GTPOC, enter the heat recovery steam generation (HRSG) unit.

The outlet pressure at each expander stage is estimated in FORTRAN block STCTAIL using the same method used for compression stages.

3.2.5.3 Engine Size Constraints

The overall mass flow in a gas turbine is typically limited by the turbine nozzle as discussed in Section 2.6. In the model, the mass flows through the gas turbine are constrained by the mass flow at the turbine inlet nozzle. This constraint enables the model to respond in a realistic manner to changes in fuel gas composition such as those because of fuel gas saturation. Specifically, as the fuel heating value decreases, the fuel mass flow increases and the compressor mass flow decreases in order to deliver the correct mass flow to the turbine inlet nozzle.

The flow at the inlet of the gas turbine expander is choked; that is, the Mach number of the gas stream is unity. The choked flow condition is assumed to hold regardless of the type of fuel used due to the large pressure ratio across the first stage turbine nozzle (Eustis and Johnson, 1990). The design specification TCHOKE sets the flow of hot air at the turbine inlet nozzle corresponding to choked flow conditions by varying the compressor inlet flow.

$$M_{flow} = P\,A\sqrt{\frac{MW}{T}\sqrt{\frac{\gamma}{R}}\left(\frac{2}{\gamma+1}\right)^{\frac{\gamma+1}{\gamma-1}}} \tag{3-22}$$

where,

M_{flow} = Maximum mass flow rate through the nozzle

P = total pressure

A = critical area where the flow is choked

MW = molecular weight of gas

T	= total temperature
R	= universal gas constant
γ	= ratio of specific heats for the gas

The mass flow rate of the ambient air entering the gas turbine combustor is initialized in the ASPEN input file. The mass flow rate of the ambient air is adjusted by TCHOKE to achieve a specified turbine nozzle gas mass flow rate. The choked mass flow is calculated based on a reference mass flow, adjusted for differences in pressure, temperature, and molecular weight, and assuming that the critical area and ratio of specific heats of exhaust gas for reference and actual case are constant. The reference mass flow is estimated based on a GE MS7001F firing syngas, with an exhaust mass flow of 3,775,000 lb/hr and assuming that 12 percent of the compressor air is diverted for gas turbine blade and vane cooling similar to previous studies (Frey and Rubin, 1991).

3.2.5.4 Estimation of Cooling Air Percentages

The cooling flows in the gas turbine are extracted from the discharge at multiple compressor stages to improve characterization of the energy penalty associated with cooling air (Frey and Rubin, 1991). As indicated in Figure 3.6 and Table 3.5, a portion of the total inlet air flow to the gas turbine combustor is directed to the first and second stage turbine inlets from the third stage compressor discharge. Similarly, a portion of air from the second discharge compressor is directed to the third stage turbine inlet and a portion of air from the first stage compressor discharge is mixed with the hot gases from the third stage turbine. The cooling air percentages were estimated by calibrating the model to the overall efficiency and output specifications for a typical heavy duty gas turbine and they are specified in the FORTRAN block AIRCOOL.

3.2.5.5 Introduction of Moisture into Fuel

The reheated fuel gas from the low temperature gas cooling section, at 570 ^{o}F with 28.2 weight percent moisture in the radiant and convective design, is introduced to the gas turbine combustor along with the compressed air. After combustion and expansion stages, the gas turbine exhaust gases are routed to the HRSG section.

3.2.5.6 Design Specifications and FORTRAN blocks

The design specifications used in the gas turbine model are TCHOKE, SETHRST, GT-HEAT and BURNTEMP. TCHOKE is used to adjust the gas turbine inlet air to achieve the choked flow constraint at the turbine nozzle inlet. SETHRST sets the expander exhaust gas temperature by varying the firing temperature of the gas turbine combustor.

At high exhaust gas temperatures, the gas turbine blades' lifetime can be reduced. To prevent possible damage to the gas turbine blades, the temperature of the gas turbine exhaust gas is controlled such that it is kept below 1120 ^{o}F. The control temperature of 1120 ^{o}F is obtained from published data (Holt, 1998). This is achieved by varying the gas turbine firing temperature in the SETHRST design specification until the desired expander exhaust gas temperature is obtained.

The design specification, TCHOKE was discussed in Section 3.2.5.3.

GT-HEAT sets the combustor heat loss to four percent of the heat input to the gas turbine combustor by varying the fuel flow. In this design specification, the mass flow of coal is varied until the desired combustor heat duty is achieved. The unit operation block GT-MIX1, mixes the hot exhaust gases from the gas turbine combustor with cool air from the compression stages of the gas turbine before sending the hot gases to the first stage of gas turbine expanders. BURNTEMP sets the firing temperature of the gas by ensuring that there is no heat loss from the mixer, GT-TMIX1, after it mixes the hot exhaust gas from the combustor with the cool air from the first stage of compression.

The FORTRAN block STCTAIL initializes parameters such as temperatures, pressures, and conversion efficiencies for a wide range of flowsheet unit operations, such as the gas turbine. GTHOC and AIRCOOL are FORTRAN blocks associated with the gas turbine, with the former calculating the actual fuel heating value which is used for estimating the gas turbine efficiency, and the latter setting the gas turbine internal cooling air flows to fractions of the total inlet airflow. These fractions were obtained by the calibration of the gas turbine.

3.2.5.7 Calibration of the Gas Turbine Model

In order to calibrate the gas turbine model, a simple cycle system was simulated for natural gas and one gas turbine and key input assumptions in the simulation were varied in order to match published specifications for the exhaust gas temperature, simple cycle efficiency, and net power output for a commercial gas turbine. The simple cycle efficiency, power output, and exhaust gas temperature vary with the isentropic efficiencies of compressors and expanders of the gas turbine, as illustrated in Figure 3.7. The curves shown in the Figure 3.7 were obtained from sensitivity analysis of the simple cycle gas turbine model. For natural gas firing, published data are available for a "Frame 7F" type of gas turbine. For example, the published values for a General Electric MS7001F gas turbine are a simple cycle efficiency of 36.35 percent on a lower heating value basis, a power output of 169.9 MW, an exhaust mass flow of 3,600,000 lb/hr, and an exhaust gas temperature of 1,116 ^{o}F (Farmer, 1997). The required turbine isentropic efficiency is selected from Figure 3.7 (a) based upon the desired exhaust temperature; in this case, an isentropic efficiency of 87.2 percent was selected. A compressor isentropic efficiency of 91.8 percent is selected based on Figure 3.7 (b) in order to obtain the correct simple cycle efficiency. The reference mass flow at the turbine inlet is adjusted to 3,470,000 lb/hr obtain the desired power output. The estimated power output of 170.0 MW, obtained from the ASPEN gas turbine model with the selected values of isentropic efficiencies, is within 0.11 percent of the published data. A similar procedure was used to calibrate the gas turbine to data for a coal gasification application. The isentropic efficiencies obtained in the case of syngas are 0.81 and 0.919 for gas turbine compressors and gas turbine expanders respectively.

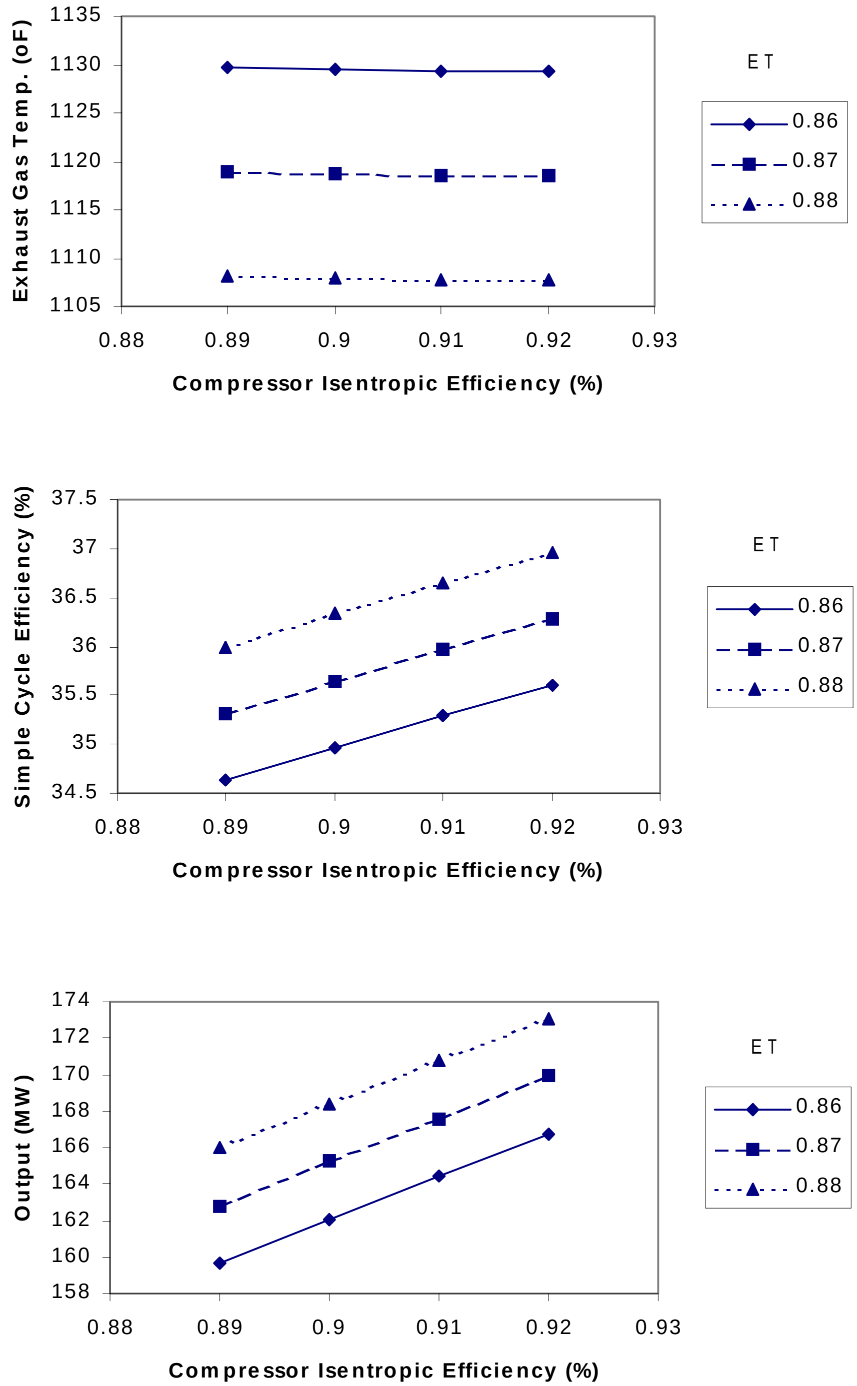

Figure 3.7 Plots of (a) Exhaust Gas Temperature , (b) Simple Cycle Efficiency, and (c) Output versus Gas Turbine Compressor Isentropic Efficiency.

Note: ET = Gas Turbine Expander Isentropic Efficiency

3.2.6 Steam Cycle

The steam cycle section of the IGCC consists of the heat recovery steam generator, auxiliaries and steam turbine. The steam cycle is shown in Table 3.6 and Figure 3.8.

3.2.6.1 Heat Recovery Steam Generation (HRSG)

The operations of the HRSG are to preheat boiler feed water, reheat intermediate pressure steam, supplement high pressure and 100 psia steam generation, and to superheat high pressure steam. The HRSG is arranged in the following order and shown in detail in Table 3.7 and Figure 3.9.

1. Superheater and reheater in parallel,
2. High pressure evaporator,
3. Economizer,
4. 100 psia boiler, and
5. Economizer.

The HRSG consists of a superheater at a pressure of 1465 psia and a temperature of 997 ^{o}F, a reheater at 997 ^{o}F, two economizers, a high pressure boiler, and a low pressure boiler. The inlet steam to the high pressure economizer and the makeup water for steam generation is initialized in the ASPEN input file through FORTRAN block SETSTEAM. The low pressure boiler is used to produce steam for the deaerator for the flue gas leaving the economizer at 366 ^{o}F. The heat losses in the HRSG process are accounted for through block QSPLIT shown in Table 3.6 and Figure 3.8

The hot exhaust gases from the gas turbine section, represented by GTPOC, are cooled by a series of heat exchangers, modeled by blocks SH-HRSG, HP-HRSG, E2-HRSG, LP-HRSG, and E1-HRSG in that order and are illustrated in Figure 3.4. The heat streams obtained from three of the blocks, namely E1-HRSG, E2-HRSG, and HP-HRSG are mixed in a mixer, simulated by QMIX. The heat stream from SH-HRSG, QSH-HRSG is split into three heat streams by the block QSPLIT. One heat stream is discarded as heat lost, one of the heat streams, QREHEAT is diverted to block TURBHEAT in steam turbine section shown in Figure 3.11, and the remaining heat stream, QSUPER, is sent to the block QMIX.

The total heat from the QMIX block, QTOTHRSG, is sent to the block ECONOMZR which simulates a heat exchanger. ECONOMZR heats a stream of water to a temperature of 553 ^{o}F. The mass flow of the stream of water, TOECON is calculated by the FORTRAN block SETSTEAM. The remaining amount of heat available is sent to block HPBOILER which simulates a high pressure steam boiler in HRSG. The steam generated by HPBOILER enters the superheater, SUPERHTR and generates superheated steam at a temperature of 997 ^{o}F. which is sent to a high pressure (350 psia) steam turbine, simulated by block TURB350 as shown in Figure 3.11.

The low pressure (1 psia) steam generated by the block TURB1, representing a steam turbine, is cooled by a heater simulated by block CONDENSR, as shown in Figure

3.8. The condensate from CONDENSR is pumped to 25 psia and delivered as WATER25 to a deaerator, simulated by the block DEAERATOR. DEAERATOR mixes the various condensates from the auxiliaries section, stream WATER25 and makeup water, which is required to makeup for the water sent to the fuel saturation unit from the steam cycle section. The mixed condensate, represented by DEAERH2O is sent to a block H2OSPLIT which simulates the splitting of the total condensate to streams TOECON, TOB100, TO565PSI, and TO65PSI. The ratios of the split are calculated by the FORTRAN block SETSTEAM.

Streams TOECON and TOB100 are sent to the blocks ECONMZR and BOIL100, respectively, in the HRSG section. BOIL100 simulates the generation of 100 psia steam. The steam from BOIL100 is split by the block SPLIT100 into streams SLXSTM and STM100, both of which are sent to the auxiliaries section shown in Figure 3.10. The unit operation blocks of the auxiliaries section are listed in Table 3.8.

The water streams TO565PSI and TO65PSI from the block H2OSPLIT are also sent to the auxiliaries section. The block CLAUS565 in the auxiliaries section heats the stream TO565PSI and generates steam of 565 psia pressure which is sent to the block TURBREHT and is further heated by the heat stream QREHEAT to a temperature of 996 ^{o}F.

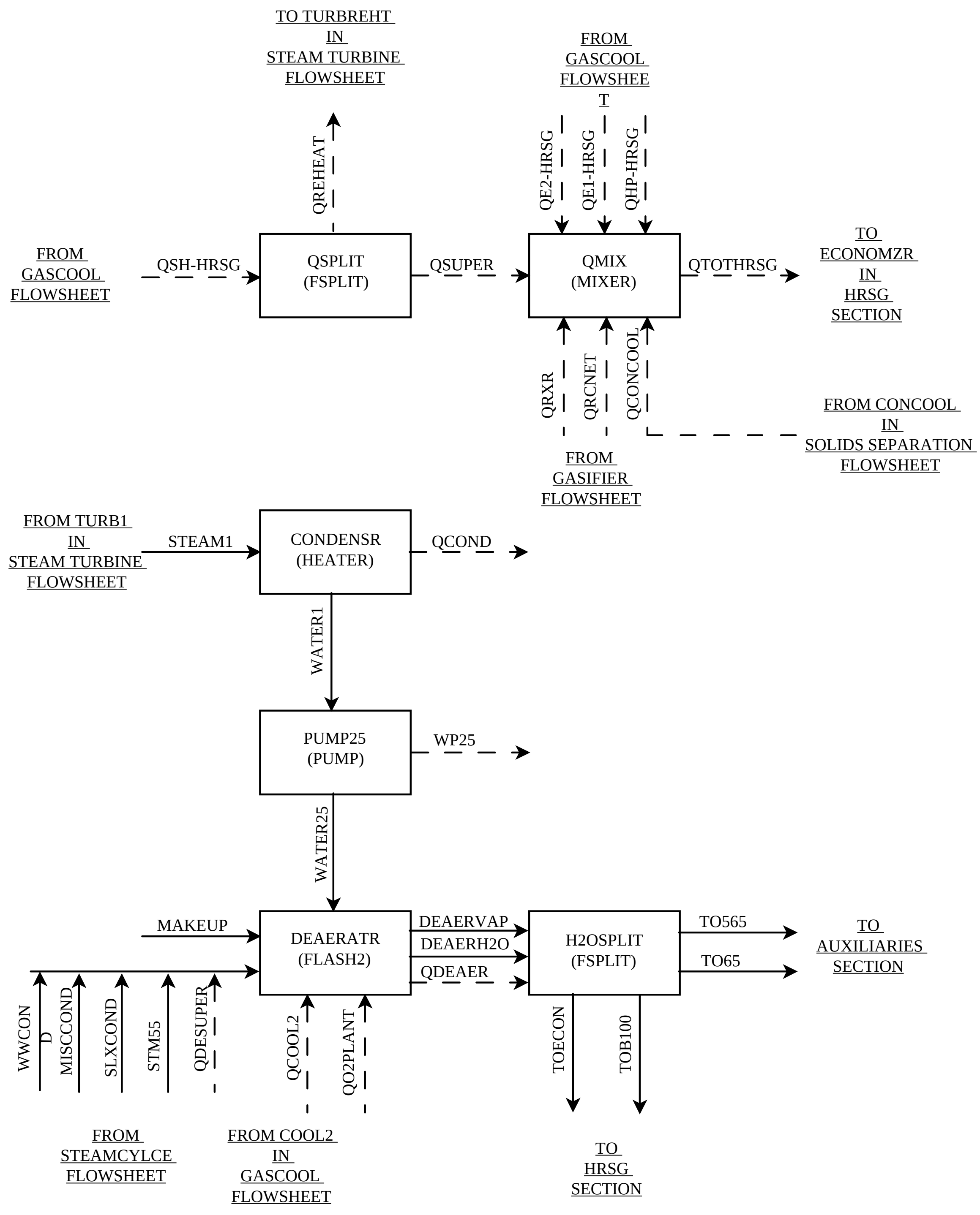

Figure 3.8 Steam Cycle Flowsheet

Table 3.6 Steam Cycle Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	QSPLIT (FSPLIT)	FRAC QRADROSS 0.03 QREHEAT 0.0388 RFRAC QSUPER 1.0	Simulates the radiation losses in the HRSG and diverts QREHEAT to REHEAT in HRSG section.
2	QMIX (MIXER)		Simulates the mixing of the various heat stream in the HRSG used in the calculation of superheated steam mass flow.
3	CONDENSR (HEATER)	Pressure = 1 psia Vfrac = 0	Simulates the block which heats the steam which comes out of the Steam Turbine section.
4	PUMP25 (COMPR)	TYPE = 1 Pressure = 25 psia	Simulates a pump which delivers the condensate to the deaerator.
5	DEAERATOR (FLASH2)	Pressure = 25 psia Vfrac = 0	Simulates the mixing of the condensates and steam.
6	H2OSPLIT (FSPLIT)	MOLE_FLOW TOECON 1.0 TOB100 1.0 TO565PSI 1.0 TO65PSI 1.0	Simulates the splitting of the total condensate into the required ratios in which the condensate will be sent to various blocks.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

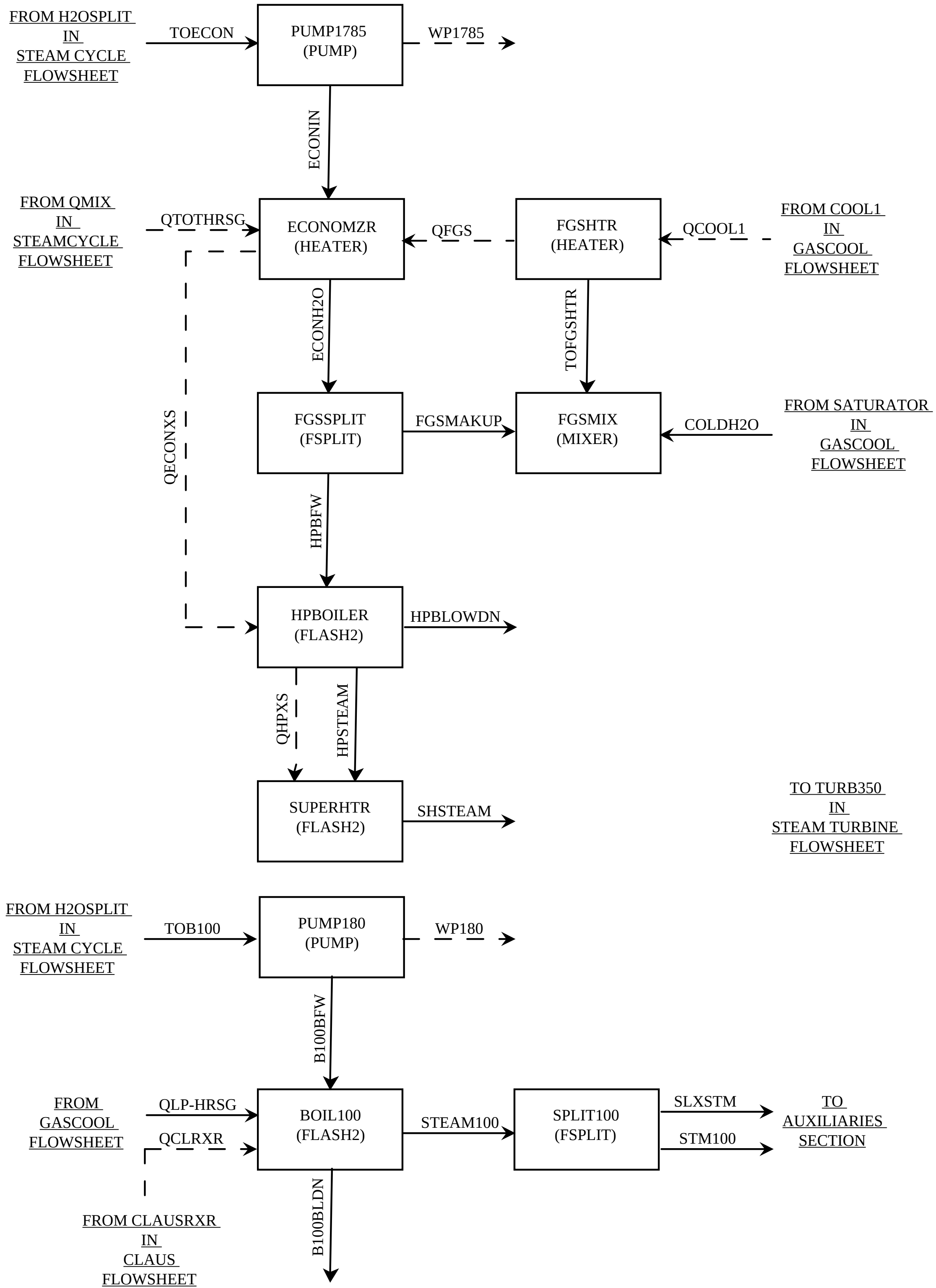

Figure 3.9 HRSG Section Flowsheet

Table 3.7 HRSG Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	PUMP1785 (COMPR)	TYPE = 1 Pressure = 1785 psia	Simulates a pump which delivers condensate to the HRSG economizer.
2	ECONOMZR (HEATER)	Temperature = 553 °F Pressure = 1625 psia	Simulates economizers 1 and 2 of HRSG.
3	QECOSPLT (FSPLIT)	FRAC QECONXS 0.81 QECOREH 0.19	Simulates the splitting of the heat stream coming out the economizer block.
4	FGSSPLIT (FSPLIT)	MOLE-FLOW FGSMAKUP 1.0 RFRAC HPBFW 1.0	This block provides hot water for fuel gas saturator.
5	FGSMIX (MIXER)	Properties SYSOP3	Simulates a mixer which mixes makeup water and cold water from the SATURATR.
6	FGSHTR (HEATER)	Properties SYSOP3 Temperature = 366 °F Pressure drop = 0 psia	Simulates a heater which heats the makeup water to the SATURATR.
7	HPBOILER (FLASH2)	Pressure = 1545 psia Vfrac = 0.97	Simulates a high pressure steam boiler in HRSG.
8	SUPERHTR (HEATER)	Pressure = 1465	Simulates the steam superheater in HRSG.
9	PUMP180 (COMPR)	TYPE = 1 Pressure = 180 psia	Simulates a pump which delivers water to the 100 psia steam boiler.
10	BOIL100 (FLASH2)	Pressure = 100 psia	This block simulates a low pressure (100 psia) steam boiler.
11	SPLIT100 (FSPLIT)	MOLE-FLOW SLXSTM 0.1 RFRAC STM100 1.0	This block splits the steam from BOIL100. The splits are set by FORTRAN block SETSTEAM.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

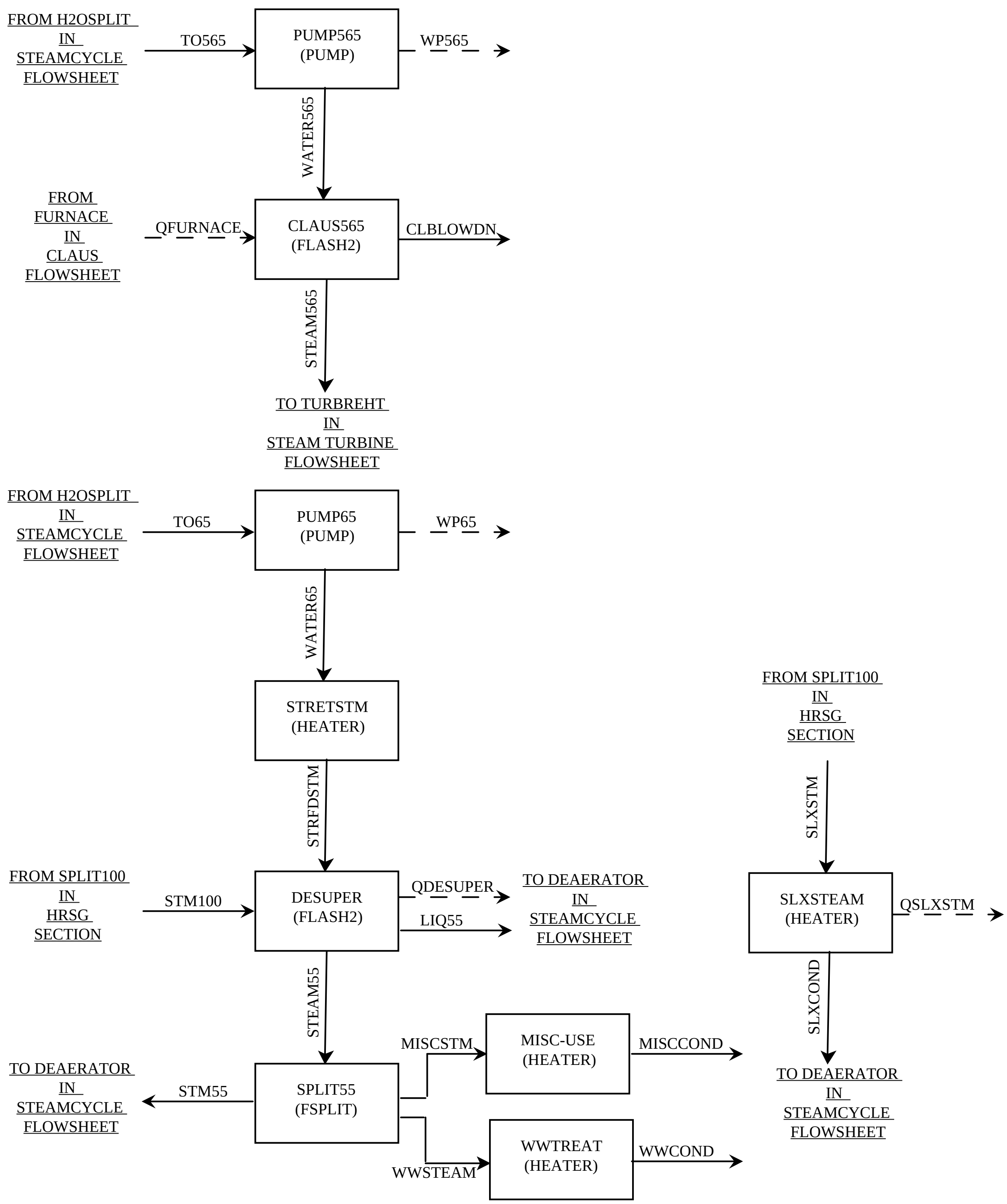

Figure 3.10 Auxiliaries Flowsheet

Table 3.8 Auxiliaries Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	PUMP565 (PUMP)	TYPE = 1 Pressure = 565 psia	This block simulates a pump which delivers water to the Claus plant steam generator.
2	CLAUS565 (FLASH2)	Pressure = 565 psia	This block simulates the Claus plant steam generator.
3	PUMP65 (PUMP)	TYPE = 65 Pressure = 65 psia	This block simulates a pump which delivers water to the BS plant steam generator.
4	STRETSTM (HEATER)	Pressure = 65 psia	This block simulates the BS plant steam generator.
5	SLXSTEAM (HEATER)	Pressure = 115 psia Vfrac = 0	This block simulates the 115 psia steam condensation in the Selexol process.
6	DESUPER (FLASH2)	Pressure = 55 psia Vfrac = 1	Simulates 55 psia steam desuperheater.
7	SPLIT55 (FSPLIT)	MOLE-FLOW WWSTEAM 1.0 MISCSTM 1.0 RFRAC STM55 1.0	This block splits the steam from DESUPER. The splits are set by FORTRAN block SETSTEAM.
8	WWTREAT (HEATER)	Pressure = 55 psia Vfrac = 0	Simulates the condensation of 55 psia steam condensation in Texaco Waste Water Treatment.
9	MISC-USE (HEATER)	Pressure = 55 psia Vfrac = 0	This block simulates the miscellaneous user of 55 psia steam.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

3.2.6.2 Steam Turbine

The details regarding the modeling of the steam turbine are given in Figure 3.11 and Table 3.9. Four steam turbines are modeled in this section: TURB350, TURB115, and TURB70, and TURB1. The steam generated in the HRSG section is expanded through three stages, consisting of a high pressure turbine which takes in steam at a pressure of 1465 psia and has an outlet pressure of 350psia followed by an intermediate pressure turbine with an inlet pressure of 310 psia and outlet pressure of 115 psia, followed by two low pressure turbines in parallel (70 psia and 1 psia outlet pressures).

The superheated steam, stream SHSTEAM, from the HRSG section enters the block TURB350 which simulates a 350 psia exhaust steam turbine. The output stream from this block, STEAM350, is steam at 350 psia. The stream STEAM350 is mixed with STEAM565 from the auxiliaries section in the block TURBHEAT simulating a mixer and is heated by QREHEAT to a temperature of 996 ^{o}F. The resulting stream, modeled by HOTSTEAM at a pressure of 350 psia, enters the block TURB115, which generates steam at 115 psia. This steam at 115 psia is split by the block SPLIT115 into streams TURB70IN and TURB1IN. The ratio of the split is decided by the design specification DEAERHT. The outlet stream modeled by TURB70IN enters the low pressure (70 psia) exhaust turbine, simulated by TURB70. The resulting stream from TURB70 is steam at 70 psia, which enters the DEAERATOR block. The output stream from TURB1, STEAM1, at a pressure of 1 psia enters the block CONDENSR.

3.2.6.3 Design Specifications and FORTRAN blocks

The design specifications used in the steam cycle section of the model are DEAERTHT and STMTEMP.

DEAERHT is used to operate the deaerator approximately adiabatically. The heat stream leaving the block DEAERHT is should be less than 100.0 BTU/hr. This design specification is achieved by varying the ratio of splitting of the stream, SPLIT115. STMTEMP sets the temperature of the stream leaving the HRSG reheat block to be equal to that of the stream leaving superheater. This is achieved by varying the split ratio of the heat stream, QSH-HRSG, which splits into heat streams QSUPER and QREHEAT. QSH-HRSG is obtained by cooling the products of combustion from the gas turbine, GTPOC in the block QSPLIT.

FORTRAN block SETMAKEUP sets the steam cycle makeup water and the FORTRAN block SETSTEAM calculates the various mass flows of water streams such as those represented by TOECON, TOB100, TO565PSI, and TO65PSI. The required water circulation rate to the heat economizers in HRSG is calculated by FORTRAN block SETSTEAM, based on the temperature of the superheated steam, 997 ^{o}F and the temperature at which the water enters the HRSG from the deaerator, 244 ^{o}F. The flow rates of water and steam to other parts of the model is also calculated by the same block.

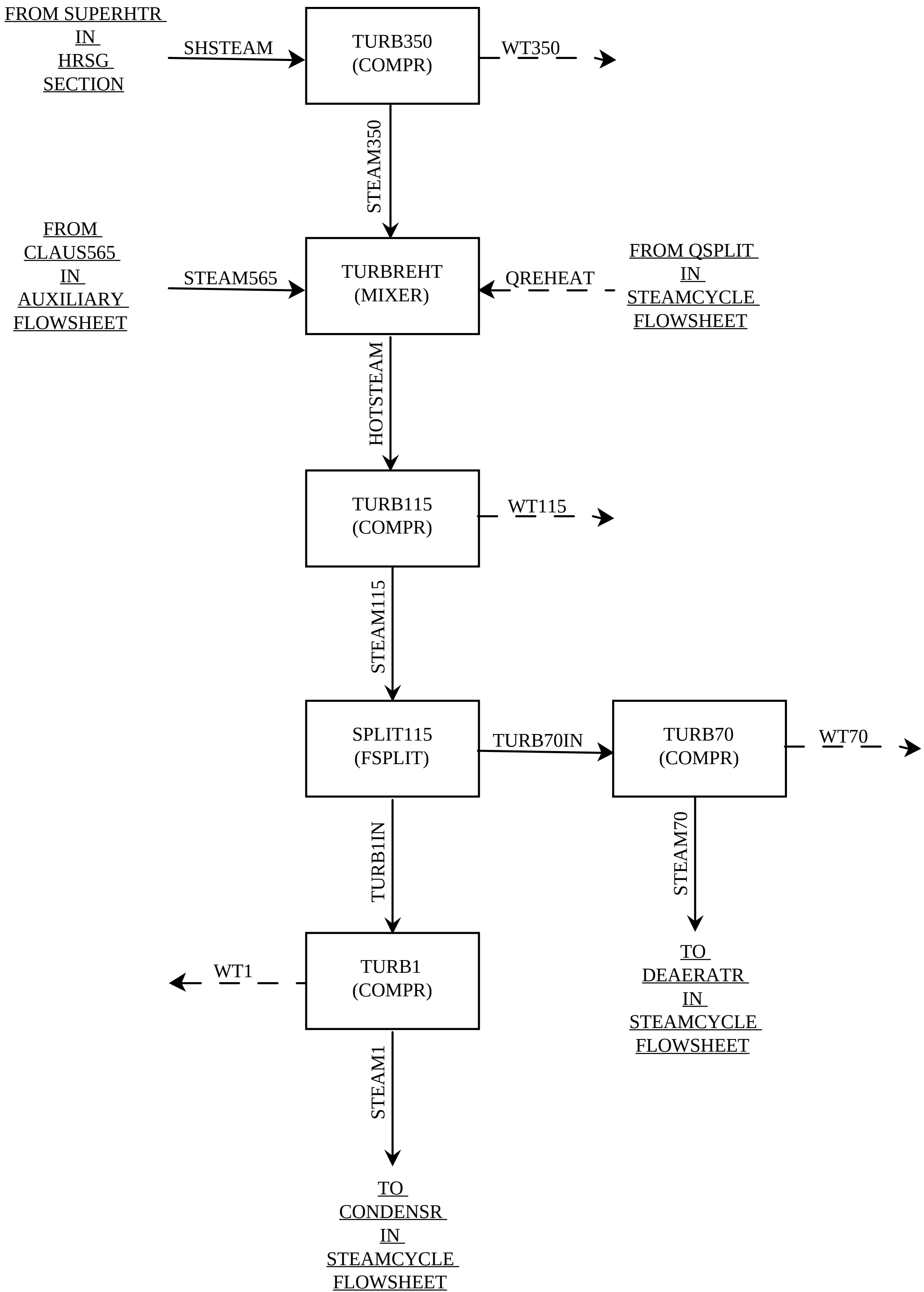

Figure 3.11 Steam Turbine Flowsheet

Table 3.9 Steam Turbine Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	TURB350 (COMPR)	TYPE = 3 Pressure = 350 psia Isoentropic = 0.847	Simulates a high pressure steam turbine.
2	TURBREHT (MIXER)		This block simulates the mixing of steams at 350 psia and 565 psia.
3	TURB115 (COMPR)	TYPE = 3 Pressure = 115 psia Isoentropic = 0.901	Simulates an intermediate pressure steam turbine.
4	SPLIT115 (FSPLIT)	FRAC TURB70IN 0.015 RFRAC TURB1IN 1.0	This block splits the steam from TURB115. The splits are set by design-spec DEAERHT.
5	TURB70 (COMPR)	TYPE = 3 Pressure = 70 psia Isoentropic = 0.85	Simulates a low pressure (70 psia) steam turbine.
6	TURB1 (COMPR)	TYPE = 3 Pressure = 1 psia Isoentropic = 0.849	Simulates a low pressure (1 psia) steam turbine.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

3.2.7 Plant Energy Balance

The plant energy balance is comprised of four energy balance calculations. They are: (1) the gas turbine section net power output estimation; (2) the estimation of the total gross power output of the steam turbine; (3) the estimation of auxiliary power consumption calculated in the ASPEN flowsheet; and (4) the estimates of auxiliary power consumption calculated in the separate cost model subroutine. The last of these calculations is elaborated upon in Chapter 4.0. The remaining three calculations are presented in this section

Assuming a generator loss of 0.5 percent, the net gas turbine power output is calculated to be the sum of the work done by the gas turbine expanders and work required by the gas turbine compressors.

The total gross output of the steam turbine is the sum of the total work done by the four steam turbines.

The auxiliary power consumption is estimated in three different sections of the performance model: (1) the power consumed by the compressors in the Claus plant and Beavon-Stretford plant; (2) the power consumption by all the pumps in the model delivering slurry or water; and (3) the power consumption by the oxygen plant assuming that 1 lbmol/hr of 95 percent purity oxygen requires 6000 watts of power. The auxiliary power consumption models are developed and included in the cost model of the IGCC system. These include models for auxiliary power consumption of coal handling, oxidant feed, gasification, low temperature gas cooling, acid gas removal, Claus and Beavon-Stretford plants, gas turbine, process condensate, boiler feed water, steam cycle, and general facilities sections.

3.3 Convergence Sequence

The convergence sequence for the model simulation is based on nine design specifications and seven FORTRAN blocks. Most of the design specifications and FORTRAN blocks have been described in earlier sections of this chapter and the rest are elaborated upon in this Section.

The FORTRAN block SETFEED maintains the water-to-coal ratio in the model by setting the mass flow of water to the gasifier based upon calculation that the coal slurry has 66.5 percent of solids by weight.

The Oxygen/Coal ratio is varied by a design specification, SETOXYG, in order to achieve the specified syngas exit temperature and overcome a two percent heat loss from the gasifier. The design specifications SETCLAIR and SETBSAIR set the air flow rates to the Claus unit and the Beavon-Stretford tail gas treatment units, respectively. SETCLAIR is designed such that the air provided to the Claus plant is enough to convert one-third of the hydrogen sulfide to elemental sulfur. SETBSAIR provides one percent excess oxygen to completely oxidize the constituents of the tail gas sent to the Beavon-Stretford plant.

The convergence sequence starts with the initialization of key input variables in the FORTRAN block STCTAIL. Then the gasification, high temperature gas cooling, and solid separation process area sequences are called by the master sequence. This is followed by the low temperature gas cooling sequence. The Selexol process and the fuel gas saturation process area sequences are specified next. Then the gas turbine flowsheet sequence is specified followed by the Claus plant and the Beavon-Stretford plant sequences. Then the gas side of the HRSG, and the entire steam cycle sequences are specified. Finally, the FORTAN block which presents user defined results, SUMMARY is attached to the sequence followed by the cost model FORTRAN subroutine, TEXCOST.

3.4 Environmental Emissions

SO_2 emissions from IGCC systems are controlled by removing sulfur species from the syngas prior to combustion in the gas turbine. NO_x emissions tend to be low for this particular IGCC system for two reasons. The first is that there is very little fuel-bound nitrogen in the fuel gas. The second reason is that thermal NO formation is low because of the low syngas heating value and correspondingly relatively low adiabatic flame temperature. A primary purpose of the gas cleanup system is to protect the gas turbine from contaminants in the fuel. Hence, no post-combustion control is assumed. However, it is possible to further control NO_x emissions, for example, through use of Selective Catalytic Reduction (SCR) downstream of the gas turbine. The emission rates of these pollutants are lower than for conventional power plants and for many advanced coal-based power generation alternatives. CO_2 emissions are lower than for conventional coal-fired power plants because of the higher thermal efficiency of the IGCC system (e.g., nearly 40 percent in this case versus typical values of 35 percent for conventional pulverized coal-fired power plants).

3.4.1 NO_x Emissions

The generation of NO and NO_2 from the gas turbine has been modeled in the present study. Both the fuel NO_x as well as thermal NO_x have been taken into consideration for the estimation of NO and NO_2. The default assumptions made for these estimations are that fuel NO is 95 percent by volume of the fuel NO_x, and that the fraction of ammonia that is converted to fuel NO_x is 0.90. The conversion rate of nitrogen to NO_X during the gas turbine combustion is assumed to be 0.00045. Atmospheric emission rates are calculated on a lb/MMBTU basis as part of the model output.

3.4.2 Particulate Matter Estimations

PM emissions are controlled in the syngas cleanup system prior to the gas turbine and therefore, particulate matter emissions from the gas turbine are not modeled in the present model.

3.4.3 CO and CO_2 Emissions

CO emissions from the power plant are assumed to come from the gas turbine section of the plant. The fraction of CO that is converted to CO_2 in the gas turbine is assumed to be 0.99985. Aside from the gas turbine, CO_2 is also emitted by the Beavon-Stretford tail gas treatment unit. The emissions are expressed in terms of lb/kWh.

3.4.4 SO_2 Emissions

SO_2 emissions from the IGCC system are assumed to the result from combustion of syngas in the gas turbine. The SO_2 emissions from the gas turbine are due to oxidation of H_2S and COS in the fuel gas. The amount of H_2S and COS in the fuel gas can be varied by changing the removal efficiency of the Selexol process. The emissions are calculated on a lb/MMBTU basis.

4.0 DOCUMENTATION OF THE AUXILIARY POWER MODEL FOR THE COAL-FUELED TEXACO GASIFIER-BASED IGCC SYSTEM WITH RADIANT AND CONVECTIVE HIGH TEMPERATURE GAS COOLING

Significant amounts of electrical power are consumed by certain process areas of the power plant for the operation of components such as pumps and conveyors. These auxiliary power requirements reduce the net power output of the plant. The auxiliary power requirements are functions of the process variables of the system. Only a few of the auxiliary loads are modeled directly in the ASPEN performance model. They are the total power consumption by the compressors and the centrifugal pumps in the system. All other auxiliary loads are modeled in the cost model subroutine linked to the performance model. These auxiliary power models are described in this chapter.

4.1 Coal Handling

The Texaco IGCC system uses a coal slurry with typically 66.5 weight percent of solids as feed to the gasifier. Coal handling involves coal unloading, stacking, reclamation, and conveying equipment followed by three operating and one spare train of wet grinding equipment. To estimate the auxiliary power requirements of the coal handling unit, a predictive model was developed by Rocha and Frey (1997) using 13 data points obtained from the sources listed in Table 4.1. The coal feed rate was chosen as the independent variable for development of an auxiliary power model. Two models were selected for consideration: power consumed per slurry train vs. coal feed rate per slurry train; and total power consumed by the slurry preparation process area vs. total coal flow to slurry preparation. The power consumed per slurry train vs. coal feed rate per slurry train produced a standard error of 1,183 kW per train and a R^2 of 0.716, whereas the standard error for the other model is 2,949 kW for the entire plant and the R^2 value is 0.807. Because of the higher R^2 value, the latter model was selected.

$$W_{e,\,CH} = 1.04\, m_{cf,\,Ch,i} \tag{4-1}$$

where,

$W_{e,\,CH}$ = Auxiliary power consumption of the coal handling process, kW.

$m_{cf,\,CH,\,i}$ = Coal feed rate, tons/day.

$3{,}300 \leq m_{cf,\,CH,\,i} \leq 20{,}000$ tons per day as-received.

The model and data are shown in Figure 4.1. The model fit is greatly influenced by the data point that is at 20,000 tons/day gasifier coal feed rate (McNamee and White, 1986). A much better fit could occur if this value was removed from the power consumption model consideration. The data point was not removed because no reason could be found to exclude the value from the development of the power consumption model.

Table 4.1 Summary of Design Studies used for Coal Handling and Slurry Preparation Auxiliary Power Model Development

Report No.	Company	Authors	Year	Sponsor[a]	Gasifier	Coal
AP-3109	Synthetic Fuels Associates	Simbeck *et al.*	1983	EPRI	Texaco	Illinois No. 6
AP-3486	Flour Engineers	Matchak *et al.*	1984	EPRI	Texaco	Illinois No. 6
AP-4509	Energy Conversion Systems	McNamee and White	1986	EPRI	Texaco	Illinois No. 6 Texas Lignite
AP-5950	Bechtel Group	Pietruszki-ewicz	1988	EPRI	Texaco	Illinois No. 6
GS-6904	Flour Daniel	Hager and Heaven	1990	EPRI	Dow	Eastern Bituminous
TR-100319	Flour Daniel	Smith and Heaven	1991	EPRI	Destec	Illinois No. 6
MRL Texaco	Montebello Research Lab, Texaco Inc	Robin *et al.*	1991	DOE	Texaco	Pittsburg No. 8

[a]EPRI = Electric Power Research Institute
DOE = U.S. Department of Energy

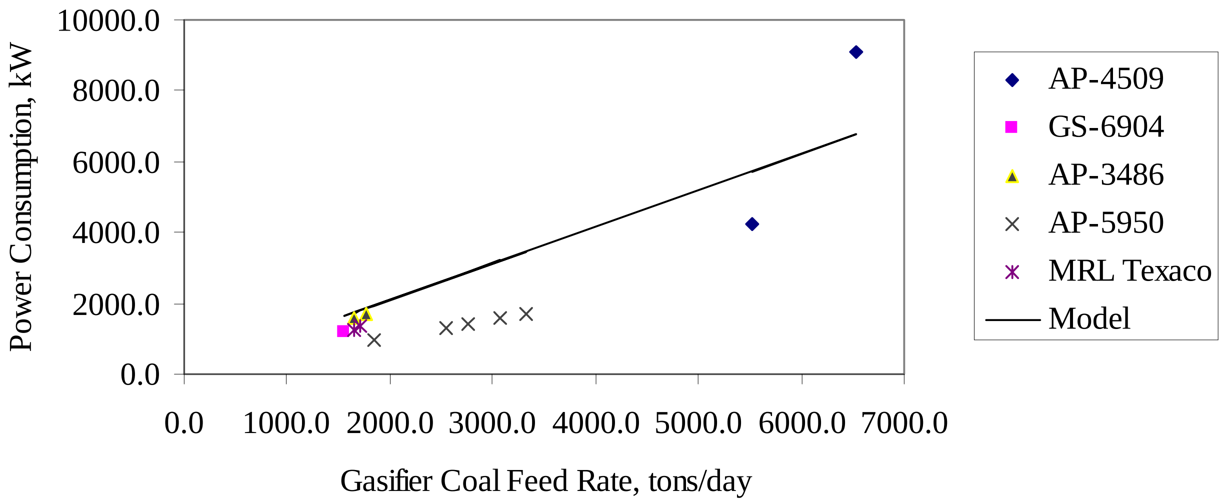

Figure 4.1 Power Requirement for the Coal Slurry Preparation Unit

4.2 Gasification

A single data point is used to estimate the auxiliary power consumption for the gasification process area based upon radiant and convective high temperature gas cooling from a study by Matchak *et al*. (1984). Coal feed rate is used as the independent variable as it is most commonly available for data analysis.

$$W_{e,\,CH} = 0.165\ N_{T,\,G}\ (m_{cf,\,G,\,i}\ /\ N_{o,\,G}) \qquad (4\text{-}2)$$

where,

$W_{e,\,CH}$ = Auxiliary power consumption of the gasification process, kW.

$m_{cf,\,G,\,i}$ = Coal feed rate, tons/day.

The model development of gasification auxiliary consumption by Rocha and Frey (1997) involved six data points, five of which were obtained from Pietruszkiewicz *et al*. (1988) and one from Matchak *et al*. (1984). The former data points formed a straight line and hence the latter data point, which was lower than other data points, was dropped from the model development. However, the five data points indicate that a linear scaling assumption was used by Pietruszkiewicz *et al*. (1988) and the model developed by Rocha and Frey (1997) is overpredictive at the data point excluded from the model. As the design of the IGCC system in this study was based extensively from the findings of Matchak *et al*. (1984), the dropped data point was used instead for the development of the current model.

4.3 Other Process areas

The auxiliary power consumptions of other process areas such as oxidant feed, low temperature gas cooling, Selexol, process condensate treatment, general facilities, pump and compressor power consumption in the Claus, Beavon-Stretford, and steam cycle systems are calculated using regression models developed by Frey and Rubin, 1990. For the convenience of the reader, the models are briefly presented here. For additional details, please refer to Frey and Rubin (1990).

4.3.1 Oxidant feed

The auxiliary power consumption model was developed in the present study modifying a model developed by Frey and Rubin (1990) for oxidant feed power consumption. A single data point with oxygen flow rate as the independent variable in the study by Anand *et al*, (1992) was used modify the original model to reflect the latest published data. The auxiliary power consumption model for oxidant feed section in MW is given by:

$$W_{e,\,OF} = (0.9466 + 3.73 - 4\ T_A + 9.019 \times 10^{-6}\ T_A^{\ 2})\ (0.00526\ M_{O,G,i}) \tag{4-3}$$

where,

$M_{O,G,i}$ = Oxygen gas flow to the gasifier, lb/hr.

T_A = Ambient temperature = 59 ^{o}F.

4.3.2 Low Temperature Gas Cooling

The auxiliary power consumption model for the low temperature gas cooling (LTGC) section was developed as part of the current study using a single data point from the study by Matchak *et al*. (1984) in MW is given by

$$W_{e,\,LT} = 4.3887 \times 10^{-5}\ M_{syn,\,LT,\,o} \tag{4-4}$$

where,

$M_{syn,\,LT,\,o}$ = Molar flowrate of syngas to LTGC section, lbmole/hr.

4.3.3 Selexol

The auxiliary power consumption model for Selexol process in MW was developed by Frey and Rubin (1990) using 18 data points with and R^2 of 0.881 and is given by

$$W_{e,\,S} = 0.348 + 4.78 \times 10^{-4}\, M_{syn,S,o}^{\;0.839} \tag{4-5}$$

where,

$4{,}000 \leq M_{syn,S,o} \leq 74{,}500$ lb/hr

$M_{syn,S,o}$ = Molar flow rate of syngas entering Selexol process, lbmole/hr.

The standard error of the estimate is 550 kW.

4.3.4 Claus Plant

The auxiliary power consumption model for Claus plant in MW was developed by Frey and Rubin (1990) using 20 data points with an R^2 of 0.870 and is given by

$$W_{e,\,C} = 2.1 \times 10^{-5}\, M_{s,C,o} \tag{4-6}$$

where,

$1{,}000 \leq M_{s,C,o} \leq 30{,}800$ lb/hr

$M_{s,C,o}$ = Mass flow of sulfur from Claus plant, lb/hr.

The standard error of the estimate is 67 kW.

4.3.5 Beavon-Stretford Unit

The auxiliary power consumption model for Beavon-Stretford plant in MW was developed by Frey and Rubin (1990) using 6 data points with an R^2 of 1.00 and is given by

$$W_{e,\,BS} = 0.0445 + 0.00112\ M_{s,BS,o} \tag{4-7}$$

where,

$9{,}000 \leq M_{s,BS,o} \leq 18{,}000$ lb/hr

$M_{s,BS,o}$ = Mass flow of sulfur from BS plant, lb/hr.

4.3.6 Process Condensate Treatment

The process condensate treatment plant has the following auxiliary power consumption model, which is developed for the present Texaco IGCC radiant and convective gasification system using a single data point from the study Matchak *et al.* (1984).

$$W_{e,\,PC} = 3.397 \times 10^{-6}\ M_{s,BD} \tag{4-8}$$

where,

$M_{s,BD}$ = Scrubber blowdown flowrate, lb/hr.

4.3.7 Steam Cycle

The boiler feed water (BFW) system supplies the water for steam generation in the HRSG. BFW consists of raw makeup water and the steam turbine condensate. The steam cycle auxiliary power load is due to the BFW treatment section and it is given as the sum of all the work done by the pumps dealing with this section. These pumps are modeled in the ASPEN flowsheet.

$$W_{BFW} = P_{1785} + P_{565} + P_{180} + P_{65} + P_{25} \tag{4-9}$$

where,

W_{BFW} = Auxiliary power consumption by boiler feedwater section, MW

P_{1785} = Work done on the centrifugal pump which delivers BFW at 1785 psia, MW

P_{565} = Work done on the centrifugal pump which delivers BFW at 565 psia, MW

P_{180} = Work done on the centrifugal pump which delivers BFW at 180 psia, MW

P_{65} = Work done on the centrifugal pump which delivers BFW at 65 psia, MW

P_{25} = Work done on the centrifugal pump which delivers BFW at 25 psia, MW

4.3.8 General Facilities

The general facilities include power requirements for cooling water systems, plant and instrument air, fuel system, potable and utility water, nitrogen system, process condensate and effluent water treating. The general facilities auxiliary power load is estimated as a fraction of all other auxiliary loads with a typical value of 10 percent. Based on Frey and Rubin (1990) the auxiliary power load model in MW is given by:

$$W_{e,\,GF} = 0.1\,(W_{e,\,CH} + W_{e,\,CH} + W_{e,\,OF} + W_{e,\,LT} + W_{e,\,S} + W_{e,\,C} + W_{e,\,BS} + W_{e,\,PC} + W_{BFW}) \tag{4-10}$$

The sum of all the above auxiliary power loads gives the total auxiliary power consumption of the power plant, $W_{e,\,AUX}$ in MW.

$$W_{e,\,AUX} = W_{e,\,CH} + W_{e,\,CH} + W_{e,\,OF} + W_{e,\,LT} + W_{e,\,S} + W_{e,\,C} + W_{e,\,BS} + W_{e,\,PC} + W_{BFW} + W_{e,\,GF} \tag{4-11}$$

4.4 Net Power Output and Plant Efficiency

The net plant power output is the total power generated from the gas turbines and steam turbines less the total auxiliary power consumption. The gas and steam turbines have been modeled as a series of compressors and turbines in ASPEN using the unit operation block COMPR. This unit operation block requires outlet pressure and isoentropic efficiencies as parameters. The power consumed by the compressors and the power generated by the turbines are calculated by the ASPEN performance model. The net power output is calculated as part of the cost model which is a part of the FORTRAN subroutine TEXCOST called by the ASPEN input file. The net power output in MW is given by

$$MW_{net} = MW_{GT} + MW_{ST} - W_{e,\,AUX} \tag{4-12}$$

The net plant efficiency on a higher heating value basis is given by

$$\eta = 3.414 \ x \ 10^6 \ \frac{MW_{net}}{M_{cf,CH,i} \ x \ HHV} \qquad (4\text{-}13)$$

where,

η = net plant efficiency.

$M_{cf,\ CH,\ i}$ = Coal feed rate, lb/hr.

HHV = Higher heating value of fuel, BTU/lb.

5.0 CAPITAL, ANNUAL, AND LEVELIZED COST MODELS OF THE COAL-FUELED TEXACO GASIFIER-BASED IGCC SYSTEM WITH RADIANT AND CONVECTIVE HIGH TEMPERATURE GAS COOLING

This chapter documents the cost model developed for the coal-fueled Texaco gasifier-based IGCC plant with radiant and convective high temperature gas cooling. The direct capital costs for all the important process areas of oxidant feed section, coal handling and slurry preparation, gasification section, low temperature gas cooling section, Selexol section, Claus sulfur recovery section, Beavon-Stretford tail gas removal section, boiler feedwater system, process condensate system, gas turbine section, heat recovery steam generator section, steam turbine section, and general facilities are described in that order. The annual, and levelized costs of the model are elaborated upon later in the chapter.

5.1 Direct Capital Cost

New direct cost models for the major process areas in the IGCC system are presented. For the purpose of estimating the direct capital costs of the plant, the IGCC plant is divided into thirteen process areas as listed in Table 5.1. The direct cost of a process section can be adjusted for other years than that year for which they were developed using the appropriate Chemical Engineering Plant Cost Index (PCI) (Chemical Engineering Magazine, 1984-1999) as shown in Table 5.2. For example, if a direct capital cost model, DC_{1989} was developed based on January 1989 dollars, then the direct capital cost in January 1998 dollars, DC_{1998}, is given by:

$$DC_{1998} = DC_{1989}\,\frac{388.0}{351.5} \tag{5-1}$$

Table 5.1 Process Areas for Cost Estimation of an IGCC System.

Area Number	Cost Section
10	Oxidant Feed
20	Coal Handling
30	Gasification
40	Low Temperature Gas Cooling
50	Selexol
60	Claus Plant
70	Beavon-Stretford Plant
80	Boiler Feedwater System
85	Process Condensate Treatment
91	Gas Turbine
92	Heat Recovery Steam Generators
93	Steam Turbine
100	General Facilities

Table 5.2 Plant Cost Index Values

Year	Plant Cost Index
1983	315.5
1984	320.3
1985	324.7
1986	323.5
1987	318.3
1988	336.3
1989	351.5
1990	354.7
1991	360.0
1992	359.5
1993	357.2
1994	361.4
1995	376.1
1996	380.9
1997	383.3
1998	388.0

5.1.1 Oxidant Feed Section

This process section typically has an air compression system, an air separation unit, and an oxygen compression system per train. The minimum number of operating trains is two and there are no spare trains. The number of trains depend on the total mass flow rate of oxygen. A regression model was developed by Frey and Rubin (1990) to estimate the direct capital cost of oxidant feed section. The regression model is applicable to all oxygen-blown gasification systems as the model development involved performance and cost data from 31 oxygen plants taken from 14 studies of oxygen-blown IGCC systems. 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. For further details on the regression model, see Frey and Rubin (1990). The direct cost model for the oxidant feed section is:

$$DG_{OF} = 14.35 \frac{N_{T,OF} \quad T_a^{0.067}}{(1-\eta_{Ox})^{0.073}} \left(\frac{M_{O,G,i}}{N_{O,OF}} \right)^{0.852} \quad (R^2 = 0.936; n = 31) \tag{5-2}$$

where,

$20 \leq T_a \leq 95;\ {}^oF$

$625 \leq \left(\frac{M_{O,G,i}}{N_{O,OF}} \right) \leq 11,350\, lbmole/hr; and$

$0.95 \leq \eta \leq 0.98$

Standard error = $10.8 million January 89 dollars

5.1.2 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. A typical coal handling section contains one operating train and no spare train. A train consists of a bottom dump railroad car unloading hopper, vibrating feeders, conveyors, belt scale, magnetic separator, sampling system, deal coal storage, stacker, reclaimer, as well as some of type of dust suppression system. Two studies (McNamee and White, 1986; Matchak *et al*., 1984) assumed a double boom stacker and bucket wheel reclaimer system. The studies by Smith and Heaven (1992) and Hager and Heaven (1990) assumed a combined stacker reclaimer. Pietruszkiewicz *et al*. (1988) specified conveyors to perform the stacking operation and a rotary plow feeder for the reclaim system.

Slurry preparation trains typically have one to five operating trains with one spare train. The typical train consists of vibrating feeders, conveyors, belt scale, rod mills, storage tanks, and positive displacement pumps to feed the gasifiers. All of the equipment for both the coal handling and the slurry feed are commercially available. This typical train design is assumed in two reports (McNamee and White, 1986; Matchak *et al*., 1984).

A regression model was developed for the direct capital cost of coal handling and slurry preparation using the data collected for possible independent variables affecting direct capital cost. The data sources are shown in Table 4.1 and Figure 5.1 shows the data points. 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 upon the overall flow to the plant rather than on per train basis. This is because a better value of R^2 was obtained in the former case. The regression model derived is:

$$DC_{CH} = 5.466\ M_{cf,G,I} \tag{5-3}$$

where,

$R^2 = 0.882$, n = 16

DC_{CH} = Direct capital cost of gasification section in $ 1000

$3{,}300 \leq M_{cf,G,I} \leq 25{,}000$ tons/day

Standard error = $11.2 million January 89 dollars

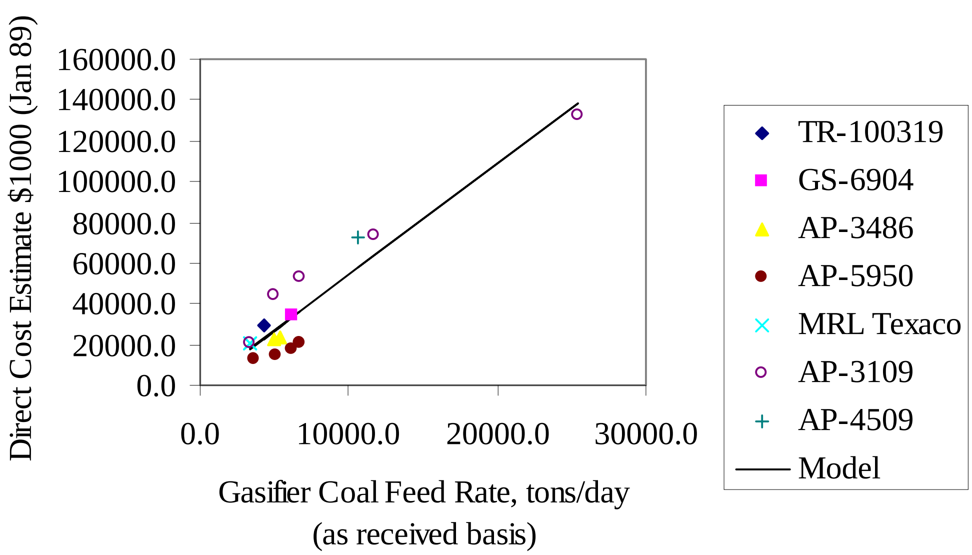

Figure 5.1 Direct Cost for the Coal Handling and Slurry Preparation Process Area.

5.1.3 Gasification Section

The Texaco gasification section of an IGCC plant contains gasifier, gas cooling, slag handling, and ash handling sections. For IGCC plants of 400 MW to 1100 MW, typically four to eight operating gasification trains are used along with one spare train.

The model for the direct capital cost of the gasification section was developed using data collected from various studies sponsored by the Electric Power Research Institute (EPRI) and the U.S. Department of Energy. Of the data collected, the coal flow rate on an as-received basis was the most readily available predictive variable. This was used as the primary predictive variable, since the size and cost of the gasifier is proportional to the coal flow rate. Moisture and ash free coal flow rate, oxidant flow rate, temperature of the gasifier, and the pressure of the gasifier were other possible predictive variables considered. However, as-received coal flowrate was found to be the most useful variable. The direct capital cost model for the gasification section is:

$$DC_G = 216\, N_{T,G} \left(\frac{M_{cf,G,i}}{N_{O,G}} \right)^{0.677} \qquad (R^2 = 0.438; n = 5) \qquad (5\text{-}4)$$

where,

$1{,}200 \le M_{cf,G,I} \le 1{,}600$ tons/day per train (as-received).

Standard error = 1.3 million January 89 dollars

Although the R^2 value for this model is relatively low, the gasifiers are typically of a relatively narrow size range. Hence the low R2 is influenced by the fact that there is a relatively narrow domain of values for the predictive variables in this data set.

5.1.4 Low Temperature Gas Cooling

The low temperature gas cooling (LTGC) section consists primarily of a series of three shell and tube heat exchangers. A cost model was previously developed for this process section for a KRW gasifier with cold gas cleanup by Frey and Rubin (1990), with temperature, pressure, and mass flow of syngas leaving the LTGC section. However, this cost model could not be applied for the present study as the pressure of the syngas leaving the LTGC in the current model is greater than 435 psia, the pressure for which the original model was developed. Since the original cost model development included data points from Texaco studies and since the design basis of this process area is the same in both the KRW gasifier-based and Texaco gasifier-based systems, the cost model was modified to fit a data point of the study by Matchak *et al*., (1984). The syngas mass flow is assumed to be the major determinant of the process area capital cost as in the original cost model. The direct cost model developed is:

$$DC_{LT} = 2.379 \left(\frac{M_{syn,LT,o}}{N_{O,LT}} \right)^{0.79} N_{T,LT} \qquad (5\text{-}5)$$

where,

$$16{,}000 \leq \left(\frac{M_{syn,LT,o}}{N_{O,LT}} \right) \leq 37{,}200 \, lbmole/hr.$$

5.1.5 Selexol Section

Hydrogen sulfide in the syngas is removed through counter-current contact with the Selexol solvent. The cost of the Selexol section includes the acid gas absorber, syngas knock-out drum, syngas heat exchanger, flash drum, lean solvent cooler, mechanical refrigeration unit, lean/rich solvent heat exchanger, solvent regenerator, regenerator air-cooled overhead condenser, acid gas knock-out drum, regenerator reboiler, and pumps and expanders associated with the Selexol process. The cost model is same as the one developed by Frey and Rubin (1990) for a KRW gasifier-based IGCC system with cold gas cleanup. The number of operating trains is calculated based on the syngas mass flow rate and the limits for syngas flow rate per train used to develop the regression model as given below. A minimum of two operating trains and no spare trains are typically assumed. The direct capital cost model for the Selexol section is:

$$DC_S = \frac{0.420 N_{T,S}}{(1-\eta)^{0.059}} \left(\frac{M_{syn,S,i}}{N_{O,S}} \right)^{0.980} \quad (R^2 = 0.909;\ n = 28) \tag{5-6}$$

where,

$$2{,}000 \leq \left(\frac{M_{syn,G,i}}{N_{O,S}} \right) \leq 67{,}300 \, lbmole/hr; and$$

$$0.835 \leq \eta_{HS} \leq 0.997.$$

Standard error = 5.1 million January 89 dollars

This model is valid for H_2S removal efficiencies between 83.5 and 99.7 percent.

5.1.6 Claus Sulfur Recovery Section

The Claus plant contains a two-stage sulfur furnace, sulfur condensers, and catalysts. The cost model is same as the one developed by Frey and Rubin (1990). The number of trains are estimated based on the recovered sulfur mass flow rate and the allowable range of recovered sulfur mass flow rate per train used to develop the regression model. The number of total trains is the number of operating trains and one spare train. Typically, one or two operating trains are used. The direct capital cost model as developed by Frey and Rubin (1990) is:

$$DC_C = 6.28 N_{T,C} \left(\frac{M_{s,C,o}}{N_{O,C}} \right)^{0.668} \quad (R^2 = 0.994;\ n = 21) \tag{5-7}$$

where,

$$695 \le \left(\frac{M_{s,C,o}}{N_{O,C}} \right) \le 18{,}100 \, lbmole/hr$$

Standard error = 235,000 January 89 dollars

5.1.7 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 streams and with the mass flow rate of the sulfur produced. The regression model developed by Frey and Rubin (1990) was based only on the sulfur produced by the Beavon-Stretford process. The number of trains for this area is the same as the number of trains for the Claus plant process area. The direct capital cost model for this process area is:

$$DC_{BS} = 57.5 + 66.2 N_{T,BS} \left(\frac{M_{s,BS,o}}{N_{O,BS}} \right)^{0.645} \quad (R^2 = 0.998; n = 7) \tag{5-8}$$

where,

$75 \leq M_{s,BS,o} \leq 1{,}200$ lb/hr.

Standard error = 260,000 January 89 dollars

5.1.8 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, developed by Frey and Rubin (1990) for a KRW gasifier-based IGCC system, 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. The number of trains used for this commercially available process area is one, with no spare. The direct capital cost model for this process area is:

$$DC_{BFW} = 0.145 M_{rw}^{0.307} M_{pw}^{0.435} \quad (R^2 = 0.991; n = 14) \tag{5-9}$$

where,

$24{,}000 \leq M_{rw} \leq 614{,}000$ lb/hr; and

$234{,}000 \leq M_{pw} \leq 3{,}880{,}000$ lb/hr

5.1.9 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, since the blowdown from the gas scrubbing unit is the larger of the flow streams entering the process condensate treatment section. The regression model developed for this process area's capital cost is:

$$DC_{PC} = 9700 \left(\frac{M_{SBD}}{300000} \right)^{0.6} \tag{5-10}$$

5.1.10 Gas Turbine Section

A number of design factors affect the cost of a gas turbine in an IGCC system. For example, firing of medium-BTU coal gas, as opposed to high-BTU natural gas, requires modification of the fuel nozzles and gas manifold in the gas turbine, which is designed primarily for operating on natural gas. The gas turbine fuel inlet temperature is another important design factor as low fuel inlet temperature may cause liquid condensation leading to corrosion. The cost model for the gas turbine was developed for a GE Frame 7F gas turbine by Frey and Rubin (1990). In the model, the gross gas turbine electrical input is estimated in MW. The number of gas turbines is estimated based on an assumption of 190 MW output for each GE Frame 7F unit. There are no spare gas turbines. The cost model is:

$$DC_{GT} = 32000\ N_{T,GT} \tag{5-11}$$

5.1.11 Heat Recovery Steam Generator

The heat recovery steam generator (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 cost of the HRSG is expected to depend on factors such as the high pressure steam flow rate to the steam turbine, the pressure of the steam, the gas turbine exhaust gas volume flow rate, the number of steam drums, and, to a lesser extent, the boiler feed water or saturated steam flow rates in each of the heat exchangers in the

HRSG. A simple regression model based only on the high pressure steam flow rate to the steam turbine was developed by Frey and Rubin (1990) and is given by:

$$DC_{LT} = -5364 + 7.21 \times 10^{-3} N_{T,HR} P_{hps,HR,o}^{1.526} \left(\frac{M_{hps,HR,o}}{N_{O,HR}} \right)^{0.242} \quad (R^2 = 0.966;\ n = 10) \qquad (5\text{-}12)$$

where,

$650 \le P_{hps,HR,o} \le 1545\ psia;\ and$

$66,000 \le \left(\frac{M_{hps,HR,o}}{N_{O,HR}} \right) \le 640,000\ lbmole/hr.$

Standard error = 6.0 million January 89 dollars

5.1.12 Steam Turbine

A typical steam turbine consists of high-pressure, intermediate-pressure, and low-pressure turbine stages, a generator, and an exhaust steam condenser. The cost of a steam turbine is expected to depend on the mass flow rate of steam through the system, the pressures in each stage, and the generator output, among other factors. The cost model, developed by Frey and Rubin (1990), assumes only one steam turbine and is given by:

$$DC_C = 158.7 W_{ST,E} \quad (R^2 = 0.958;\ n = 9) \qquad (5\text{-}13)$$

where,

$200 \le W_{ST,E} \le 550$ MW.

Standard error = 5.5 million January 89 dollars

5.1.13 General Facilities

The general facilities section includes cooling water systems, plant and instrument air, potable and utility water, and electrical system. Most studies assume that general facilities are approximately 15 to 17 percent of direct costs (Frey and Rubin, 1990; Matchak *et al.*, 1984). In the present study the direct cost of the general facilities is assumed to be approximately 17 percent of the direct costs of the all the other process sections and is given by:

$$GF = f_{GF} \sum_{i=1}^{12} DC_i \tag{5-14}$$

where,

f_{GF} = 0.17.

5.2 Total Plant Costs

The total plant costs of an IGCC power plant include the process facilities capital costs, indirect construction costs, engineering and home office fees, sales tax, allowances for funds used during construction (AFDC), project contingency, and total process contingencies.

The equations for the plant cost model are the same as those given in Frey and Rubin (1990) and are not repeated here. However, the model is briefly described.

Indirect construction costs include worker benefits, supervision and administrative labor, purchased and rented construction equipment, and construction facilities. Engineering and home office fees include the costs associated with engineering, office expenses, and fees or profit to the engineer. Sales tax cost is specific to the state where the power plant is constructed and is estimated as the tax on material costs. AFDC is the estimated debt and equity costs of capital funds necessary to finance the construction of new facilities. Startup costs include one month of fixed operating costs and one month of variable costs based on full plant capacity.

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. Project contingency is used in deterministic cost estimates to represent the expected increase in the capital cost estimate that would result from a more detailed estimate for a specific project at a particular site.

5.3 Total Capital Requirement

The total capital requirement (TCR) includes the total plant investment, prepaid royalties, spare parts inventory, preproduction (or startup) costs, inventory capital, initial chemicals and ctatlyst charges, and land costs. The methodology for calculating TCR is given in detail in Frey and Rubin (1990).

5.4 Annual Costs

The annual costs of an IGCC plant consists of fixed and variable operating costs. The fixed operating costs are annual costs including operating labor, maintenance labor, maintenance materials, and overhead costs associated with administrative and support labor. The variable operating costs include consumables, fuels, slag and ash disposal, and byproduct credits. For more details on the annual cost models, please refer to Frey and Rubin (1990).

5.5 Levelized Costs

The total capital requirement, fixed operating cost, and operating variable cost are used to calculate the cost of producing electricity that is available for sale from the power plant, based on the net electrical output from the power plant. The calculated cost of electricity is also known as total annualized cost and is the levelized annual revenue requirement to cover all of the capital and operating costs for the economic life of the plant.

$$C_{elec} = \frac{[1{,}000\, f_{cr}\, TCR + f_{vclf}\,(FOC + VOC)]\left(\dfrac{1{,}000\, mills}{dollar}\right)}{MW_{net}\, 8{,}760\, c_f} \qquad (5\text{-}15)$$

where,

C_{elec} = The cost of electricity in mills per kWh

TCR = Total capital requirement in $1000

FOC = Fixed operating costs in dollars

VOC = Variable operating costs in dollars

MW_{net} = Net power output in MW

f_{cr} = Fixed charge factor = 0.1034

f_{vclf} = Variable levelization cost factor = 1.0

C_f = Capacity Factor = 0.65

6.0 APPLICATION OF THE PERFORMANCE, EMISSIONS, AND COST MODEL OF THE COAL-FUELED IGCC SYSTEM WITH RADIANT AND CONVECTIVE GAS COOLING TO A DETERMINISTIC CASE STUDY

An example case study is presented here to illustrate the use of the new IGCC system model. The key steps in running the ASPEN simulation model of the Texaco gasifier-based IGCC system are: (1) specify input assumptions; (2) execute the model; (3) collect results; and (4) interpret the results.

6.1 Input Assumptions

Model input assumptions were developed for the performance and cost model based upon a review of design and performance parameters obtained from literature (Frey and Rubin, 1990; Frey and Rubin, 1991; Matchak *et al.*, 1984; Farmer, 1997; Holt, 1998). The assumed composition of the 3.9 weight percent (dry basis) sulfur Illinois No. 6 coal is given in Table 3.1. The model is configured to represent three parallel trains of heavy duty "Frame 7F" gas turbines.

Table 6.1 summarizes a number of the input assumptions for the example case study, with a focus on the key inputs for the gasifier and gas turbine process areas of the model. Many of these assumptions have been previously described in the technical description of the technology. Two of the assumptions listed in the table are initial values that may be modified during the simulation. These are the Oxygen/Coal ratio in the gasifier and the Turbine Inlet Temperature in the gas turbine. The Oxygen/Coal ratio is varied by a design specification in order to achieve the specified syngas exit temperature and overcome a two percent heat loss from the gasifier. The Turbine Inlet Temperature may be lowered from the initial value of 2,350 ^{o}F in order to maintain the exhaust gas temperature below 1,120 ^{o}F. There are literally hundreds of other input assumptions to the model. Only the most significant ones affecting plant design and operation are shown here. The cost model assumptions used in this case study are similar to those reported by Frey and Rubin (1991).

6.2 Model Results

The version of ASPEN used in the present study is the one developed by US Department of Energy. To execute the ASPEN model, an input file is prepared using standard ASPEN keywords and is submitted to a multi-step process leading to model execution. In the first step, the input file is translated into a FORTRAN program, which is then compiled and linked to the extensive library of ASPEN unit operation and other subroutines. The model is then executed and produces numerous output files. This particular case study was executed on a VAX 4000 located at Carnegie Mellon University, and the clock time for the run was approximately 5 minutes.

Selected performance and cost results from the model output are summarized in Tables 6.2 and 6.3. The overall energy balance is indicated in Table 6.2. The plant is estimated to produce a net of 863 MW with an overall plant efficiency of 39.4 percent on a higher heating value basis. The breakdown of plant power production and internal plant power consumption for auxiliaries is given in the Table 6.2. Buchanan *et al*. (1998) mentions a first-of-a-kind (FOAK) IGCC plant which has a gasifier that closely resembles the radiant and convective design adopted in the present study. The efficiency of the FOAK plant is given to be 40.1 percent which is comparable to the efficiency obtained by the current model. The FOAK plant produces 543 MW on a higher heating value basis.

Estimated emission rates for SO_2, NO_x, particulate matter (PM), and CO_2, are provided in Table 6.2. SO_2 emissions from IGCC systems are controlled by removing sulfur species from the syngas prior to combustion in the gas turbine. NO_x emissions tend to be low for this particular IGCC system because there is very little fuel-bound nitrogen in the fuel gas and thermal NO formation is low due to the low syngas heating value and correspondingly relatively low adiabatic flame temperature. PM emissions are controlled in the syngas cleanup system prior to the gas turbine. A primary purpose of the gas cleanup system is to protect the gas turbine from contaminants in the fuel. Hence, no post-combustion control is assumed. However, it is possible to further control NO_x emissions, for example, through use of Selective Catalytic Reduction (SCR) downstream of the gas turbine. The emission rates of these pollutants are lower than for conventional power plants and for many advanced coal-based power generation alternatives. CO_2 emissions are lower than for conventional coal-fired power plants because of the higher thermal efficiency of the IGCC system (e.g., nearly 40 percent in this case versus typical values of 35 percent for conventional pulverized coal-fired power plants).

The estimated costs for the IGCC system given in Table 6.3 include capital, annual, and levelized costs. These costs are inclusive of the entire power plant, including the environmental control system. The breakdown of total capital cost of $1,732/kW includes a 47.1 percent contribution from direct costs, a 5.4 percent contribution from process contingencies, a 12.2 percent contribution from project contingencies, and a 13.1 percent contribution from allowances for funds used during construction. The remaining contributions are from other indirect costs and startup costs. The largest annual cost is for fuel consumption. The byproduct credit for sale of elemental sulfur offsets the incremental variable costs for all consumables other than fuel. The levelized cost of electricity, based upon a 65 percent capacity factor, is 50.9 mills/kWh (5.09 cents/kWh). This cost of electricity is comparable to that of many other coal-based power generation systems evaluated using similar financial assumptions.

Table 6.1 Summary of Selected Base Case Input Values for the Texaco Gasifier-Based IGCC System with Radiant and Convective High Temperature Gas Cooling

Description	Value
Gasifier Process Area	
Gasifier Pressure, psia	615
Gasifier Outlet Temperature, ^{o}F	2,400
Oxygen/Coal Ratio, lb O_2 / lb Coal (Initial Value)	0.915
Slurry Water/Coal Ratio, lb H_2O / lb Coal	0.504
Radiant Cooler Outlet Temperature, ^{o}F	1,500
Convective Cooler Outlet Temperature, ^{o}F	650
Radiant Cooler Heat Loss, %	6
Gas Turbine Process Area	
Inlet Syngas Temperature, ^{o}F	570
Fuel Moisturization, wt-% of Clean Gas	28.2
Pressure Ratio	15.5
Turbine Inlet Temperature, ^{o}F (Initial Value)	2,350
Compressor Isentropic Efficiency, %	81.0
Expander Isentropic Efficiency, %	91.9
Generator Efficiency, %	98.5

Table 6.2 Summary of Selected Performance Model of the Coal-Fuel System with Radiant and Convective High Temperature Gas Cooling Point Estimate Results from the Example Case Study

Description, Units	Value
Gas Turbine Output, MW	579.5
Steam Turbine Output, MW	400.8
Auxiliary Power Demand	
Coal Handling, MW	7.3
Oxidant Feed, MW	83.5
Gasification, MW	1.2
Low T. Cool., MW	2.4
Selexol, MW	4.8
Claus, MW	0.4
Beavon-Stretford, MW	1.3
Steam Cycle, MW	5.3
Process Condensate, MW	0.6
General Facilities, MW	10.7
Total Auxiliary Load, MW	117.4
Net Power Output, MW	862.9
Heat Rate, BTU/kWh (HHV basis)	8,664
Efficiency, % (HHV basis)	39.4
SO_2 Emissions, lb/10^6 BTU	0.22
NO_x Emissions, lb/10^6 BTU	0.13
Particulate Matter (PM) Emissions, lb/10^6 BTU	< 0.03
CO_2 Emissions, lb/kWh	1.70

Table 6.3 Summary of Cost Model Results for the Example Case Study (1998 Dollars)

Description, Units	Value
Capital Cost Summary ($/kW)	
Total Direct Cost	815
Total Indirect Costs	299
Process Contingencies	94
Project Contingency	211
Total Plant Cost	1,419
AFDC (see note below)	227
Total Plant Investment	1,647
Startup Costs and Land	43
Total Capital Requirement[a]	1,732
Fixed Operating Cost, $/(kW-yr)	50.4
Incremental Variable Costs, mills/kWh	1.2
Byproduct Credit, mills/kWh	-1.5
Fuel Cost, mills/ kWh	10.9
Variable Operating Cost, mills/kWh	10.6
Cost of Electricity, mills/kWh	50.9

Note: AFDC = Allowances for Funds used During Construction
Fuel Cost, $/MMBTU = 1.26 (Jan 1998 Dollars) (Buchanan *et al*., 1998)
Capital Recovery Factor = 0.1034
a = Total Capital Requirement inlcudes Total Plant Investments, Startup costs and Land, Inventory Capital, Initial Catalysts and Chemicals

7.0 DOCUMENTATION OF THE PLANT PERFORMANCE, EMISSIONS, AND COST SIMULATION MODEL IN ASPEN OF THE COAL-FUELED TEXACO-GASIFIER BASED IGCC SYSTEM WITH TOTAL QUENCH HIGH TEMPERATURE GAS COOLING

The performance model of an oxygen-blown Texaco gasifier based IGCC system with total quench high temperature gas cooling (referred to here as the "total quench model") is documented in this chapter. The development of the model is based primarily on the findings of a study conducted by Electric Power Research Institute (Matchak *et al*., 1984). This design is adopted as it provides extensive information on the mass flows of streams, temperatures, pressures, power production and consumption, and costs associated with each process section of the plant. Most of the major process sections are modeled in similar method as for the radiant and convective method. Tables and figures are listed for those process areas which are modeled differently from those in the radiant and convective Texaco gasifier-based IGCC system. The convergence sequence for the present model is described along with the FORTRAN blocks and design specifications used in the model.

7.1 Major Process Sections in the Total Quench IGCC Process Simulation Model

Most of the major flowsheet sections in the process simulation model of the total quench-based system, such as coal slurry preparation, gasification, particulate scrubbing, acid gas removal, Claus sulfur recovery, Beavon-Stretford tail gas treatment, and gas turbine, are similar in design to those in the radiant and convective-based model. The flowsheet sections in the total quench model that are significantly different from their counterparts in the radiant and convective design which are the high temperature gas cooling section, low temperature gas cooling section, fuel gas saturation, and steam cycle, are described below. The other process are modeled in the same manner as descibed in Chapter 3.

7.1.1 Gasification and High Temperature Gas Cooling

Figure 7.1, and Table 7.1 illustrate the structure and input assumptions of the gasification and high temperature gas cooling models. The gasification process is similar to that in the radiant and convective design. The crude gas leaving the gasificaton unit is at a temperature of 2400 ^{o}F to 2600 ^{o}F. As shown in Figure 7.1, the hot gas is introduced directly into a water quench chamber located below the gasifier vessel. In the model, the hot gas is simulated by the stream RXROUT. RXROUT enters the unit operation block QUENMIX, which simulates a mixer. Quench water and the hot gas from the block GASIFIER are mixed in QUENMIX. The resulting output stream, modeled by QUENGAS, flows to the unit operation block QUENHEAT which simulates a heater. QUENHEAT cools the QUENGAS stream to a temperature of 433 ^{o}F.

A design specification, SETQUEN, is used for setting the temperature of the output stream from the QUENHEAT block. The mass flow of the quench water, represented by QUWATER, is varied until the temperature of the stream represented by COOLGAS is 433 ^{o}F. The quenched gas is sent to the particulate removal section.

The other components of this process area, such as the blocks SLURPUMP, COALCONV, MAKESOOT, MAKESLAG, and SLAGOUT, are the same as described in Chapter 3.0.

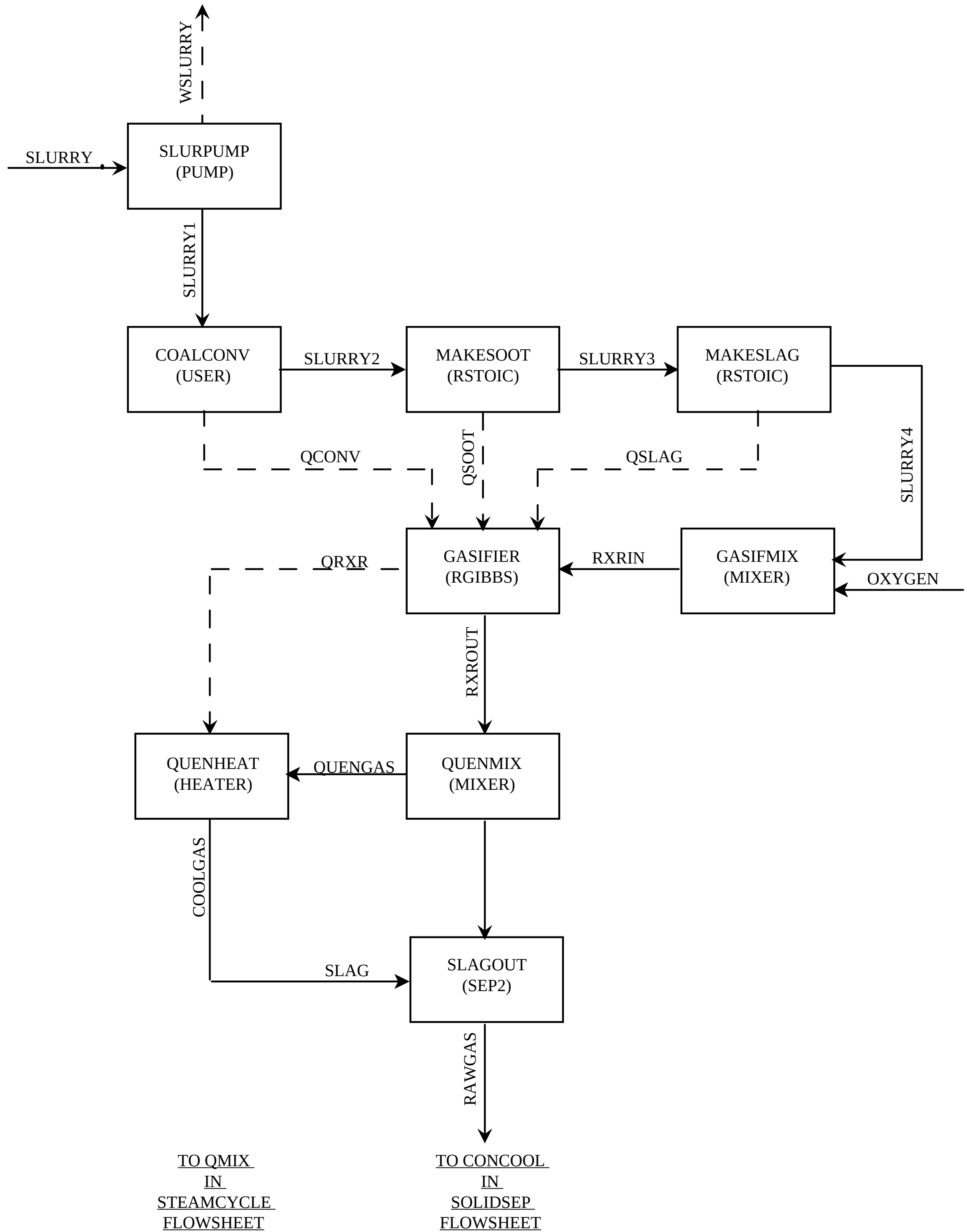

Figure 7.1 Gasification Flowsheet

Table 7.1 Gasification Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	SLURPUMP (PUMP)	TYPE=2 Pressure = 650 psia Efficiency = 0.65	This block simulates Coal-Water Slurry Pump which delivers slurry to the gasifier burners.
2	COALCONV (USER)		This block decomposes coal into its elements using the subroutine USRDEC
3	MAKESOOT (RSTOIC)	Temperature = 59 ^oF Pressure drop = 0 psia	Simulates the stoichiometric reaction which produces soot based on the coal's ultimate analysis.
4	MAKESLAG (RSTOIC)	Temperature = 59 ^oF Pressure drop = 0 psia	Simulates the stoichiometric reaction which produces slag based on the coal's ultimate analysis.
5	GASIFMIX (MIXER)		Represents a Mixer which mixes the coal slurry and the oxidant feed.
6	GASIFIER (RGIBBS)	Temperature = 2400 ^oF Pressure = 615 psia NAT = 6 NPHS = 1 NPX = 2 NR = 9 IDELT = 1	This block simulates the stoichiometric reactions associated with the Gasifier Reactor.
7	QUENMIX (MIXER)		Simulates a MIXER which mixes quench water with the raw gas from the gasifier. The amount of water is decided by the design-spec SETQUEN.
8	QUENHEAT (HEATER)	Pressure = 572 psia	This block heats the quenched gas so that it achieves a temperature of 433 ^oF

(continued on next page)

Table 7.1. Continued

9	SLAGOUT (SEP2)	COMP FRAC COAL = 1.0 ASH = 1.0 SLAG = 1.0 SOOT = 0.0	This block places slag into the Gasifier bottoms stream.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

7.1.2 Low-Temperature Gas Cooling and Fuel Gas Saturation

Figure 7.2 and Table 7.2 illustrate the structure of the low temperature gas cooling section of the total quench model. The scrubbed gas from the solids separation section, represented by TOBSAT100, is cooled by heat exchange with the circulating saturator water. The scrubbed gas is first cooled by heat exchanger BSAT100. The heat recovered here is used to generate 100 psia steam in the HRSG section. Blocks COOLA, COOL1, and BSAT55 are other heat exchangers which cool the raw gas to 332 ^{o}F. The gas is further cooled to 130 ^{o}F by heating vacuum condensate and makeup water from block DAERATOR in the steam cycle. The raw gas at 130 ^{o}F is cooled to 101 ^{o}F in the trim cooler. The condensate from all the above mentioned heat exchangers is collected in the condensate collection drum, CONDMIX. The cooled gas is sent to the Selexol acid gas removal unit.

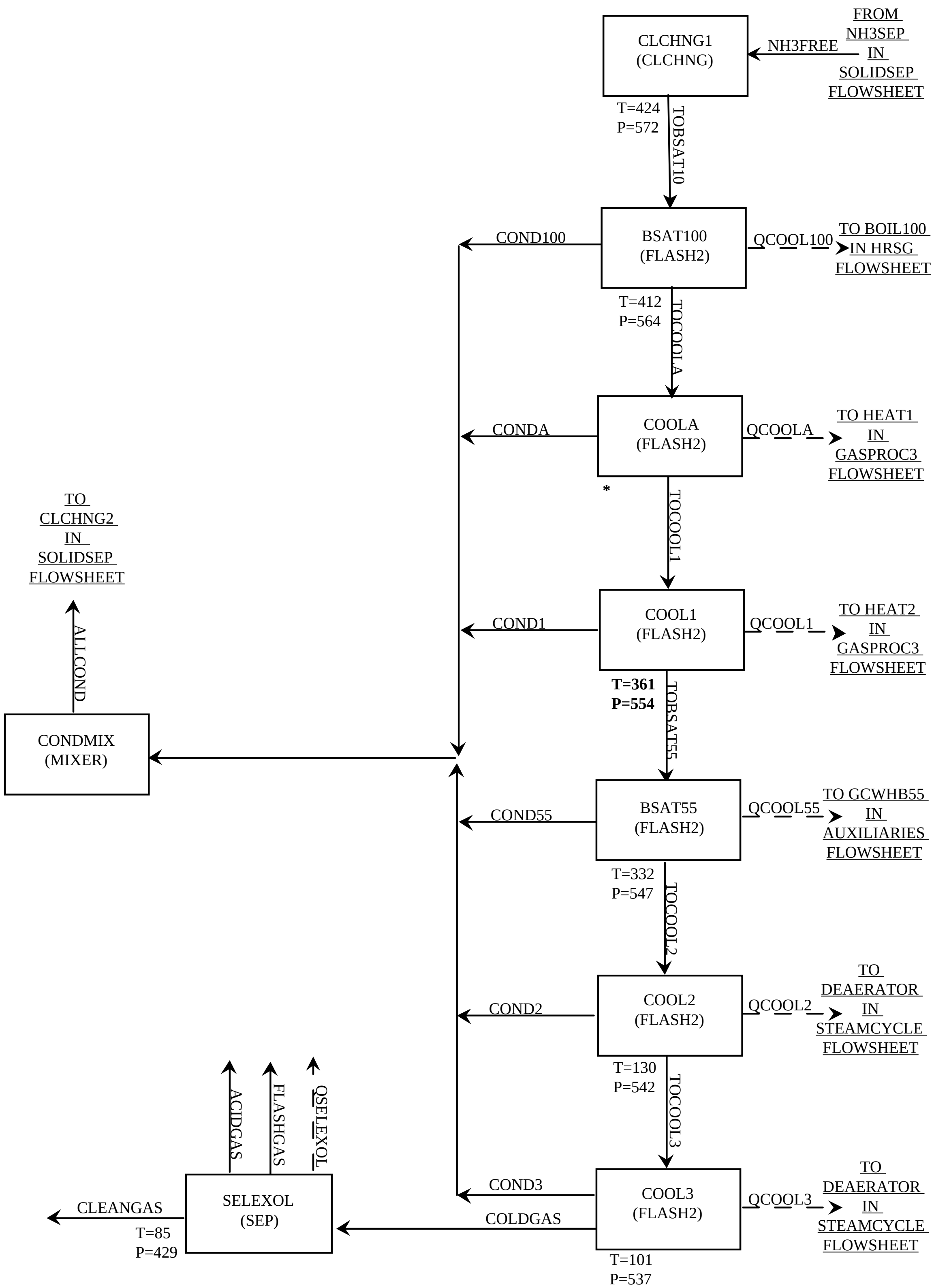

Figure 7.2 Low Temperature Gas Cooling Flowsheet

Table 7.2 Low Temperature Gas Cooling Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	CLCHNG1 (CLCHNG)		This block changes stream class from MIXCINC to Conventional.
2	BSAT100 (FLASH2)	Temperature = 412 °F Pressure drop = 8 psia	This block simulates a heat exchanger which reduces the temperature of the syngas to 412 °F from 424 °F across a pressure drop of 8 psia.
3	COOLA (FLASH2)	Temperature = 396.1 °F Pressure drop = 5 psia	This block simulates a heat exchanger which reduces the temperature of the syngas to 396.1 °F from 412 °F across a pressure drop of 5 psia.
3	COOL1 (FLASH2)	Temperature = 361 °F Pressure drop = 5 psia	This block simulates a heat exchanger which reduces the temperature of the syngas to 361 °F from 391.1 °F across a pressure drop of 5 psia.
4	BSAT55 (FLASH2)	Temperature = 332 °F Pressure drop = 7 psia	This block simulates a heat exchanger which reduces the temperature of the syngas to 361 °F from 323 °F across a pressure drop of 7 psia.
5	COOL2 (FLASH2)	Temperature = 130 °F Pressure drop = 5 psia	This block simulates a heat exchanger which reduces the temperature of the syngas to 130 °F from 332 °F across a pressure drop of 5 psia.
6	COOL3 (FLASH2)	Temperature = 101 °F Pressure drop = 5 psia	This block simulates a heat exchanger which reduces the temperature of the syngas to 101 °F from 130 °F across a pressure drop of 5 psia.
7	CONDMIX (MIXER)		This block simulates the mixing of all condensates in this section.

(continued on next page)

Table 7.2. Continued

8	SELEXOL (SEP)	CLEANGAS T = 85 °F, P = 429 psia ACID GAS T = 120 °F, P=22 psia FLASH GAS T = 58 °F, P = 115 psia	This block separates the syngas into Acid Gas, Flash Gas, and Clean Gas.
9	RMHEAT (HEATER)	Temperature = 421 °F Pressure = 500 psia	Simulates the cooling of the high pressure boiler feed water from the HRSG.
9	HEAT1 (HEATER)	Pressure = 500 psia	This block splits the HOTH2O required for saturation of fuel gas to 28.2 wt % moisture. The split is set by the FORTRAN block SATURH2O.
10	HEAT2 (HEATER)	Pressure = 500 psia	Simulates the cooling of the hot BFW.
9	WARMCOOL (HEATER)	Temperature = 350 °F Pressure = 429 psia	Simulates the mixing of the CLEANGAS and SATCOM.
10	HOTSPLIT (FSPLIT)	MOLE-FLOW SATCOM1 1.0 RFRAC WARMH2O 1.0	Simulates the heating of the saturated gas such that the fuel gas temperature before entering REHEAT is 347 °F.
11	PUMP1K1 (PUMP)	TYPE = 1 Pressure = 500 psia	Simulates a pump which delivers water at 500 psia.
12	WMIX (MIXER)		Simulates a mixer
13	COOLSPLT (FSPLIT)	MOLE-FLOW TOHEAT12 1.0 RFRAC COOLH2O 1.0	This block splits a given stream into two streams. The split is calculated in FORTRAN block SETINIT
14	COLDCOOL (HEATER)	Temperature = 252 °F Pressure = 429 psia	Simulates the cooling of water to 252 °F and 429 psia
15	COLDSPLT (FSPLIT)	MOLE-FLOW COLDH2O 1.0 RFRAC SATCOM2 1.0	This block splits a given stream into two streams. The split is calculated in FORTRAN block SETINIT
16	PUMP1K2 (PUMP)	TYPE = 1 Pressure = 500 psia	Simulates a pump which delivers water at 500 psia.
17	HMIX (MIXER)		Simulates a mixer

(continued on next page)

Table 7.2. Continued

18	SATMIX (MIXER)		This block mixes cleangas from Selexol, with water so that the moisture content of clean gas is 40.0% by weight.
19	SATHEAT (HEATER)	Pressure = 419 psia	This block heats the mixture of clean gas and water so that the mixture is saturated.
20	REHEAT (HEATER)	Pressure = 414 psia	Simulates a Fuel Gas Reheater – Cold Side.
21	SH-HRSG (HEATER)	Temperature = 856 ^{o}F Pressure drop = 0 psi	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
22	HP-HRSG (HEATER)	Temperature = 639 ^{o}F Pressure drop = 0 psi	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
23	E3-HRSG (HEATER)	Temperature = 541 ^{o}F Pressure drop = 0 psi	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
24	IP-HRSG (HEATER)	Temperature = 469 ^{o}F Pressure drop = 0 psi	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
25	E2-HRSG (HEATER)	Temperature = 420 ^{o}F Pressure drop = 0 psi	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.
26	LP-HRSG (HEATER)	Temperature = 365 ^{o}F Pressure drop = 0 psi	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.

(continued on next page)

Table 7.2. Continued

27	E1-HRSG (HEATER)	Temperature = 307 ^{o}F Pressure drop = 0 psi	This block is part of the Heat Recovery Steam Generation Section and removes heat from the products of combustion of the Gas Turbine.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

Figure 7.3 shows the details of the fuel gas saturation unit. The syngas leaving the Selexol acid gas recovery unit, CLEANGAS is saturated with moisture before the gas enters the gas turbine combustor. This is done with the intent of raising the net plant power output and to control NO_x emissions from the gas turbine, as previously described in Section 2.5. The steam saturation increases the mass throughput and the heat capacity of the inlet pressurized fuel gas stream to the gas turbine resulting in an increase in the gas turbine power output.

The clean gas, modeled as stream CLEANGAS, from the Selexol process, enters the saturation unit at 85 ^{o}F and 429 psia. The saturation unit is provided with two stages in order to achieve a high moisture content of 40 weight percent in the fuel gas. Large quantities of heat are required to achieve the high moisture content in the fuel gas because large amount of cold water from the saturator and boiler feed water from the HRSG have to be heated. This required heat is supplied from the raw gas during low temperature gas cooling. The saturated gas is also reheated in this unit to 520 ^{o}F using high pressure boiler feedwater from HRSG.

The model of saturator for the total quench system is different from that described in Section 2.5. Instead of direct contact of syngas with water, the heat transfer between the clean syngas and the saturator water are modeled. The amount of water required to saturate the clean syngas to 40 weight percent moisture is calcuated, and the heat required to vaporize this amount of water is obtained from blocks HEAT1 and HEAT2. Finally, the water vapor is mixed with the clean syngas and reheated in REHEATR before the syngas is sent to the gas turbine. HEAT1 and HEAT2 simulate blocks which heat the circulating water using heat recovered from unit operartion blocks COOLA and COOL1 respectively. WARMCOOL cools the hot water entering the saturator unit to an intermediate temperature. The cooled hot water, HOTH21, is split into two streams, WARMH2O and SATCOM1, by HOTSPLIT. PUMP1K1 is a pump, which increases the pressure of WARMH2O to 500 psia. A mixer WMIX mixes the 500 psia WARMH2O and the heated water from HEAT2. The mixed stream, TOSPLT, is split by COOLSPLT into two streams, COOLH2O and TOHEAT12. The stream COOLH2O is cooled to a temperature of 252 ^{o}F in block COLDCOOL, and split into COLDH2O and SATCOM2. PUMP1K2 increases the pressure of COLDH2O to 500 psia before it is sent to HEAT2 block. TOHEAT12 is sent to the block HEAT1, where it is heated to become HOTW2. CLEANGAS, SATCOM1, and SATCOM2 are mixed and heated to a temperature of 370 ^{o}F in the block SATHEAT. The high pressure boiler feedwater from the HRSG, FGSMAK, is cooled to 421 ^{o}F by RMHEAT. The heat recovered from FGSMAK is used to heat the saturated clean gas to 520 ^{o}F. The reheated fuel gas, GTFUEL flows to the gas turbine combustors.

The saturation section flowsheet contains a FORTRAN block SETINIT which calculates the required amount of water to be added to clean gas to make its moisture content 40.0 percent by weight. SETINIT obtains the mass flow of clean gas entering the saturator block and calculates mass flow of the saturated gas, SATGAS. The equation used for this purpose is given by,

$$M_{water} = \frac{\eta}{100-\eta} M_{Clean\,Gas} \tag{7-1}$$

where,

M_{water} = Mass flow of water to be added to the clean syngas, lb/hr

= SATCOM1 + SATCOM2

η = weight percent of moisture to be present in the saturated syngas

M_{CG} = Mass flow of clean syngas from acid gas removal section (dry basis), lb/hr

The FORTRAN block SETINIT also calculates the split ratios for the blocks HOTSPLIT and COLDSPLT using similar methods as in the FORTRAN block SETSTEAM, elaborated upon in Section 3.2.6.3. SETINIT also sets the mass flow makeup water to the steam cycle equal to the mass flow of water added to the clean syngas.

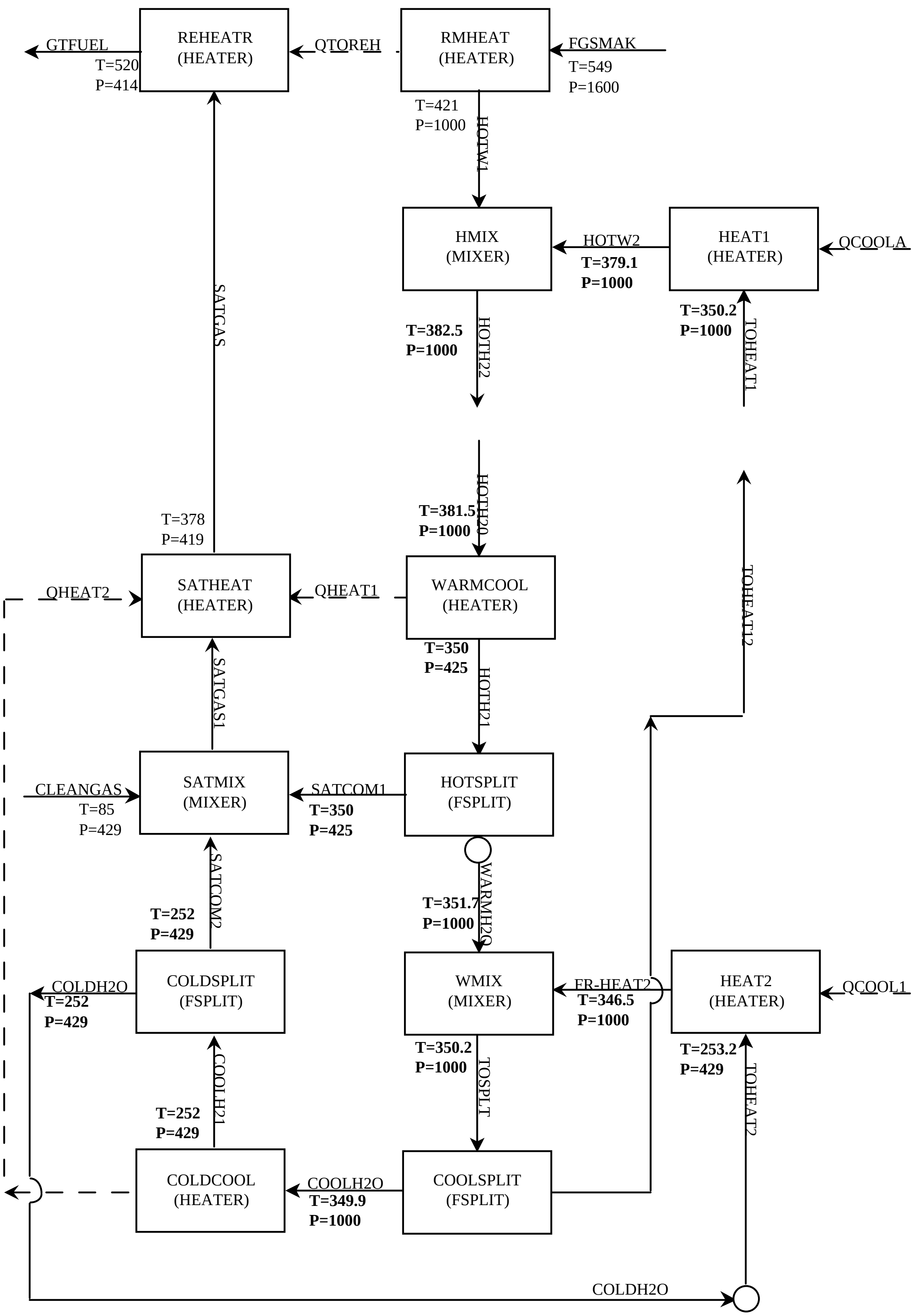

Figure 7.3 Fuel Gas Saturation Flowsheet

7.1.3 Steam Cycle

The steam cycle designed for the total quench model is similar to the one designed for radiant and convective IGCC system except for a few differences. The HRSG section in the total quench model has two extra economizers and an intermediate pressure evaporator. The auxiliaries section has an additional 55 psia centrifugal pump and a 55 psia steam boiler. The steam turbine section has only one low pressure (1 psia) steam turbine unlike in the case of radiant and convective model in which there is also a 70 psia low pressure steam turbine. The rest of the steam cycle is the same as that described in Section 3.2.6.

7.1.3.1 Heat Recovery Steam Generation (HRSG)

The operations of the HRSG are to preheat boiler feed water, reheat intermediate pressure steam, supplement high pressure and 100 psia steam generation, and to superheat high pressure steam. Figure 7.4 and Table 7.3 illustrate the model of the HRSG section. The HRSG is arranged in the following order:

1. Superheater and reheater in parallel,
2. High pressure evaporator,
3. Economizer,
4. Intermediate pressure evaporator,
5. Economizer,
6. 100 psia boiler, and

7. Economizer.

Most of the HRSG section design is similar to the HRSG design in the radiant and convective model. The key additions are ECONOMZ3, which models two economizers, and the intermediate pressure boiler, IPBOILER, which generates saturated steam of 350 psia. This steam is combined in the high pressure power turbine, TURBREHT, with the high pressure steam (565 psia), STEAM565, from the Claus plant.

7.1.3.2 Auxiliaries Section

The auxiliaries section has similar design to that in the radiant and convective model as shown in Figure 7.5 and Table 7.4. The key difference is the additional generation of 55 psia steam by a waste heat boiler, GCWHB55 which is sent to the block SPLIT55.

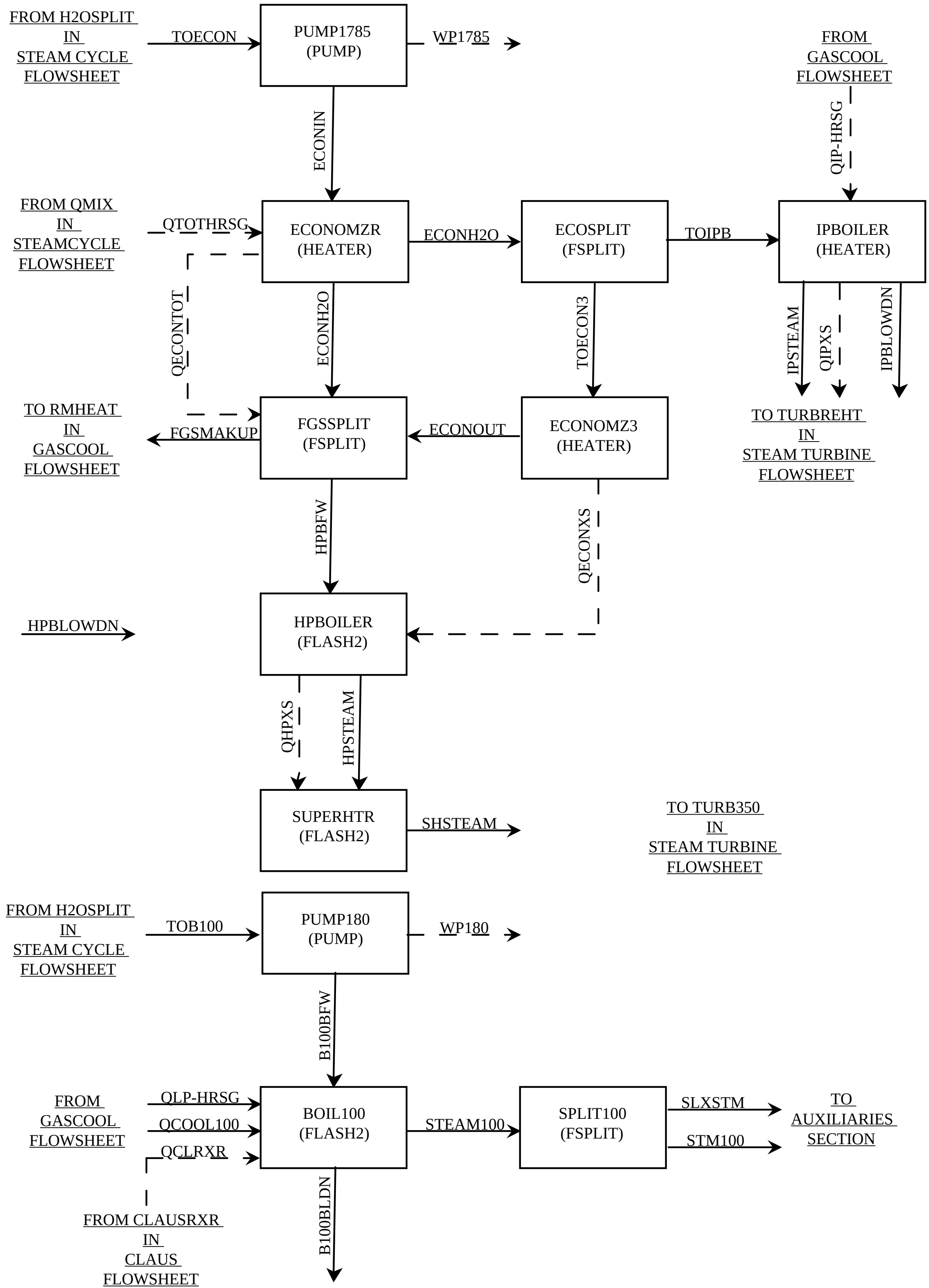

Figure 7.4 HRSG Section Flowsheet

Table 7.3 HRSG Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	PUMP1785 (COMPR)	TYPE = 1 Pressure = 1785 psia	Simulates a pump which delivers condensate to the HRSG economizer.
2	ECONOMZR (HEATER)	Temperature = 553 ^{o}F Pressure = 1625 psia	Simulates economizers 1 and 2 of HRSG.
3	ECOSPLT (FSPLIT)	MOLE-FLOW TOIPB 1.0 RFRAC TOECON3 1.0	Simulates the splitting of the heat stream coming out the economizer block.
4	ECONOMZ3 (HEATER)	Temperature = 549 ^{o}F Pressure = 1600 psia	Simulates economizer 3 of HRSG.
5	FGSSPLIT (FSPLIT)	MOLE-FLOW FGSMAKUP 1.0 RFRAC HPBFW 1.0	This block provides hot water for fuel gas saturator.
6	HPBOILER (FLASH2)	Pressure = 1545 psia Vfrac = 0.97	Simulates a high pressure steam boiler in HRSG.
7	SUPERHTR (HEATER)	Pressure = 1465	Simulates the steam superheater in HRSG.
8	IPBOILER (HEATER)	Pressure = 350 psia Vfrac = 0.97	Simulates a 350 psia steam boiler.
9	PUMP180 (COMPR)	TYPE = 1 Pressure = 180 psia	Simulates a pump which delivers water to the 100 psia steam boiler.
10	BOIL100 (FLASH2)	Pressure = 100 psia	This block simulates a low pressure (100 psia) steam boiler.
11	SPLIT100 (FSPLIT)	MOLE-FLOW SLXSTM 0.1 RFRAC STM100 1.0	This block splits the steam from BOIL100. The splits are set by FORTRAN block SETSTEAM.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

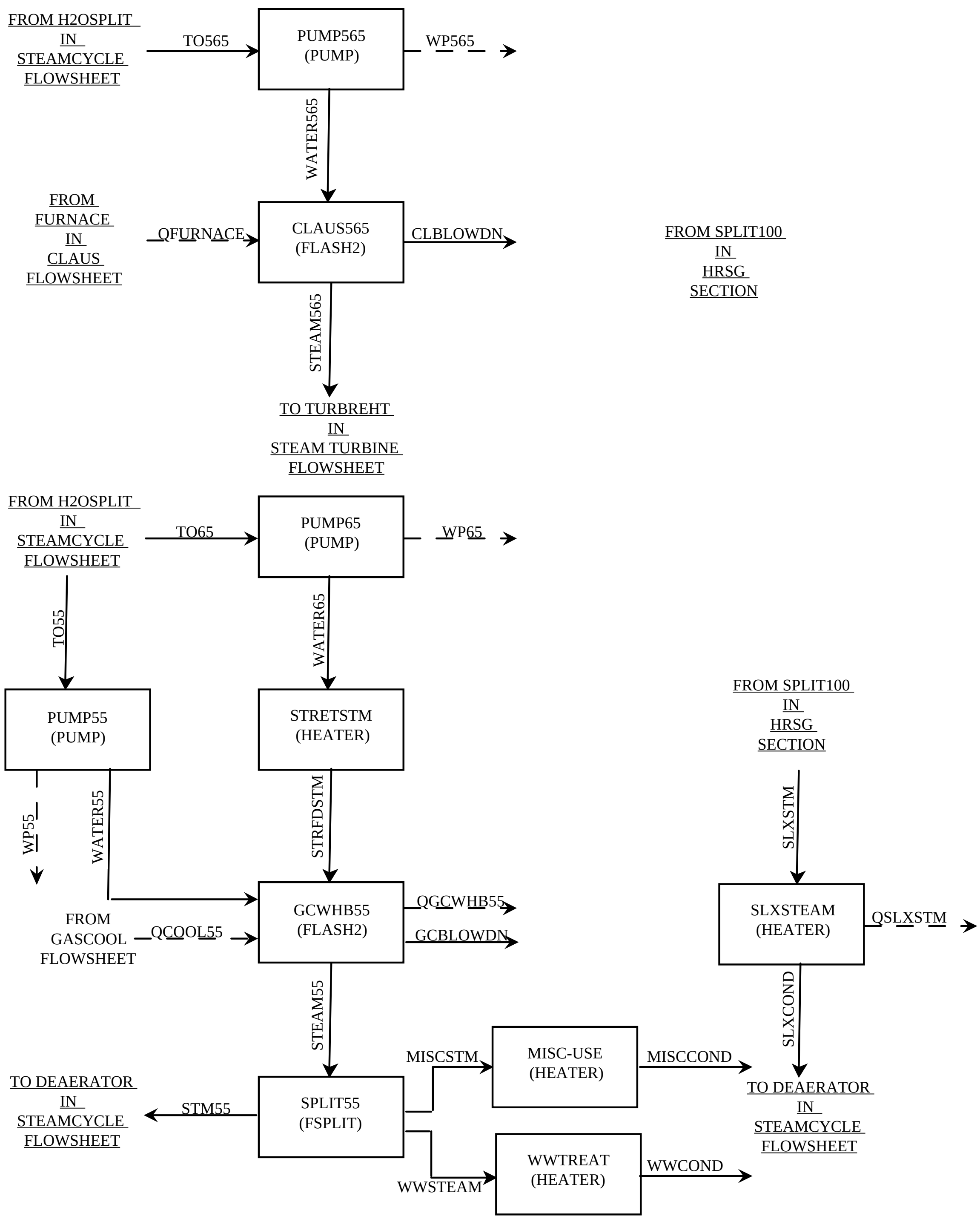

Figure 7.5 Auxiliaries Flowsheet

Table 7.4 Auxiliaries Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	PUMP565 (PUMP)	TYPE = 1 Pressure = 565 psia	This block simulates a pump which delivers water to the Claus plant steam generator.
2	CLAUS565 (FLASH2)	Pressure = 565 psia	This block simulates the Claus plant steam generator.
3	PUMP65 (PUMP)	TYPE = 65 Pressure = 65 psia	This block simulates a pump which delivers water to the BS plant steam generator.
4	STRETSTM (HEATER)	Pressure = 65 psia	This block simulates the BS plant steam generator.
5	SLXSTEAM (HEATER)	Pressure = 115 psia Vfrac = 0	This block simulates the 115 psia steam condensation in the Selexol process.
6	PUMP55 (PUMP)	TYPE = 1 Pressure = 55 psia	Simulates a pump that delivers water to GCWHB55.
6	GCWHB55 (FLASH2)	Pressure = 55 psia Vfrac = 1	Simulates 55 psia steam heater.
7	SPLIT55 (FSPLIT)	MOLE-FLOW WWSTEAM 1.0 MISCSTM 1.0 RFRAC STM55 1.0	This block splits the steam from DESUPER. The splits are set by FORTRAN block SETSTEAM.
8	WWTREAT (HEATER)	Pressure = 55 psia Vfrac = 0	Simulates the condensation of 55 psia steam condensation in Texaco Waste Water Treatment.
9	MISC-USE (HEATER)	Pressure = 55 psia Vfrac = 0	This block simulates the miscellaneous user of 55 psia steam.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

7.1.3.3 Steam Turbine

The details regarding the modeling of the steam turbine section are given in Figure 7.6 and Table 7.5. Three steam turbines are modeled in this section: TURB350, TURB90, and TURB1. The steam generated in the HRSG section is expanded through these three turbine stages, consisting of a 350 psia pressure exhaust turbine followed by an intermediate pressure turbine of exhaust pressure 90 psia, followed by a low pressure (1 psia) exhaust turbine.

The superheated steam from the HRSG section, SHSTEAM enters the block TURB350, which simulates a 350 psia steam turbine. The output stream of TURB350, STEAM350 is mixed in the block TURBREHT with the stream STEAM565 from the auxiliaries section and stream IPSTEAM from the HRSG section. The output stream of TURBREHT, HOTSTEAM is sent to intermediate pressure (90 psia) steam turbine, TURB90. The stream STEAM90 from TURB90 flows to the low pressure steam turbine, TURB1 generating 1 psia steam which flows to the block CONDENSR in the steam cycle flowsheet The work streams, WT350, WT90, and WT1 are summed to estimate the shaft power input to the generator.

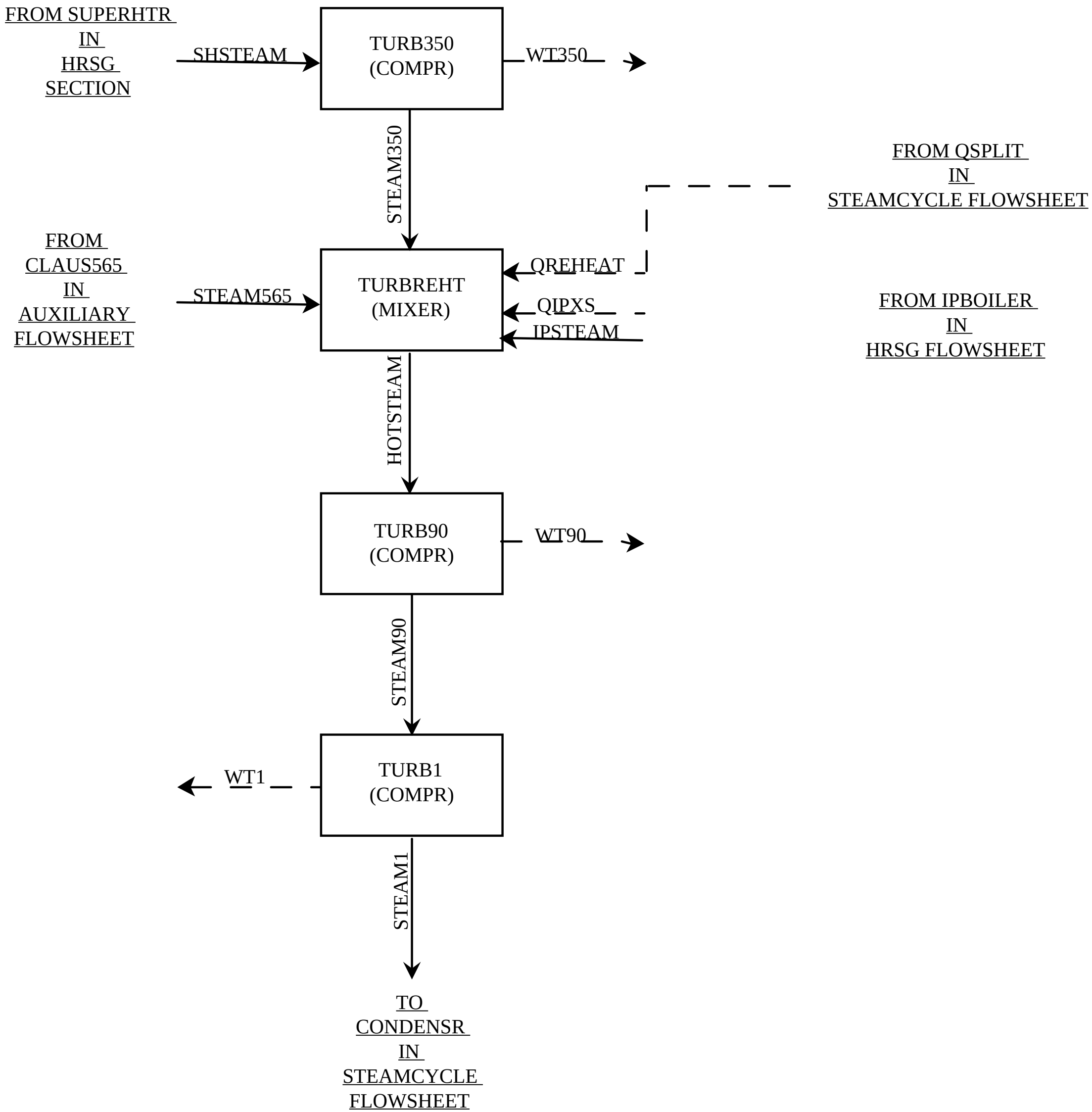

Figure 7.6 Steam Turbine Flowsheet

Table 7.5 Steam Turbine Section Unit Operation Block Description

NO	BLOCK ID (ASPEN BLOCK NAME)	BLOCK PARAMETERS	DESCRIPTION
1	TURB350 (COMPR)	TYPE = 3 Pressure = 350 psia Isoentropic eff. = 0.847	Simulates a high pressure steam turbine.
2	TURBREHT (MIXER)		This block simulates the mixing of steams at 350 psia and 565 psia.
3	TURB90 (COMPR)	TYPE = 3 Pressure = 90 psia Isoentropic eff. = 0.901	Simulates an intermediate pressure steam turbine.
4	TURB1 (COMPR)	TYPE = 3 Pressure = 1 psia Isoentropic eff. = 0.849	Simulates a low pressure (1 psia) steam turbine.

The user assigned unit operation block identification and the ASPEN unit operation block name are given.
For a glossary of ASPEN block names, please see Table A.1 in Appendix A.
For a glossary of ASPEN block parameters, please see Table A.2 in Appendix A.

7.1.4 Plant Energy Balance

The plant energy balance is comprised of four energy balance calculations. They are: (1) the gas turbine section power output estimation; (2) the estimation of the total gross power output of the steam turbine; (3) the estimation of power consumption of auxiliary pumps modeled in the ASPEN flowsheet; and (4) the estimate of all other process area auxiliary loads. The latter are calculated in the cost model subroutine. The approach to calculating the plant energy balance is the same as described in the Section 3.2.7.

The auxiliary power consumption models for oxidant feed, coal slurry preparation, Beavon-Stretford plant, general facilities section are similar to those used in the radiant and convective design as elaborated upon in Chapter 4.0. The sections which use different auxililary power models than those in the radiant and convective design are described below.

7.1.4.1 Gasification

Only two data points were available for the determination of the auxiliary power consumption model for the gasification section based upon water quench high temperature syngas cooling. The two data points were obtained from studies by Matchak *et al*. (1984) and Robin *et al*. (1993). A linear model with zero intercept was developed based upon the coal flow rate (as-received basis) per gasifier train and is shown in Figure 7.7. The auxiliary model developed has a standard error of 16 kW for the entire plant and R^2 of 0.970.

$$W_{e,\,CH} = 0.111\, N_{T,\,G}\, (m_{cf,\,G,\,i} / N_{o,\,G}) \tag{7-2}$$

where,

$W_{e,\,CH}$ = Auxiliary power consumption of the gasification process, kW.

$m_{cf,\,G,\,i}$ = Coal feed rate, tons/day.

$1300 \le m_{cf,\,G,\,i} \le 2400$ tons/day per train as received.

The R^2 variable is very high because only two data points were available.

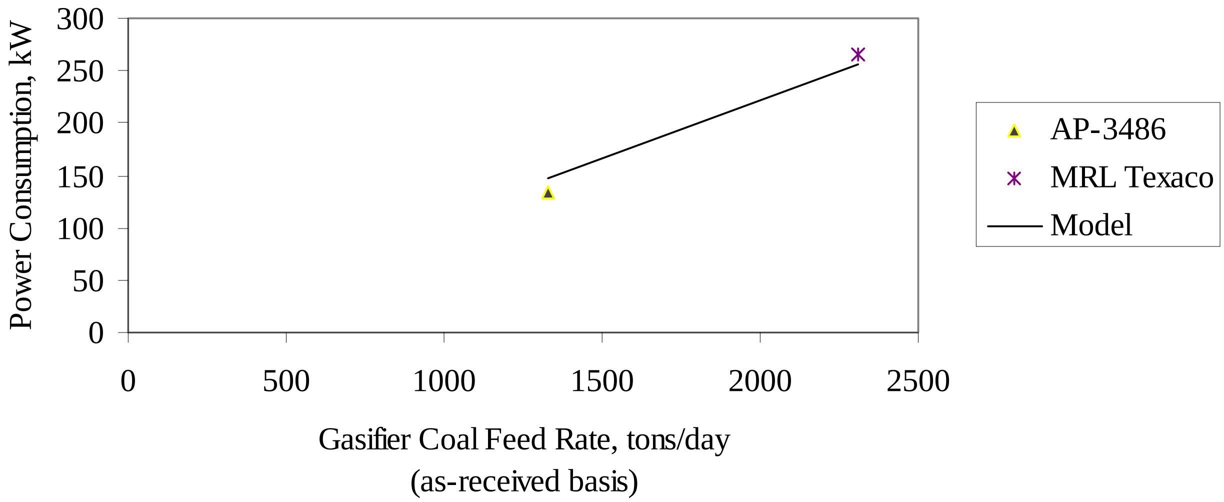

Figure 7.7 Power Requirement for the Gasification Section for Total Quench

7.1.4.2 Low Temperature Gas Cooling

The auxiliary power consumption model for the low temperature gas cooling (LTGC) section was developed using a single data point from the study by Matchak *et al*. (1984) and is given by:

$$W_{e,\,LT} = 3.211\ M_{SN,LT,O} \tag{7-3}$$

where,

$W_{e,\,LT}$ = Auxiliary power consumption for LTGC section, MW

$M_{SN,LT,O}$ = Molar flowrate of syngas to LTGC section, lbmole/hr.

7.1.4.3 Selexol

The auxiliary power consumption model for the Selexol section was developed as part of the current study using a single data point from the study by Matchak *et al*. (1984) The auxiliary power consumption model for Selexol process in MW is given by

$$W_{e,\,S} = 2.07 \text{ x } 10^{-5}\ M_{syn,S,I} \qquad (7\text{-}4)$$

where,

$M_{SYN,S,I}$ = Molar flow rate of syngas entering Selexol process, lbmole/hr.

7.1.4.4 Claus Plant

The auxiliary power consumption model for the Claus plant was developed as part of the current study using a single data point from the study by Matchak *et al*. (1984) The auxiliary power consumption model for Claus plant in MW is given by

$$W_{e,\,C} = 1.4055 \text{ x } 10^{-5}\ M_{s,C,o} \qquad (7\text{-}5)$$

where,

$M_{S,C,O}$ = Mass flow of sulfur from Claus plant, lb/hr.

7.1.4.5 Process Condensate Treatment

The process condensate treatment plant has the following auxiliary power consumption model, which is developed for the present Texaco total quench gasification system using a single data point from the study Matchat *et al*., (1984) and is given in MW by the equation:

$$W_{e, PC} = 9.289 \times 10^{-7} M_{S,BD} \quad (7\text{-}6)$$

where,

$M_{S,BD}$ = Scrubber blowdown flowrate, lb/hr.

7.2 Convergence Sequence

The convergence sequence for the total quench model simulation is similar to the convergence sequence specified in the radiant and convective design as described in Section 3.3. The additional blocks used for designing the total quench section of the model replace the high temperature gas cooling section of the model containing the radiant and convective design and the additional economizers in the total quench model are added the convergence sequence developed for the radiant and convective model.

7.3 Environmental Emissions

NO_X, particulate, SO_2, and CO_2 emissions are modeled in the same method as done in the radiant and convective design.

7.4 Capital Cost Model

This section documents the cost model developed for the Texaco gasifier-based IGCC plant with total quench high temperature gas cooling. New direct capital cost models for major process sections are presented here. For the purpose of estimating the direct capital costs of the plant, the IGCC plant is divided into thirteen process areas as listed in Table 5.1. The direct cost of a process section can be adjusted for other years using the appropriate Chemical Engineering Plant Cost Index (PCI) as shown in Table 5.2 and described in Section 5.1.

The direct capital cost models for coal handling section, oxidant feed section, Claus recovery section, Beavon-Stretford plant, gas turbine, boiler feedwater system, process condensate system, heat recovery steam generation system, steam turbine section, and general facilities are the same as those in radiant and convective design for coal. The process area direct capital costs for gasification, low temperature gas cooling, and Selexol are different from those in the radiant and convective system and are described here.

7.4.1 Gasification Section

The Texaco gasification section of an IGCC plant contains gas scrubbing, gas cooling, slag handling, and ash handling. For IGCC plants of 400 MW to 1100 MW, typically four to eight operating gasification trains are used along with one spare train (Matchak *et al.*, 1984).

Only two data points were available for the development of this cost model. The data points are not conducive to cost model development using regression analysis, since a straight line connecting them would have a negative slope. Therefore, a representative value based upon the average of the two points is used to represent the direct cost of a single gasifier train. A plot of the data is given Figure 7.8. From the two data points an approximation was determined to be 10 million January dollars per train. Since the data are based upon a coal feed rate of 1,300 to 2,300 tons/day (as-received basis), the average cost is assumed to apply for individual trains in this size range. The direct capital cost in January 1989 dollars for the gasification section is:

$$DC_G = 10{,}000{,}000\ N_{T,G} \tag{7-7}$$

where,

$N_{T,G}$ = Number of trains

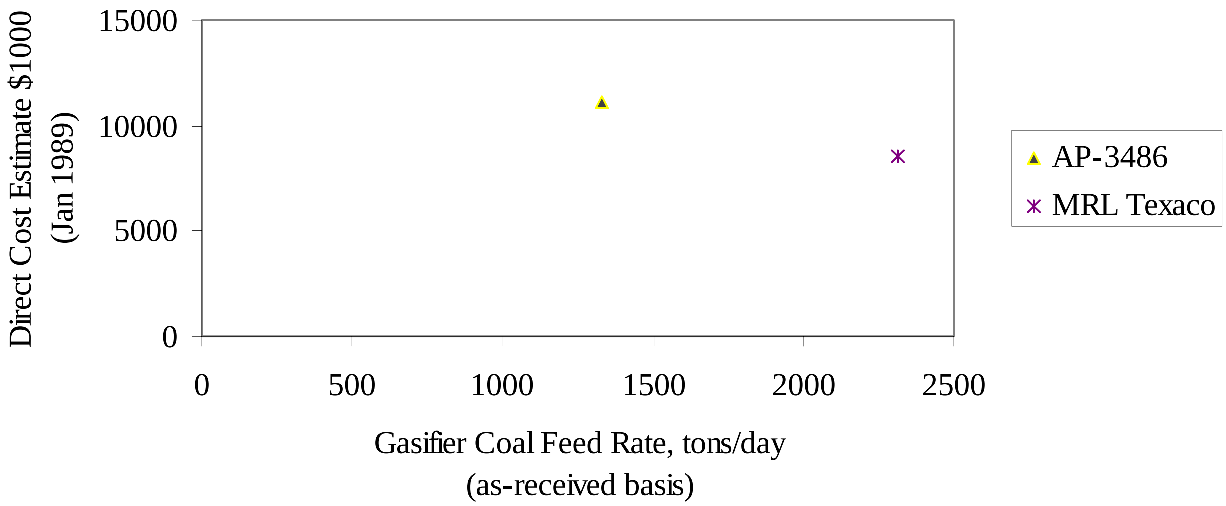

Figure 7.8 Direct Cost for Total Quench Cooled Gasifier

7.4.2 Low Temperature Gas Cooling

The direct cost model for the low temperature gas cooling section of the total quench model is similar to the one used for radiant and convective model, with a small modification such that the total quench model reflects a data point obtained from the study by Matchak *et al*. (1984). The direct cost model developed is:

$$DC_{LT} = 4.5492\, N_{T,LT} \left(\frac{M_{syn,LT,o}}{N_{O,LT}} \right)^{0.79} \qquad (7\text{-}8)$$

where,

$M_{syn,LT,o}$ = Molar flow of syngas from the LTGC section, lbmol/hr

$N_{O,LT}$ = Number of operating trains

$N_{T,LT}$ = Total number of trains

7.4.3 Selexol Section

The same direct cost model for Selexol section is used as that in the radiant and convective design except for a small modification of the coefficient in the equation. This modification was done to match a data point obtained from the study by Matchak *et al*. (1984). The direct capital cost model for the Selexol section is:

$$DC_S = \frac{0.2746\, N_{T,S}}{(1-\eta)^{0.059}} \left(\frac{M_{syn,S,i}}{N_{O,S}} \right)^{0.980} \tag{7-9}$$

where,

$$2,000 \le \left(\frac{M_{syn,G,i}}{N_{O,S}} \right) \le 67,300\, lbmole/hr;\, and$$

$$0.835 \le \eta_{HS} \le 0.997.$$

200 • $W_{ST,E}$ • 550 MW.

8.0 APPLICATION OF THE PERFORMANCE, EMISSIONS, AND COST MODEL OF THE COAL-FUELED IGCC SYSTEM WITH TOTAL QUENCH GAS COOLING TO A DETERMINISTIC CASE STUDY

An example case study is presented here to illustrate the use of the new IGCC system model for the coal-fueled system with total quench high temperature gas cooling. The key steps in running the ASPEN simulation model of the Texaco gasifier-based IGCC system are: (1) specify input assumptions; (2) execute the model; (3) collect results; and (4) interpret the results.

8.1 Input Assumptions

Model input assumptions were developed for the performance and cost model similar to those developed for radiant and convective model. The model is configured to represent three parallel trains of heavy duty “Frame 7F” gas turbines. Table 8.1 summarizes a number of the input assumptions for the example case study, with a focus on the key inputs for the gasifier and gas turbine process areas of the model. Many of these assumptions have been previously described in the technical description of the technology. Two of the assumptions listed in the table are initial values that may be modified during the simulation. These are the Oxygen/Coal ratio in the gasifier and the Turbine Inlet Temperature in the gas turbine. The Oxygen/Coal ratio is varied by a design specification in order to achieve the specified syngas exit temperature and overcome a two percent heat loss from the gasifier. The Turbine Inlet Temperature may be lowered from the initial value of 2,350 ^{o}F in order to maintain the exhaust gas temperature below 1,120 ^{o}F. There are literally hundreds of other input assumptions to the model. Only the most significant ones affecting plant design and operation are shown here. The cost model assumptions used in this case study are similar to those reported by Frey and Rubin (1991).

8.2 Model Results

The version of ASPEN used in the present study is the one developed by US Department of Energy. To execute the ASPEN model, an input file is prepared using standard ASPEN keywords and is submitted to a multi-step process leading to model execution. In the first step, the input file is translated into a FORTRAN program, which is then compiled and linked to the extensive library of ASPEN unit operation and other subroutines. The model is then executed and produces numerous output files. This particular case study was executed on a VAX 4000 located at Carnegie Mellon University, and the clock time for the run was about 5 minutes.

Selected performance and cost results from the model output are summarized in Tables 8.2 and 8.3. The overall energy balance is indicated in Table 8.2. The plant is estimated to produce a net of 793 MW with an overall plant efficiency of 35.0 percent on a higher heating value basis. The breakdown of plant power production and internal plant power consumption for auxiliaries is given in the Table 8.3.

Estimated emission rates for SO_2, NO_x, particulate matter (PM), and CO_2 are provided in Table 8.2. The estimated costs for the IGCC system given in Table 8.3 include capital, annual, and levelized costs. These costs are inclusive of the entire power plant, including the environmental control system. The breakdown of total capital cost of $1,540/kW includes a 47.2 percent contribution from direct costs, a 4.7 percent contribution from process contingencies, a 12.1 percent contribution from project contingencies, and a 13.1 percent contribution from allowances for funds used during construction. The remaining contributions are from other indirect costs and startup costs. The largest annual cost is for fuel consumption. The byproduct credit for sale of elemental sulfur offsets the incremental variable costs for all consumables other than fuel. The levelized cost of electricity, based upon a 65 percent capacity factor, is 47.7 mills/kWh (4.77 cents/kWh). This cost of electricity is comparable to that of many other coal-based power generation systems evaluated using similar financial assumptions.

Table 8.1 Summary of the Base Case Parameters Values for the Texaco Coal Gasification Total Quench System

Description, Units	Value
Gasifier Process Area	
Gasifier Pressure, psia	615
Gasifier Outlet Temperature, °F	2,400
Oxygen/Coal Ratio, lb O_2 / lb Coal (Initial Value)	0.915
Slurry Water/Coal Ratio, lb H_2O / lb Coal	0.504
Quench Cooler Outlet Temperature, °F	433
Gas Turbine Process Area	
Inlet Syngas Temperature, °F	526
Fuel Moisturization, wt-% of Clean Gas	40.0
Pressure Ratio	15.5
Turbine Inlet Temperature, °F (Initial Value)	2,350
Compressor Isentropic Efficiency, %	81.0
Expander Isentropic Efficiency, %	91.9
Generator Efficiency, %	98.5

Table 8.2 Summary of Selected Performance Model Results from the Example Case Study

Description, Units	Value
Gas Turbine Output, MW	615.0
Steam Turbine Output, MW	293.7
Auxiliary Power Demand	
Coal Handling, MW	7.6
Oxidant Feed, MW	86.3
Gasification, MW	0.8
Low T. Cool., MW	1.8
Selexol, MW	1.2
Claus, MW	0.3
Beavon-Stretford, MW	1.3
Steam Cycle, MW	4.6
Process Condensate, MW	1.3
General Facilities, MW	10.5
Total Auxiliary Load, MW	115.6
Net Power Output, MW	793.0
Heat Rate, BTU/Kwh (HHV basis)	9,478
Efficiency, %	35.0
SO_2 Emissions, lb/10^6 BTU	0.22
NO_x Emissions, lb/10^6 BTU	0.12
Particulate Matter (PM) Emissions, lb/10^6 BTU	< 0.03
CO_2 Emissions, lb/kWh	1.91

Table 8.3 Summary of Cost Model Results for the Example Case Study (1998 Dollars)

Description, Units	Value
Capital Cost Summary ($/kW)	
Total Direct Cost	728
Total Indirect Costs	267
Process Contingencies	73
Project Contingency	187
Total Plant Cost	1,256
AFDC (see note below)	201
Total Plant Investment	1,457
Startup Costs and Land	68
Total Capital Requirement[a]	1,540
Fixed Operating Cost, $/(kW-yr)	42.6
Incremental Variable Costs, mills/kWh	1.6
Byproduct Credit, mills/kWh	1.7
Fuel Cost, mills/ kWh	12.3
Variable Operating Cost, mills/kWh	12.2
Cost of Electricity, mills/kWh	47.7

Note: AFDC = Allowances for Funds used During Construction
Fuel Cost, $/MMBTU = 1.26 (Jan 1998 Dollars)
Capital Recovery Factor = 0.1034
[a] = Total Capital Requirement inlcudes Total Plant Investments, Startup costs and Land, Inventory Capital, Initial Catalysts and Chemicals

9.0 UNCERTAINTY ANALYSIS

Process technologies that are still in the research phase are subject to uncertainty with respect to prediction of performance, emissions, and costs. Insights into risks of such new technologies are obtained by analyzing the uncertainties associated with them. Uncertainty and sensitivity analysis of technology assessment models are done to find out which assumptions and uncertainties may affect the conclusions significantly.

In any type of modeling effort, the limitations of data and of knowledge about the system should be reflected in the model results. Uncertainties are prevalent in the early stages of any technology development effort and hence must be incorporated in the analysis and design of the technology. Uncertainty analysis has been described as "the computation of the total uncertainty induced in the output by quantified uncertainty in inputs and models, and the attributes of the relative importance of the input uncertainties in terms of their contribution" (Morgan and Henrion, 1990). Incorporating uncertainties in the development of new technology model helps in: (1) identifying robust solutions to process design questions and to eliminate inferior design options; (2) identifying key problems areas in a technology failure; (3) comparing competing technologies on a consistent basis to determine the risks associated with adopting a new technology; and (4) evaluating the effects that additional research might have on comparisons with conventional technology (Frey and Rubin, 1991).

In probabilistic analysis, uncertainties in model input parameters are represented using probability distributions. Using probabilistic simulation techniques, simultaneous uncertainties in any number of model input parameters can be propagated through a model to determine the combined effect on model outputs. The result of a probabilistic simulation includes both the possible range of values for model output parameters and information about the likelihood of obtaining various results. This provides insight into risks and potential pay-offs of a new technology. Statistical analysis on the inputs and output data can be used to identify trends without the need to re-run the analysis. Thus, probabilistic analysis can be used to the identify the uncertainties in a process that matter the most.

9.1 Methodology for Probabilistic Analysis

9.1.1 Characterizing Uncertainties

There are three general areas of uncertainty that should be explicitly reflected in engineering models. There are uncertainties in: (1) process performance parameters (e.g., heat losses and removal efficiencies); (2) process area capital cost; and (3) process operating cost (Frey and Rubin, 1992b). The approaches to developing probability distributions for model input parameters are similar in many ways to the approach one might take to pick a single "best guess" number for deterministic (point-estimate) analysis or to select a range of values to use in sensitivity analysis. However, the development of estimates of uncertainty usually requires more detailed thinking about possible outcomes and their relative likelihoods.

The steps involved in estimating uncertainties for model input parameters are (Frey and Rubin, 1992,a):

1. Review the technical basis for uncertainty in the process;
2. Identify candidate parameters that should be treated as uncertain;
3. Determine the sources of information regarding uncertainty for each parameter; and
4. Develop estimates of uncertainty depending on the availability of information.

Estimates of uncertainty in terms of ranges and probability distributions for model input parameters can be based on: (1) published judgements in the literature; (2) published information, both quantitative and qualitative, that can be used to infer a judgement about uncertainty; (3) statistical analysis of data; and (4) judgements elicited from technical experts with relevant expertise (Frey and Rubin, 1990; Morgan and Herrion, 1990).

Probability distributions for uncertainty can be developed from available data using statistical techniques. The data can be fitted to a particular distribution using various statistical tests. When the data available are limited, engineering insight can be used to supplement the data in coming up with an appropriate probability distribution for the uncertain variable.

When sufficient data are not available, judgements from technical experts can be elicited to obtain an appropriate probability distribution for the uncertain variable. In designing elicitation protocol, it is important to take into account heuristics by which judgements about uncertainty may be. Some heuristics can lead to biases in the judgements. However, protocols can be designed to counteract these sources of bias.

9.1.2 Types of Uncertain Quantities

Many types of random variation should be considered in developing a probability distribution for a variable. These are briefly discussed in Frey and Rubin (1991) and are reviewed here.

9.1.2.1 Variability

Variability is caused due to variations in the process itself. For example, a variation in the coal composition will cause a variation in the net efficiency of the plant. Variability can be represented as a probability distribution.

9.1.2.2 Uncertainty

Uncertainty represents the lack of knowledge regarding the true value of a quantity. There are a number of types of uncertainty which can be considered while developing a probability distribution for a variable. Variability is conceptually distinct from uncertainty (Frey and Rubin, 1992b). For example, for a given coal composition, the carbon conversion may be uncertain.

1. Statistical Error – is associated with imperfections in measurement techniques. Statistical analysis of test data is thus one method for developing a representation of uncertainty in a variable.

2. Systematic Error – The mean value of a quantity may not converge to the "true" mean value because of biases in measurement and procedures. Such biases may arise from imprecise calibration, faulty reading of meters, and inaccuracies in the assumptions used to infer the actual quantity of interest from the observed readings of other quantities.

Uncertainty may also arise due to lack of experience with a process. This type of uncertainty often cannot be treated statistically because it requires predictions about something that has yet to be built or tested. This type of uncertainty can be represented using technical estimates about the range and likelihood of possible outcomes.

9.1.3 Some Types of Probability Distributions

An expert may specify a judgement regarding uncertainties using different types of probabilistic distributions. One way of representing a probability distribution is the cumulative distribution function (CDF), which shows the probability fractiles on the y-axis and the value of the fractile associated with each fractile on the x-axis. Some commonly used probability distributions are shown in Figure 9.1.

1. Uniform - represents uniform probability of obtaining a value between upper and lower limits.
2. Triangle - - represents uniform probability of obtaining a value between upper and lower limits with values biased toward a modal value specified.
3. Normal – is a symmetric distribution with mean, mode, and median at the same point. It is often assumed in statistical analysis as the basis for unbiased measurement errors.
4. Lognormal – is a positively skewed distribution and has a long tail to the right.

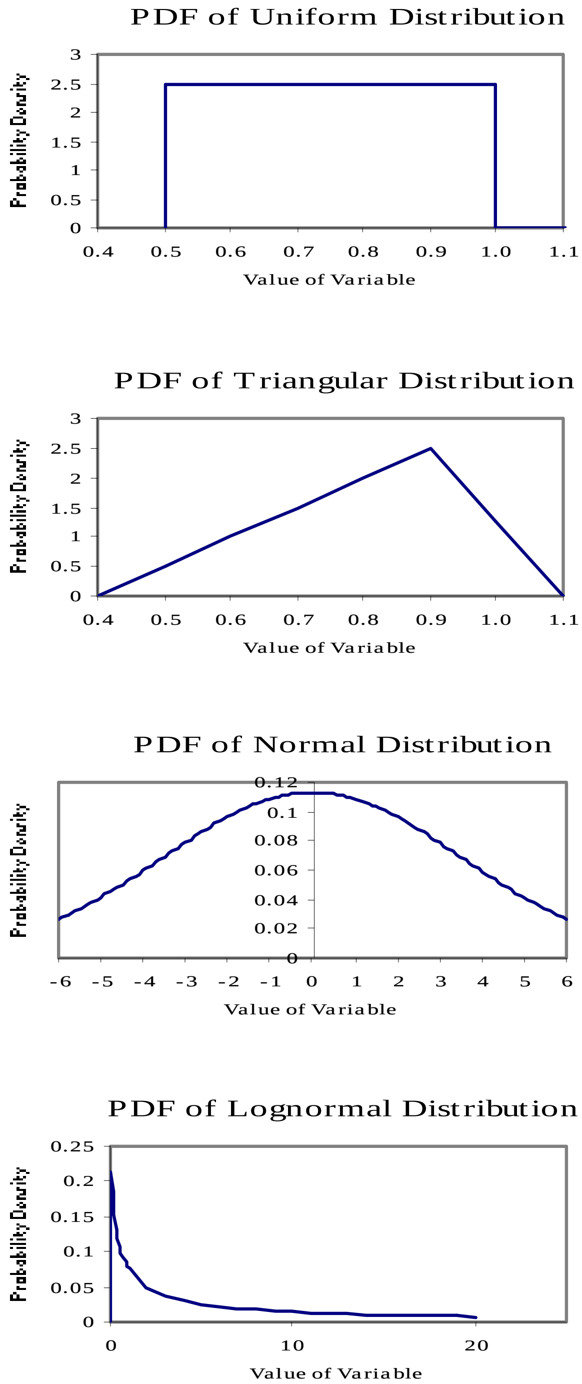

Figure 9.1 Examples of Probability Density Functions

9.1.4 Monte Carlo Simulation

A probabilistic modeling environment is required to analyze uncertainties in advanced process technologies. Monte Carlo simulation is one such typical environment (Ang and Tang, 1984). In this approach, model is run repeatedly, using different values for each of the uncertain input parameters each time. The values of each of the uncertain input parameters are generated based on the probability distribution for the parameters. In each repetition in the simulation, one value for each of the input parameters is sampled simultaneously. The set of sampled values generated for each of the model output variables can be analyzed statistically treating them as experimental set of data.

The execution of the model for a given set of samples in a repetition is deterministic, although the generation of samples values for the input parameters is probabilistic. However, the Monte Carlo method has the advantage that these deterministic simulations are repeated in manner that yields important insights into the sensitivity of the model to variations in the input parameters, as well as into the likelihood of obtaining any particular outcome.

Monte Carlo methods allow the modeler to use any type of probability distribution for which values can be generated on a computer, rather than to be restricted to forms which are analytically tractable. The set of samples obtained for model outputs can be represented as cumulative distribution functions and summarized using typical statistics such as mean and variance.

In a random Monte Carlo simulation, one approach to generating sample values is to use the inverse CDF method. A random number generator is used to generate uniformly distributed numbers between 0 and 1 for each uncertain variable. Thus, the generated random numbers are used to represent the fractile of the random variable for which a sample is to be generated. The sample values for the random variables are calculated using the inverse cumulative distribution functions (CDF's) based on randomly generated fractiles.

Latin Hypercube Sampling (LHS) is an alternative to random Monte Carlo simulation. In LHS, the fractiles that are used as inputs to the inverse CDF are not randomly generated. Instead, the probability distribution for the random variable of interest is first divided into ranges of equal probability, and one sample is taken from each equal probability range. However, the order of the samples is random over the course of simulation, and the pairing of samples between two or more random input variables is usually treated as independent. In random LHS, one sample is randomly taken from each equal probability interval, while in median LHS one sample is taken from the median of the interval (Morgan and Henrion, 1990).

LHS methods guarantee that values from the entire range of distribution will be sampled proportional to the probability density of the distribution. Thus the input samples typically cover a full span of each parameter's probability density function compared to when the random Monte Carlo method is used (McKay *et al*., 1979). The number of samples required to adequately represent the CDF for a distribution is less for LHS than for random Monte Carlo sampling. The LHS method was employed in the present study.

9.1.5 Methods for Identifying Key Sources of Uncertainty in Model Inputs

A probabilistic modeling capability has been added to the publicly available version of ASPEN (Diwekar and Rubin, 1989). A FORTRAN program developed by Iman and Shortencarier (1984) using LHS was adopted for assigning probability distributions to model parameters and generating samples from those distributions. In order to identify the key sources of uncertainty in the model inputs, linear correlations between the input variables and model outputs can be determined. Linear correlations between uncertain input variables and the model outputs are identified using techniques such as standardized regression coefficients (SRC) and partial correlations (PCC). A FORTAN program which calculates the partial correlation and standardized regression coefficients was used for analysis of model output (Iman *et al*., 1985).

The standard regression coefficient of an input variable is used to measure the relative contribution of the uncertainty in the input variable to the uncertainty of the output variables. For this analysis, all the sample values for the input variables are standardized. The standardization process involves subtracting the mean of the variable from all the sample values and then dividing by the variable's standard deviation. A multi-variate regression is performed for an output variate based on the inputs. The relative importance of each input variate is indicated by the regression coefficient of that variate, which is the standardized regression coefficient (SRC). SRCs are the partial derivatives of the output variable with respect to each input variable. SRCs measure the shared contribution of the input to the output as all of the simulation input uncertainties are included in the regression analysis simultaneously.

The partial correlation coefficient analysis is used to identify the degree to which correlations between output and input random variables may be linear, and it is estimated in conjunction with multi-variate linear regression analysis using a step-wise procedure. The input variable most higly correlated with the output variable of interest is assumed as the starting point for construction of a stepwise linear regression model. In the regression model, the output variable is treated as a dependent variable and the most highly correlated input variable is treated as a predictive variable. The PCC technique then searches for another input variable which is most highly correlated with the residuals of the regression model already in containing the first input variable. The residual is the difference between the actual sample value of the dependent variable and the estimated sample values, using the linear regression model already containing the first input variable. The process is repeated to add more variables in the analysis. The PCC is a measure of the unique relationship between input and dependent variables that cannot be explained by variables already included in the regression analysis (Frey and Rubin, 1992).

PCC and SRC analysis is limited to cases where the relationship between input and output variables is linear. However, these techniques can be extended to monotonic non-linear cases by performing regressions on the ranks, rather than the sample values of the inputs and outputs. They are known as partial rank correlation coefficients (PRCC) and standardized rank regression coefficients (SRCC).

The regression techniques are useful for identifying the contribution of each input variable to variations in the output variable. However, they cannot be used to identify which input variables may be responsible for a shift in the central tendency of the model outputs associated with skewness in the input distributions. In such cases, sensitivity analysis is performed by gradually making one or more input variables uncertain while setting point estimates to the remaining input variables and observing the output distribution. The sensitivity analysis is continued till the current model output distribution closely resembles the original model output distribution in which all the input variables are uncertain.

9.2 Input Assumptions for Probabilistic Case Studies

In this section, the base case assumptions regarding uncertainty in specific performance and cost parameters of the Texaco gasifier-based IGCC system models developed in the present study are given as listed in Table 9.1, Table 9.2.

Tables 9.1 and 9.2 list the performance, environmental, and cost variables selected for stochastic analysis, along with the deterministic value and distributions for each of these variables. Since the gas turbine and steam cycle/steam turbine technology is well established, the performance variables of these process areas were not considered for uncertainty analysis.

A total of 40 parameters are treated as uncertain in the two cases. These include assumptions regarding the performance of the gasifier and gas turbine process areas, capital cost parameters, direct capital costs, maintenance costs, labor rate, and unit costs. The deterministic values are based upon the assumptions used in published design studies.

The estimates of uncertainty in the capital cost parameters, including engineering and home office fees, indirect construction cost factor, and project uncertainty are based on typical ranges of values for these parameters suggested by EPRI (EPRI, 1986). The basis for these estimates have been discussed by Frey and Rubin (1991).

The deterministic values for the process contingency factors had been adopted from assumptions in published design studies (e.g., Frey and Rubin, 1991; Dawkins *et al.*, 1985). For the purposes of preliminary characterization of uncertainty in capital cost, it was assumed that the process contingency factors were intended to represent the mid-point of symmetric uncertainty distributions for process area direct cost. The relative magnitudes of the contingency factors were assumed to suggest the relative magnitude of the variances to be used. Uniform distributions between the best and worst values were assumed for some of the process areas, while triangular distribution was assumed for the other process areas. The triangular distribution was selected in cases where the author felt that the published contingency factors were carefully developed. The effect of a triangular distribution, compared to a uniform distribution, is to place more "weight" on the outcomes near the published contingency factor than on the extreme high or low outcomes. An exception to the above described approach is the estimate of uncertainty in the gas turbine process area and is elaborated upon in Frey and Rubin (1991).

The estimates of uncertainty in maintenance cost factors use deterministic values from published design studies as starting points, similar to the estimates of uncertainty in direct costs. However, it is assumed that the maintenance costs are more likely to increase than decrease compared to the deterministic values. This assumption is based on the fact that IGCC systems must handle material streams containing various contaminants derived from coal conversion. These contaminants are likely to cause deposition, erosion, and corrosion problems in various parts of the systems and increase maintenance (Frey and Rubin, 1991).

The development of estimates for uncertainties in operating cost parameters including operating labor rate, units costs for ash disposal and byproduct sales, and byproduct marketing costs factor is similar as discussed in Frey and Rubin (1991).

Table 9.1 Summary of the Base Case Parameter Values and Uncertainties for the Coal-Fueled Texaco Gasifier-Based IGCC System with Radiant and Convective High Temperature Gas Cooling

Description	Units	Deterministic Value	Distribution	Parameters[a]		
GASIFIER PROCESS AREA						
Gasifier Pressure	psia	615	Normal	567.5	to	662.51
Gasifier Temperature	^oF	2400	Triangular	2400	to	2600
Oxygen/Oil Ratio	lb/hr O_2/ lb/hr C	0.915				
Water/Oil Ratio	lb/hr H_2O/ lb/hr C	0.504	Normal	0.465	to	0.543
Carbon Conversion	fraction	0.99	Triangular	0.96	to	1.00
Approach Temperature 1	^oF	-300	Triangular	-350	to	-250
Approach Temperature 2	^oF	-500	Triangular	-550	to	-450
Approach Temperature 3	^oF	-500	Triangular	-550	to	-450
Approach Temperature 4	^oF	-500	Triangular	-550	to	-450
Approach Temperature 5	^oF	-500	Triangular	-550	to	-450
Approach Temperature 6	^oF	-500	Triangular	-550	to	-450
Approach Temperature 7	^oF	-500	Triangular	-550	to	-450
Approach Temperature 8	^oF	-500	Triangular	-550	to	-450
Approach Temperature 9	^oF	-500	Triangular	-550	to	-450
GAS TURBINE PROCESS AREA						
Fuel gas temp. before Entering combustor	oF	570				
Fuel Moisturization	wt % of Clean gas	28.2				
Pressure Ratio	ratio	15.5				
Turbine Inlet Temp	oF	2,350				
Exhaust Flow	lb/sec	1,089				
Thermal NO_x	fraction of air nitrogen fixated	$4.5x10^{-5}$	Uniform	$2.5x10^{-5}$	to	$7.5x10^{-5}$
Unconverted CO	wt-% of CO in fuel gas	0.99985	Uniform	0.9998	to	0.9999

CAPITAL COST PARAMETERS							
Engineering and Home Office Fee	fraction	0.10	Triangular	0.07	to	0.13	(0.10)
Indirect Construction Cost Factor	fraction	0.20	Triangular	0.15	to	0.25	(0.20)
Project Uncertainty	fraction	0.175	Uniform	0.10	to	0.25	
General Facilities	fraction	0.20					
DIRECT COSTS[b]							
Coal Handling	% of DC	5					
Oxidant Feed	% of DC	5	Uniform	0	to	10	
Gasification	% of DC	15	Triangular	0	to	40	(15)
Selexol	% of DC	10	Triangular	0	to	20	(10)
Low Temperature Gas Cooling	% of DC	0	Triangular	-5	to	5	(0)
Claus Plant	% of DC	5	Triangular	0	to	10	(5)
Beavon-Stretford	% of DC	10	Triangular	0	to	20	(10)
Boiler Feed Water	% of DC	0					
Process Condensate Treatment	% of DC	30	Triangular	0	to	30	(10)
Gas Turbine	% of DC	12.5	Triangular	0	to	25	(12.5)
HRSG	% of DC	2.5	Triangular	0	to	5	(2.5)
Steam Turbine	% of DC	2.5	Triangular	0	to	5	(2.5)
General Facilities	% of DC	5	Triangular	0	to	10	(5)
MAINTENANCE COSTS[c]							
Coal Handling	% of TC	3					
Oxidant Feed	% of TC	2					
Gasification	% of TC	4.5	Triangular	3	to	6	(4.5)
Selexol	% of TC	2	Triangular	1.5	to	4	(2)
Low Temperature Gas Cooling	% of TC	3	Triangular	2	to	4	(3)
Claus Plant	% of TC	2	Triangular	1.5	to	2.5	(2)
Beavon-Stretford	% of TC	2					
Boiler Feed Water	% of TC	1.5					
Process Condensate Treatment	% of TC	2	Triangular	1.5	to	4	(2)
Gas Turbine	% of TC	1.5	Triangular	1.5	to	2.5	(1.5)
HRSG	% of TC	1.5					

Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					
<u>OTHER FIXED OPERATING COST PARAMETERS</u>							
Labor Rate	$/hr	19.70	Normal	17.70	to	21.70	
<u>VARIABLE OPERATING COST PARAMETERS</u>							
Ash Disposal	$/ton	10	Triangular	10	to	25	(10)
Sulfur Byproduct	$/ton	125	Triangular	60	to	125	(125)
Byproduct Marketing	fraction	0.10	Triangular	0.05	to	0.15	(0.10)
Fuel Cost	$/MMBTU	1.28	Trinagular	1.15	to	1.41	(1.28)

Table 9.2 Summary of the Base Case Parameter Values and Uncertainties for the Coal-Fueled Texaco Gasifier-Based IGCC System with Total Quench High Temperature Gas Cooling

Description	Units	Deterministic Value	Distribution	Parameters[a]		
GASIFIER PROCESS AREA						
Gasifier Pressure	psia	615	Normal	567.5	to	662.51
Gasifier Temperature	^oF	2400	Triangular	2400	to	2600
Oxygen/Oil Ratio	lb/hr O_2/ lb/hr C	0.915				
Water/Oil Ratio	lb/hr H_2O/ lb/hr C	0.504	Normal	0.465	to	0.543
Carbon Conversion	fraction	0.99	Triangular	0.96	to	1.00
Approach Temperature 1	^oF	-300	Triangular	-350	to	-250
Approach Temperature 2	^oF	-500	Triangular	-550	to	-450
Approach Temperature 3	^oF	-500	Triangular	-550	to	-450
Approach Temperature 4	^oF	-500	Triangular	-550	to	-450
Approach Temperature 5	^oF	-500	Triangular	-550	to	-450
Approach Temperature 6	^oF	-500	Triangular	-550	to	-450
Approach Temperature 7	^oF	-500	Triangular	-550	to	-450
Approach Temperature 8	^oF	-500	Triangular	-550	to	-450
Approach Temperature 9	^oF	-500	Triangular	-550	to	-450
GAS TURBINE PROCESS AREA						
Fuel gas temp. before Entering combustor	oF	526				
Fuel Moisturization	wt % of Clean gas	40.0				
Pressure Ratio	ratio	15.5				
Turbine Inlet Temp	oF	2,350				
Exhaust Flow	lb/sec	1,089				
Thermal NO_x	fraction of air nitrogen fixated	4.5×10^{-5}	Uniform	2.5×10^{-5}	to	7.5×10^{-5}
Unconverted CO	wt-% of CO in fuel gas	0.99985	Uniform	0.9998	to	0.9999

CAPITAL COST PARAMETERS							
Engineering and Home Office Fee	fraction	0.10	Triangular	0.07	to	0.13	(0.10)
Indirect Construction Cost Factor	fraction	0.20	Triangular	0.15	to	0.25	(0.20)
Project Uncertainty	fraction	0.175	Uniform	0.10	to	0.25	
General Facilities	fraction	0.20					
DIRECT COSTS[b]							
Coal Handling	% of DC	5					
Oxidant Feed	% of DC	5	Uniform	0	to	10	
Gasification	% of DC	15	Triangular	0	to	40	(15)
Selexol	% of DC	10	Triangular	0	to	20	(10)
Low Temperature Gas Cooling	% of DC	0	Triangular	-5	to	5	(0)
Claus Plant	% of DC	5	Triangular	0	to	10	(5)
Beavon-Stretford	% of DC	10	Triangular	0	to	20	(10)
Boiler Feed Water	% of DC	0					
Process Condensate Treatment	% of DC	30	Triangular	0	to	30	(10)
Gas Turbine	% of DC	12.5	Triangular	0	to	25	(12.5)
HRSG	% of DC	2.5	Triangular	0	to	5	(2.5)
Steam Turbine	% of DC	2.5	Triangular	0	to	5	(2.5)
General Facilities	% of DC	5	Triangular	0	to	10	(5)
MAINTENANCE COSTS[c]							
Coal Handling	% of TC	3					
Oxidant Feed	% of TC	2					
Gasification	% of TC	4.5	Triangular	3	to	6	(4.5)
Selexol	% of TC	2	Triangular	1.5	to	4	(2)
Low Temperature Gas Cooling	% of TC	3	Triangular	2	to	4	(3)
Claus Plant	% of TC	2	Triangular	1.5	to	2.5	(2)
Beavon-Stretford	% of TC	2					
Boiler Feed Water	% of TC	1.5					
Process Condensate Treatment	% of TC	2	Triangular	1.5	to	4	(2)
Gas Turbine	% of TC	1.5	Triangular	1.5	to	2.5	(1.5)
HRSG	% of TC	1.5					

Steam Turbine	% of TC	1.5					
General Facilities	% of TC	1.5					

<u>OTHER FIXED OPERATING COST PARAMETERS</u>

Labor Rate	$/hr	19.70	Normal	17.70	to	21.70	

<u>VARIABLE OPERATING COST PARAMETERS</u>

Ash Disposal	$/ton	10	Triangular	10	to	25	(10)
Sulfur Byproduct	$/ton	125	Triangular	60	to	125	(125)
Byproduct Marketing	fraction	0.10	Triangular	0.05	to	0.15	(0.10)
Fuel Cost	$/MMBTU	1.28	Trinagular	1.15	to	1.41	(1.28)

9.3 Probabilistic Analysis of the IGCC Model

For the probabilistic simulation, the deterministic performance, emissions, and cost model is executed a number of times using LHS, with a different set of values (samples) assigned to uncertain input parameters each time. The number of times the deterministic model is executed is equal to the number of observations or sample size selected. The sample size should be large enough to give sufficient precision to the numerical simulation as dictated by the use of the model results and at the same time ensure that the computational time and disk space usage are not excessive (Morgan and Henrion, 1990). To characterize the mean and the variance of the results and to identify key uncertainties, a sample size of 100 or greater is typically sufficient (Frey and Rubin, 1991). For the present study, a sample size of 120 was chosen. Results for all the uncertain output variables are collected at the end of each deterministic run, which can then be analyzed statistically to gain insight into the key uncertainties of the system. Such an analysis enables the identification of the key model uncertainties of the most important determinants of uncertainty in model outputs.

The results of the simulation can be summarized using statistics, such as mean and standard deviation, or using graphs of the cumulative distribution function and is discussed in Section 9.5.

The key uncertain variables contributing to the uncertainties in IGCC process performance were identified using three general approaches. Statistical analysis using regression techniques was used to identify input random variables which are most highly correlated with uncertainties in output variables. Probabilistic sensitivity analysis was used for identifying key uncertain inputs. In this approach, the interaction between the cost and performance uncertain input variables as they affect uncertainty in output variables can be studied by isolating the uncertainties. For example, one can assign distributions to one or more input variables while all other model inputs are assigned point estimates. The third approach, uncertainty screening, which is similar to probabilistic sensitivity analysis, can be used to confirm the results of a regression or probabilistic sensitivity analysis by deleting uncertainties from the model inputs which are not believed to be important and assigning them point estimates. The results of the screening study can be compared to the results obtained from the original probabilistic analysis and used to confirm that the deleted uncertainties do not affect the model output distributions.

A total of six case studies were performed for each technology to characterize the uncertainty in model outputs and to identify the model inputs that contributed most significantly to the distribution of values in the model outputs. The general procedure is illustrated here for the case of radiant and convective-based model. A similar procedure will be used for the two technologies evaluated. The discussions of the results for all the three technologies are presented in later sections. The purpose here is to to give a general description of the approach. The input assumptions for these case studies for the radiant and convective model are summarized in Table 9.3. Each case study is briefly described:

Case 1. Uncertainties were assigned to performance and cost inputs as described in Table 9.1 and summarized in Table 9.3. This case study has the largest number of uncertain model inputs of all the case studies.

Case 2. Uncertainties were assigned only to performance input variables as identified in Table 9.3. The results of this case study, when compared with Case 1, enable evaluation of the relative contribution of uncertainty in plant performance assumptions to the overall uncertainty in plant costs.

Case 3. Uncertainties were assigned only to cost input variables as identified in Table 9.3. The results of this case study, when compared with Case 1, enable evaluation of the relative contribution of uncertainty in cost assumptions to the overall uncertainty in plant costs.

Case 4. Only those uncertainties identified as key sources of uncertainty from a regression analysis of the results of Case 1 were assigned probability distributions, while all other model inputs were assigned point estimates. In the regression analysis of the results of Case 1, the model input variables having the highest correlation coefficients (greater than 0.5) with most of model outputs were selected as the key sources of uncertainty.

Case 5. Only those uncertainties identified as key sources of uncertainty from a regression analysis of the results of Case 2 were assigned probability distributions, while all other model inputs were assigned point estimates. In the regression analysis of the results of Case 2, the model input variables having the highest correlation coefficients (greater than 0.5) with most of model outputs were selected as the key sources of uncertainty.

Case 6. Only those uncertainties identified as key sources of uncertainty from a regression analysis of the results of Case 2 were assigned probability distributions, while all other model inputs were assigned point estimates. In the regression analysis of the results of Case 3, the model input variables having the highest correlation coefficients (greater than 0.5) with most of model outputs were selected as the key sources of uncertainty.

Tables 9.4 summarizes the above mentioned case studies for total quench coal model.

Table 9.3 List of Uncertainty Variables Used in Each of the Case Studies

	Original Models			Key Uncertainties		
	Performance and Cost	Performance Only	Cost Only	Performance and Cost	Performance Only	Cost Only
Case Study No	**1**	**2**	**3**	**4**	**5**	**6**
Gasifier Pressure	√	√				
Gasifier Temperature	√	√		√	√	
Water/Oil Ratio	√	√		√	√	
Carbon Conversion	√	√		√	√	
Approach Temperature 1	√	√				
Approach Temperature 2	√	√				
Approach Temperature 3	√	√				
Approach Temperature 4	√	√		√	√	
Approach Temperature 5	√	√		√	√	
Approach Temperature 6	√	√				
Approach Temperature 7	√	√				
Approach Temperature 8	√	√				
Approach Temperature 9	√	√				
Thermal Nox	√	√				
Unconverted CO	√	√				
Engineering and Home Office Fees	√		√	√		√
Indirect Construction Cost Factor	√		√	√		√
Project Uncertainty	√		√	√		√
Process Contingency						
Oxidant Feed	√		√	√		√
Gasification	√		√	√		√
Low Temperature Gas Cooling	√		√			
Selexol	√		√			
Claus	√		√			
Beavon-Stretford	√		√			
Process Condensate	√		√			
Gas Turbine	√		√	√		√
HRSG	√		√			
Steam Turbine	√		√			
General Facilities	√		√			
Maintenance Cost Factors						
Gasification	√		√			
Low Temperature Gas Cooling	√		√			
Selexol	√		√	√		√
Claus	√		√			
Process Condensate	√		√	√		√
Gas Turbine	√		√	√		√
Labor Rate	√		√			
Fuel Cost	√		√	√		√
Ash Disposal	√		√	√		√
Sulfur Byproduct	√		√			
Byproduct Marketing	√		√			

Table 9.4 List of Uncertainty Variables Used in Each of the Case Studies

	Original Models			Key Uncertainties		
	Performance and Cost	Performance Only	Cost Only	Performance and Cost	Performance Only	Cost Only
Case Study No	**1**	**2**	**3**	**4**	**5**	**6**
Gasifier Pressure	√	√				
Gasifier Temperature	√	√		√	√	
Water/Oil Ratio	√	√		√	√	
Carbon Conversion	√	√		√	√	
Approach Temperature 1	√	√				
Approach Temperature 2	√	√				
Approach Temperature 3	√	√				
Approach Temperature 4	√	√		√	√	
Approach Temperature 5	√	√		√	√	
Approach Temperature 6	√	√				
Approach Temperature 7	√	√				
Approach Temperature 8	√	√				
Approach Temperature 9	√	√				
Thermal Nox	√	√				
Unconverted CO	√	√				
Engineering and Home Office Fees	√		√	√		√
Indirect Construction Cost Factor	√		√	√		√
Project Uncertainty	√		√	√		√
Process Contingency						
Oxidant Feed	√		√	√		√
Gasification	√		√	√		√
Low Temperature Gas Cooling	√		√			
Selexol	√		√			
Claus	√		√			
Beavon-Stretford	√		√			
Process Condensate	√		√			
Gas Turbine	√		√	√		√
HRSG	√		√			
Steam Turbine	√		√			
General Facilities	√		√			
Maintenance Cost Factors						
Gasification	√		√			
Low Temperature Gas Cooling	√		√			
Selexol	√		√	√		
Claus	√		√			
Process Condensate	√		√	√		
Gas Turbine	√		√	√		
Labor Rate	√		√			
Fuel Cost	√		√	√		√
Ash Disposal	√		√	√		
Sulfur Byproduct	√		√			
Byproduct Marketing	√		√			

9.4 Model Results and Applications

Two IGCC systems are evaluated using probabilistic engineering models. All the systems are Texaco gasifer-based and include: (1) a case of coal-fueled system with radiant and convective high temperature cooling; and (2) a case of coal-fueled system with total quench high temperature cooling.

The IGCC system performance models are implemented in the ASPEN chemical process simulation modeling environment on a DEC VAX Station 3200 mini-computer using the public version of ASPEN with the stochastic modeling capability. The simulation process involves several steps. The performance model in ASPEN's keyword-based input language is read by the ASPEN package and converted to a FORTRAN program. This first step is called "input translation" and takes approximately 1 to 2 minutes. The ASPEN-generated FORTRAN program is compiled and linked, which also takes 1 to 2 minutes. The linked ASPEN flowsheet is executed. The final step in the simulation involves the writing of a report file containing the results of the simulation. The report writing step may take several minutes depending upon the amount of information requested by the user regarding the simulation. A single deterministic run takes a total time of about 5 to 6 minutes. In a stochastic run, the input translation, compilation and linking takes about 5 minutes and is done only once. The execution of the compiled program for 120 iterations and report writing takes about 100 minutes.

As part of the probabilistic modeling capability in the ASPEN simulator, four alternative approaches to regression analysis are available for analyzing model results. The outputs can be analyzed by partial correlation coefficients (PCC), standardized regression coefficients (SCC), partial rank correlation coefficients (PRCC), and standardized rank regression coefficients (SRCC).When running stochastic simulation in ASPEN, the user may specify which type of output analysis is desired. PRCC is the approach chosen to analyze the case studies.

Cases 1,2, and 3 are run in ASPEN and the distributions for the all outputs are obtained. The PRCC obtained in each case are analyzed and formed the basis for cases 4,5, and 6 respectively. The outputs which had a PRCC greater than 0.5 or less than –0.5 are identified as key uncertain variables. The PRCCs for the outputs are listed in Appendix C. The Cases 4, 5, and 6 are run in ASPEN using only the key uncertain variables identified. It is desired that the cdf's of the output variables obtained after the running Cases 4, 5, and 6 are identical to the cdf's obtained from the results of Cases 1, 2, and 3 respectively. However, the distributions of the uncertainties in the costs of electricity and total capital requirement in Cases 1 and 4 were not similar due to the skewness of some assumptions regarding the unit costs of consumables, some process contingency factors ,and some maintenance cost factors. This skewness results in a shift in the central value of the uncertainty in the cost of electricity when uncertainties in performance and costs are considered. To avoid this shift in the central value of the uncertainty in the cost of electricity, probabilistic sensitivity analysis was performed by introducing each of the unit cost assumptions, process contingency factors, and maintenance cost factors one by one as a key uncertainty factor. After introducing one variable as an uncertain parameter, the model was run again and the distributions were examined. This procedure was continued till the distributions of the cost of electricity of Cases 1 and 4 are similar.

Tables 9.5, and 9.6 list the distributions of selected outputs obtained from executing Case 1 for radiant and convective model and total quench coal model respectively.

Table 9.5 Selected Outputs Collected by the Model for Uncertainty Analysis

Parameter	Units	Best Guess"”	$f_{.50}$	μ	$\sigma 2$	$f_{.05}$	$f_{.95}$
Plant Performance							
Plant Thermal Efficiency	Fraction	39.41	38.91	38.88	0.46	38.03	39.52
Net Plant Output	MW	862.9	867.5	867.4	4.04	860.8	873.5
Gross Power Output	MW	980.3	988.4	988.6	6.09	978.3	998.7
Total Auxiliary Power Consumption	MW	117.4	121.0	121.2	2.41	117.6	125.9
Fuel Consumption	lb/kWh	0.68	0.69	0.69	0.01	0.68	0.70
Sulfur Byproduct Production	lb/MWh	24.88	24.95	24.94	0.10	24.77	25.10
Plant Emissions							
SO2 Emissions	lb/10^6 BTU	0.221	0.224	0.224	0.013	0.205	0.245
NOX Emissions	lb/10^6 BTU	0.129	0.140	0.141	0.041	0.078	0.205
CO2 Emissions	lb/kWh	1.70	1.71	1.71	0.01	1.70	1.71
Plant Costs							
Total Capital Cost	$/kW	1732	1748	1756	80.7	1619	1892
Total Direct Cost	$/kW	815	822	822	5.1	815	832
Total Plant Cost	$/kW	1419	1433	1439	67.4	1324	1553
Total Plant Investment	$/kW	1647	1663	1670	78.2	1536	1802
Fixed Operating Cost	$/kW-year	50.35	51.94	52.27	3.31	47.19	58.15
Variable Operating Cost	mills/kWh	10.59	11.27	11.20	0.54	10.28	12.16
Fuel Cost	mills/kWh	10.91	11.09	11.07	0.49	10.19	11.89
Byproduct Credit	mills/kWh	1.52	1.29	1.26	0.19	0.89	1.50
All Others	mills/kWh	1.20	1.38	1.39	0.13	1.22	1.65
Cost of Electricity	mills/kWh	50.88	52.00	52.27	3.88	49.38	55.58

Table 9.6 Selected Outputs Collected by the Model for Uncertainty Analysis

Parameter	Units	Best Guess"”	$f_{.50}$	μ	σ	$f_{.05}$	$f_{.95}$
Plant Performance							
Plant Thermal Efficiency	Fraction	35.03	34.46	34.40	0.50	33.51	35.14
Net Plant Output	MW	793.0	793.9	793.8	2.14	789.9	797.1
Gross Power Output	MW	908.6	913.5	913.3	3.86	906.7	920.1
Total Auxiliary Power Consumption	MW	115.6	119.3	119.5	2.54	115.5	124.1
Fuel Consumption	lb/kWh	0.76	0.78	0.78	0.01	0.76	0.80
Sulfur Byproduct Production	lb/MWh	27.49	27.71	27.70	0.16	27.43	27.96
Plant Emissions							
SO_2 Emissions	lb/10^6 BTU	0.219	0.222	0.222	0.012	0.203	0.243
NO_x Emissions	lb/10^6 BTU	0.120	0.130	0.130	0.038	0.072	0.190
CO_2 Emissions	lb/kWh	1.91	1.93	1.93	0.01	1.91	1.95
Plant Costs							
Total Capital Cost	$/kW	1540	1549	1557	67.8	1447	1677
Total Direct Cost	$/kW	729	736	736	5.4	729	746
Total Plant Cost	$/kW	1256	1263	1269	56.6	1178	1370
Total Plant Investment	$/kW	1457	1465	1473	65.6	1367	1589
Fixed Operating Cost	$/kW-year	42.55	44.06	44.20	1.91	41.35	47.89
Variable Operating Cost	mills/kWh	12.23	13.04	12.98	0.62	11.97	14.10
Fuel Cost	mills/kWh	12.28	12.54	12.51	0.56	11.49	13.45
Byproduct Credit	Mills/kWh	1.68	1.43	1.40	0.21	1.00	1.67
All Others	Mills/kWh	1.63	1.85	1.87	0.15	1.67	2.16
Cost of Electricity	Mills/kWh	47.67	48.71	49.01	1.64	46.49	51.85

The distributions for three important model outputs - total capital cost, levelized cost of electricity, and plant thermal efficiency are discussed in Section 9.5.

The regression analysis of results of Cases 1, 2 and 3 for the two models indicated that three performance parameters were significantly correlated with uncertainty in plant efficiency including carbon conversion, water to feedstock ratio, and gasifier temperature. For both the total capital cost and levelized cost of electricity, uncertainty in project cost contingency, engineering and home office fees, indirect construction costs, process contingency for gas turbine, and fuel cost were found to be influential from the analysis of Cases 1,2 and 3 for all the three models.

9.5 Analysis of Results

The performance and cost models of the the two models were run using the set of assumptions regarding uncertainties in process performance and costs shown in Tables 9.3 and 9.4. A deterministic simulation of each of the models was also run which is based on "best guess" values for the values for the parameters, which are treated as uncertain in the probabilistic simulation. The deterministic simulation is intended to be representative of the estimates for plant performance and cost that would be obtained in lieu of probabilistic simulation, frequency distributions for variables calculated in the performance and cost models can be estimated. The plant thermal efficiency, the total capital requirement, and the cost of electricity are the key outputs analyzed in the following sections for the two models.

9.5.1 Plant Thermal Efficiency

The distributions of the plant thermal efficiencies are collected for two models and results are discussed below. For each model, the plant thermal efficiency of the deterministic simulation and the frequency distributions of the plant thermal efficiency for Cases 2 and 5 are plotted on a graph and analyzed.

9.5.1.1 Coal-Fueled Texaco-based IGCC system with Radiant and Convective Design

The uncertainty in the plant thermal efficiency for the radiant and convective model is shown in Figure 9.2. The deterministic result of 39.41 percent is shown as a vertical dotted line in the graph. The probabilistic simulation (Case 2) indicates that the mean is 38.88 percent and the median (50th percentile) is 38.91 percent, both of which are less than the deterministic value. The key uncertain variables are the gasifier temperature, water-to-coal ratio, carbon conversion, and the 4^{rth} and 5th approach temperatures. The number of uncertain variables reduced from 15 to 5. Case 5 results in similar outputs for thermal efficiency. This indicates that there is little difference between the two cases. Thus, the uncertainties screened out of case studies need not be the subject of any further study and these screened model inputs were assigned point estimates. Table 9.7 indicates that the range of the efficiencies enclosed by the 90 percent confidence interval of the distribution is from 38.05 to 39.51 percent for Case 5. The probability distribution is negatively skewed. There is a 5 percent probability that the efficiency could be less than 38.00 percent and it may go as low as 37.5 percent. There is a 15 percent probability that the efficiency would be higher than the deterministic estimate of 39.41 percent and it could go as high as 39.75 percent. Therefore, if only a point estimate was used to predict the plant thermal efficiency, then the efficiency will be overestimated 85 percent of the time.

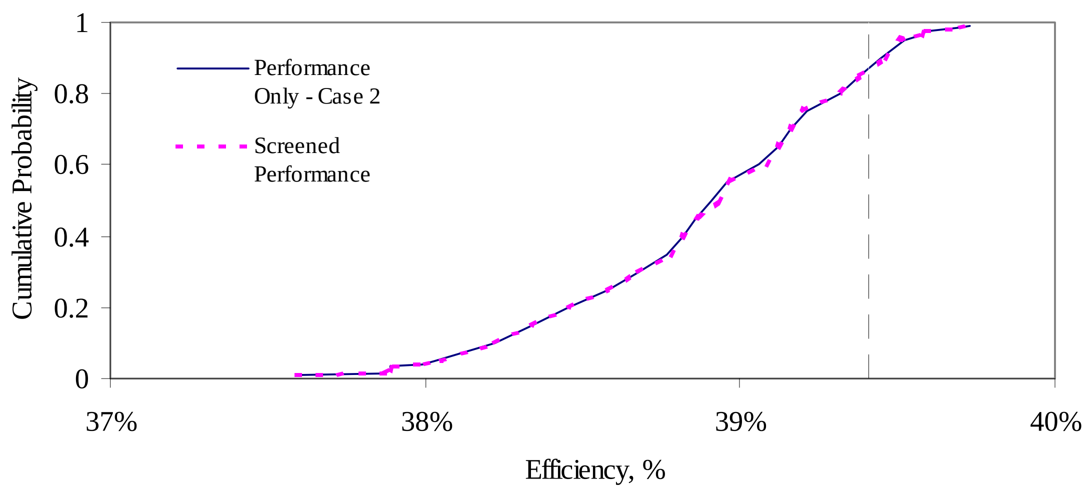

Figure 9.2 Comparison of Probabilistic Results for Cases 2 and 5 for thePlant Thermal Efficiency for Radiant and Convective Model

Table 9.7 Summary of Results from Deterministic and Probabilistic Simulations of IGCC with Original and Screened Sets of Uncertainties

Parameter	Units	Deterministic	$f_{.50}$	μ	σ	$f_{.05}$	$f_{.95}$
Plant Thermal Efficiency							
Base Case 1	%	39.41	38.91	38.88	00.46	38.03	39.52
Case 2	%	39.41	38.91	38.88	00.46	38.03	39.52
Case 5	%	39.41	38.91	38.88	00.45	38.05	39.51
Total Capital Requirement							
Base Case 1	MW	1732	1748	1756	81	1619	1892
Case 3	MW	1732	1734	1741	80	1605	1872
Case 6	MW	1732	1737	1741	79	1605	1871
Levelized Cost of Electricity							
Base Case 1	MW	50.88	52.00	52.27	1.97	49.38	55.58
Case 3	MW	50.88	51.47	51.78	1.93	48.81	54.71
Case 6	MW	50.88	50.37	51.78	1.86	48.73	54.58

9.5.1.2 Coal-Fueled Texaco-based IGCC system with Total Quench Design

The uncertainty in the plant thermal efficiency for the coal-fueled total quench model is shown in Figure 9.3. The number of uncertain variables reduced from 15 to 5 and they are the same as those for the radiant and convective model. The vertical dotted line in the graph represents the deterministic result of 35.03 percent. The probabilistic simulation (Case 2) indicates that the mean is 34.40 percent and the median (50th percentile) is 34.46 percent, both of which are less than the deterministic value. Table 9.8 indicates the range of the efficiencies enclosed by the 90 percent confidence interval of the distribution to be from 33.46 to 35.05 percent for Case 5. The probability distribution is negatively skewed similar to that in the radiant and convective model. If only the deterministic result of 35.03 percent is considered, then the plant thermal efficiency can be overestimated 90 percent of the time. The plant thermal efficiency may go as low as 32.5 percent and as high as 35.3 percent.

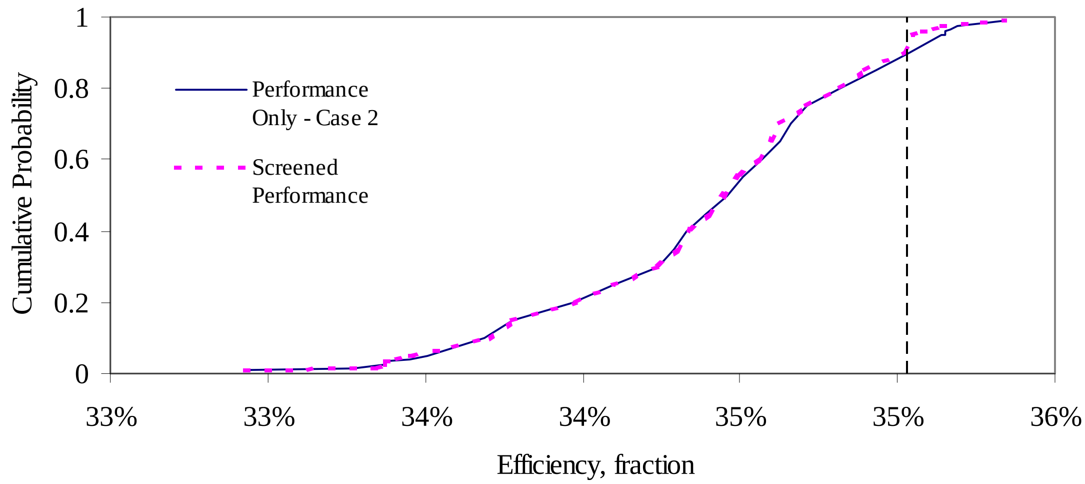

Figure 9.3 Comparison of Probabilistic Results for Cases 2 and 5 for thePlant Thermal Efficiency for Total Quench Coal Model

Table 9.8 Summary of Results from Deterministic and Probabilistic Simulations of IGCC with Original and Screened Sets of Uncertainties

Parameter	Units	Deterministic	$f_{.50}$	μ	σ	$f_{.05}$	$f_{.95}$
Plant Thermal Efficiency							
Base Case 1	%	35.03	34.46	34.40	00.49	33.51	35.14
Case 2	%	35.03	34.46	34.40	00.49	33.51	35.14
Case 5	%	35.03	34.45	34.39	00.48	33.46	35.05
Total Capital Requirement							
Base Case 1	MW	1540	1549	1557	68	1447	1677
Case 3	MW	1540	1533	1540	66	1432	1654
Case 6	MW	1540	1534	1540	66	1427	1649
Levelized Cost of Electricity							
Base Case 1	MW	47.67	48.71	49.00	1.64	46.88	51.85
Case 3	MW	47.67	48.06	48.41	1.58	46.07	50.95
Case 6	MW	47.67	48.00	48.41	1.57	46.19	50.89

9.5.2 Total Capital Cost

The deterministic and probabilistic results for the total capital cost for the two IGCC models are discussed in the following sections. The key uncertain variables for the total capital costs in each model are identified. The results are plotted on a graph in for each model.

9.5.2.1 Coal-Fueled Texaco-based IGCC system with Radiant and Convective Design

The uncertainty in the total capital cost is shown in Figures 9.4, 9.5, and 9.6. Figure 9.4 shows the CDF' s of total capital requirements of Case 1 and Case 4, Figure 9.5 for Case 2 and Case 5, and Figure 9.6 for Cases 3 and 6. The deterministic result of 1732 $/kW is shown as a vertical dotted line in the all the above figures. The above figures and Table 9.7 indicate that uncertainties in cost parameters have the strongest influence on the total capital cost distribution.

In Case 2 and Case 5 with uncertainties only in performance input variables, the total capital cost requirement distribution has a narrow 90% confidence interval from 1731 $/kW to 1767 $/kW as indicated in Figure 9.5. Therefore, the performance uncertainties have less influence on the costs of the model.

In Cases 3 and 6, with uncertainties only in cost input variables, the 90 percent confidence interval of total capital cost distribution is from 1605 $/kW to 1871 $/kW. The probabilistic simulation with all cost uncertain variables (Case 3) indicates that the mean is 1741 $/kW and the median (50th percentile) 1734 $/kW. The probabilistic simulation with key cost uncertain variables (Case 6) gives a mean of 1741 $/kW and median of 1737 $/kW. This indicates that there is little difference between the two cases. Thus, the uncertainties screened out of case studies need not be the subject of any further study and the screened model inputs were assigned point estimates. The total number of uncertainties in costs reduced from a total of 25 to 11. The key uncertain variables identified are engineering and home office fees, Indirect construction cost factor, project contingency, process contingency factors for oxidant feed, gasification, and gas turbine, maintenance cost factors Selexol, process condensate, and gas turbine, and units costs for fuel cost and ash disposal.

From the Figure 9.6 showing case 3 and 6, the probability that the total capital cost is greater than the deterministic value of 1732 $/kW is 50 percent. Therefore, if only deterministic results are considered, then the total capital cost is overestimated 50 percent of the time.

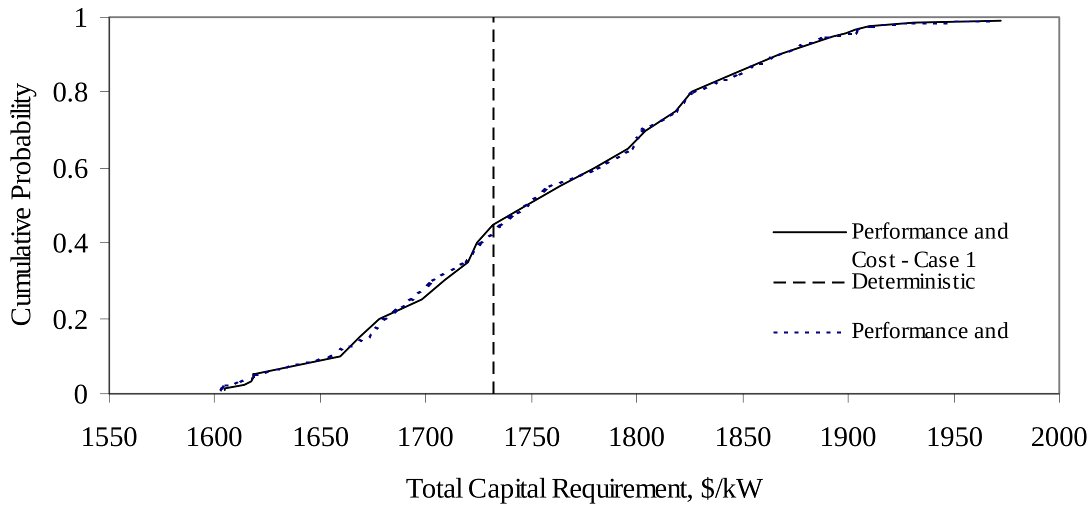

Figure 9.4 Comparison of Probabilistic Results for Cases 1 and 4 for theTotal Capital Requirement for Radiant and Convective Model

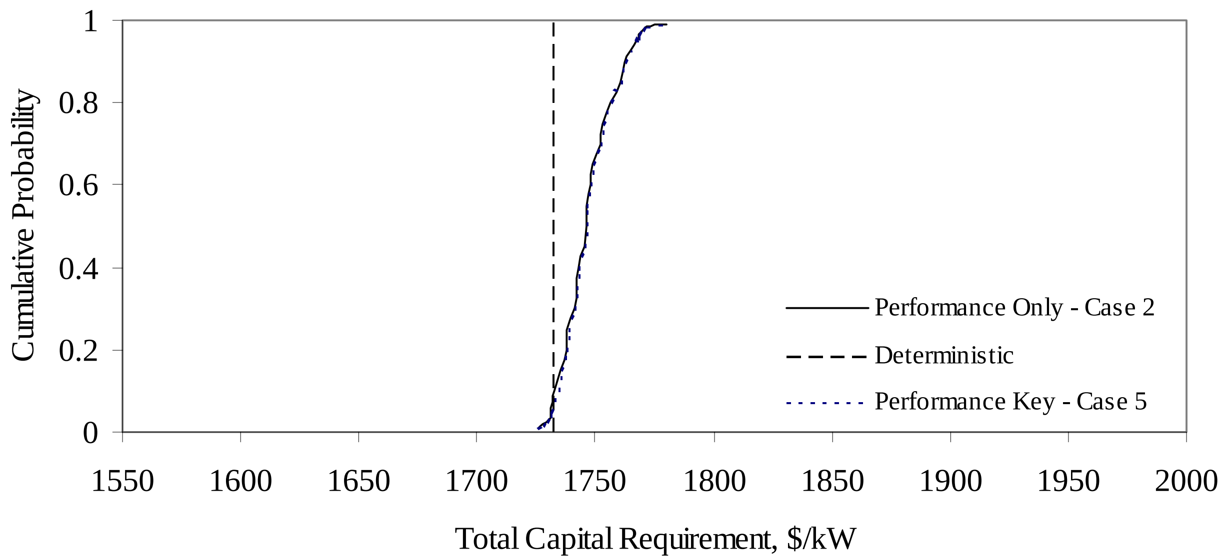

Figure 9.5 Comparison of Probabilistic Results for Cases 2 and 5 for theTotal Capital Requirement for Radiant and Convective Model

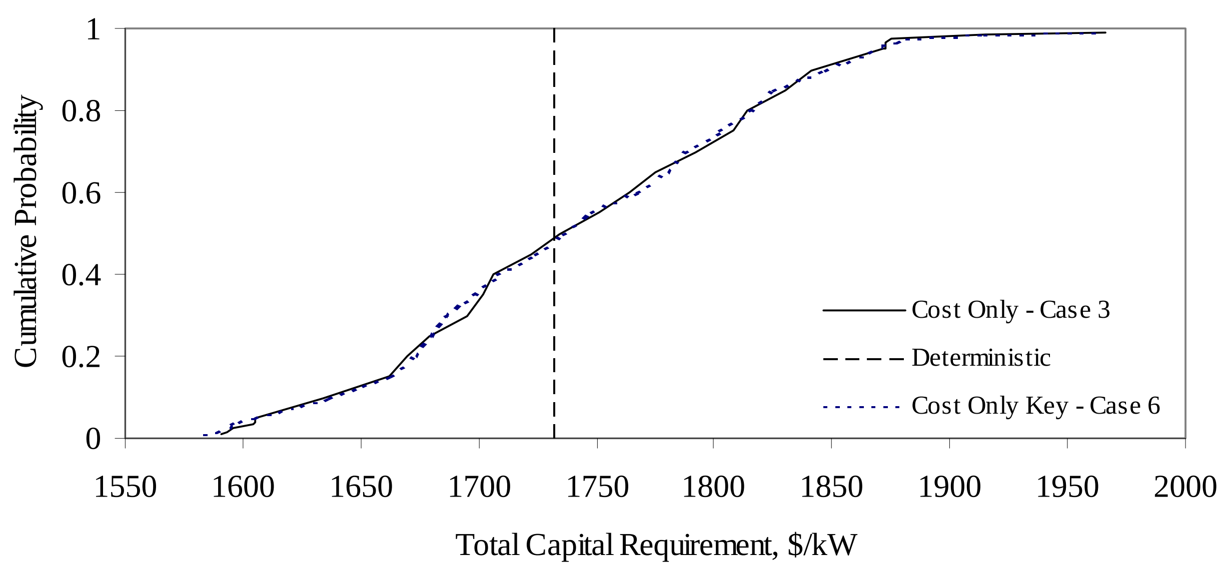

Figure 9.6 Comparison of Probabilistic Results for Cases 3 and 6 for theTotal Capital Requirement for Radiant and Convective Model

9.5.2.2 Coal-Fueled Texaco-based IGCC system with Total Quench Design

The uncertainty in the total capital cost is shown in Figures 9.7 9.8, and 9.9. Figure 9.7 shows the CDF' s of total capital requirements of Case 1 and Case 4, Figure 9.8 for Case 2 and Case 5 , and Figure 9.9 for Cases 3 and 6. The deterministic result of 1540 $/kW is shown as a vertical dotted line in the all the above figures. As in the radiant and convective model. the uncertainties in cost parameters have the strongest influence on the total capital cost distribution.

The total capital cost requirement distribution has a narrow 90% confidence interval from 1557 $/kW to 1576 $/kW, in Cases 2 and 5, indicating that the performance only uncertainties have small influence on the total capital cost distributions.

In Cases 3 and 6, with uncertainties only in cost input variables, the 90 percent confidence interval of total capital cost distribution is from 1427 $/kW to 1649 $/kW. The probabilistic simulation with all cost uncertain variables indicates that the mean is 1540 $/kW and the median (50th percentile) is 1533 $/kW, which are similar to those in the Case 6. The key uncertain cost inputs identified for this model are engineering and home office fees, Indirect construction cost factor, project contingency, process contingency factors for oxidant feed, gasification, and gas turbine and units costs for fuel cost. The total number of uncertain cost inputs reduced from 25 to 7. From the Figure 9.9 showing case 3 and 6, the probability of the total capital cost is greater than the deterministic value of 1540 $/kW is about 50 percent. Therefore, in the case of selecting only a point estimate model for predicting costs, then the total capital cost is overestimated 50 percent of the time.

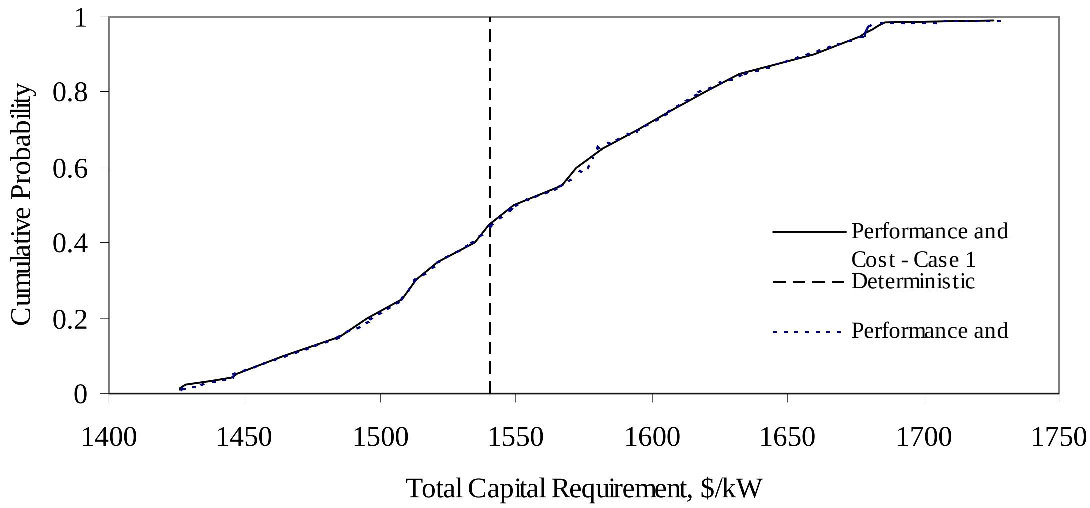

Figure 9.7 Comparison of Probabilistic Results for Cases 1 and 4 for theTotal Capital Requirement for Total Quench Coal Model

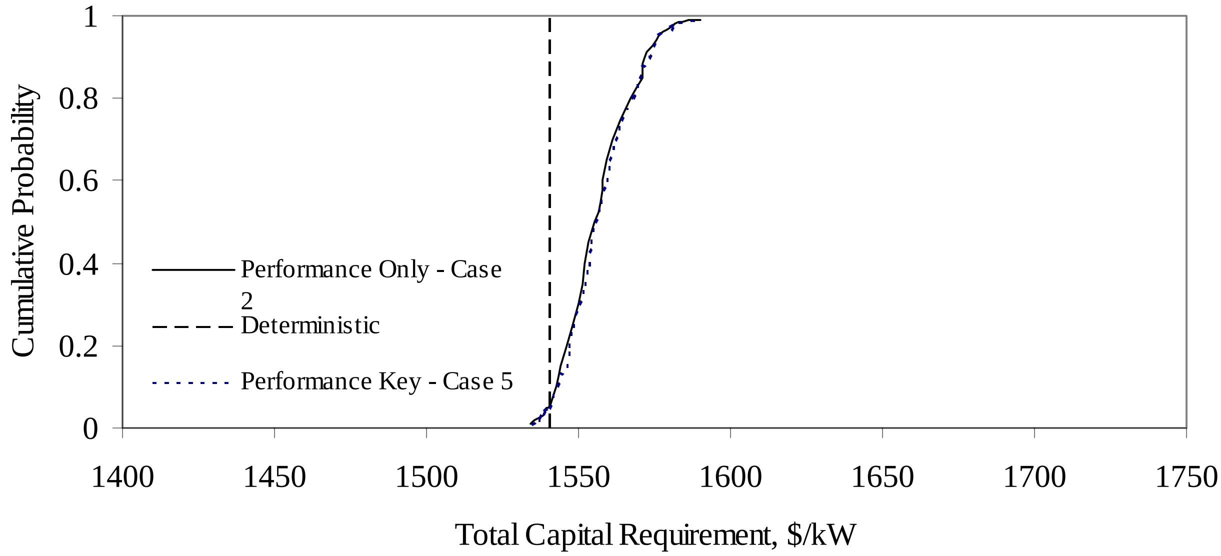

Figure 9.8 Comparison of Probabilistic Results for Cases 2 and 5 for theTotal Capital Requirement for Total Quench Coal Model

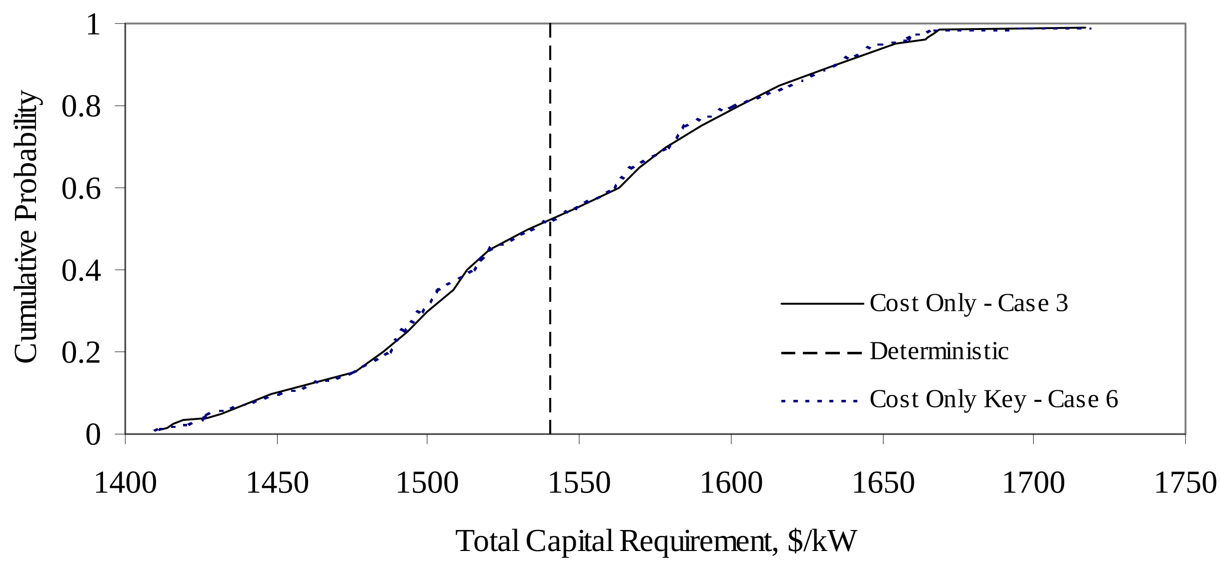

Figure 9.9 Comparison of Probabilistic Results for Cases 3 and 6 for theTotal Capital Requirement for Total Quench Coal Model

9.5.3 Cost of Electricity

The following sections discuss the results of probabilistic simulations for the two IGCC models for the cost of electricity. The deterministic results are analyzed with respect to the probabilistic results for each model. In the two models, the uncertainties in cost parameters have the dominating influence on the cost of electricity. Therefore, Cases 3 and 6, which contain uncertainties in only cost inputs, are discussed in detail. The key uncertainties in cost parameters influencing the cost of electricity are the same as those described for the total capital cost in each IGCC model.

9.5.3.1 Coal-Fueled Texaco-based IGCC system with Radiant and Convective Design

The uncertainty in the levelized cost of electricity is shown in Figures 9.10, 9.11, and 9.12. Figure 9.10 shows the CDF' s of cost of electricity of Case 1 and Case 4, Figure 9.11 for Case 2 and Case 5 , and Figure 9.12 for Cases 3 and 6. The deterministic result of 50.88 mills/kWh is shown as a vertical dotted line in the all the above figures.

In Cases 3 and 6, with uncertainties only in cost input variables, the 90 percent confidence interval of cost of electricity distribution is from 48.73 mills/kWh to 54.58 mills/kWh. The probabilistic simulation with all cost uncertain variables (Case 3) indicates that the mean is 51.78 mills/kWh and the median (50th percentile) 51.47 mills/kWh both of which are higher than that of the deterministic simulation of 51.88 mills/kWh. From the Figure 9.12 showing Case 3 and 6, the total capital cost is underestimated 65 percent of the time if only the point estimate simulation is used for predicting the cost of electricity.

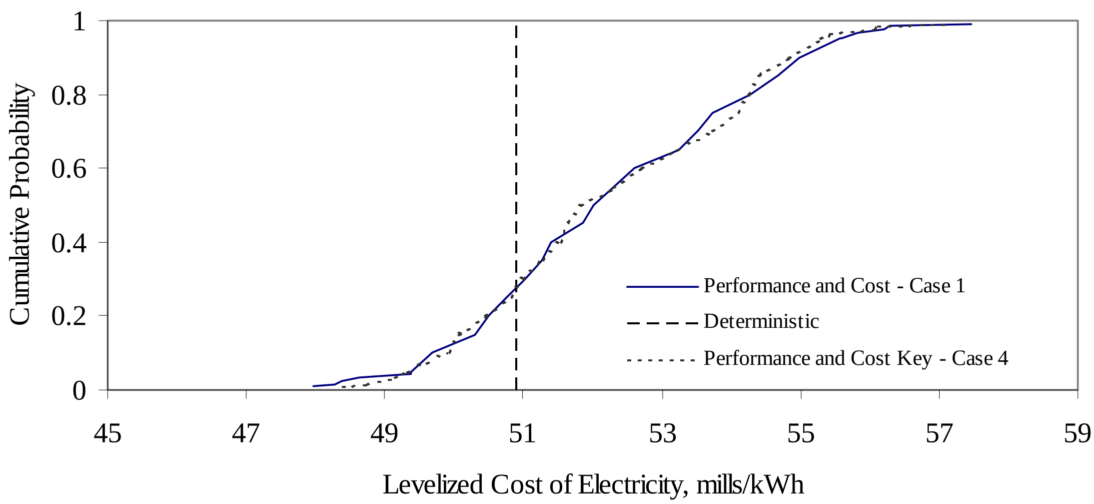

Figure 9.10 Comparison of Probabilistic Results for Cases 1 and 4 for the Levelized Cost of Electricity for Radiant and Convective Model

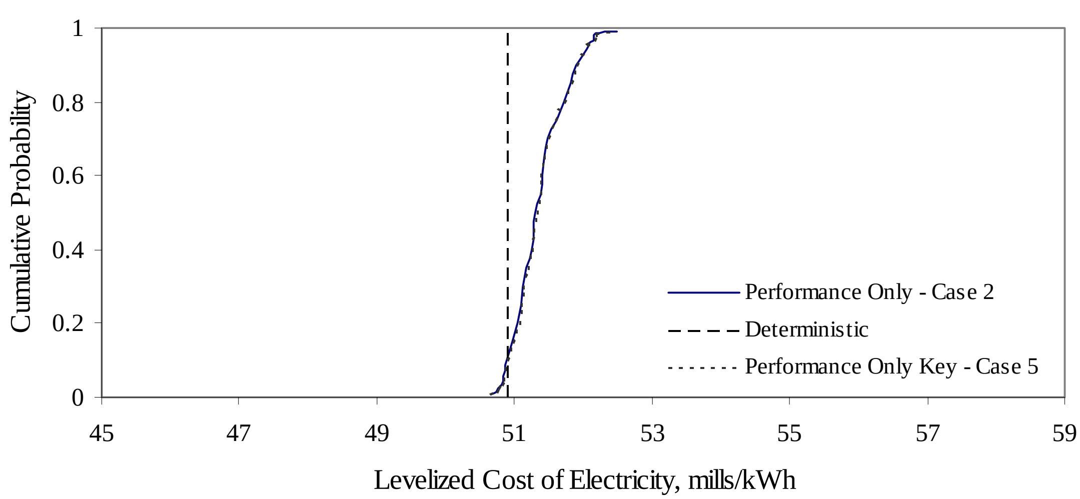

Figure 9.11 Comparison of Probabilistic Results for Cases 2 and 5 for the Levelized Cost of Electricity for Radiant and Convective Model

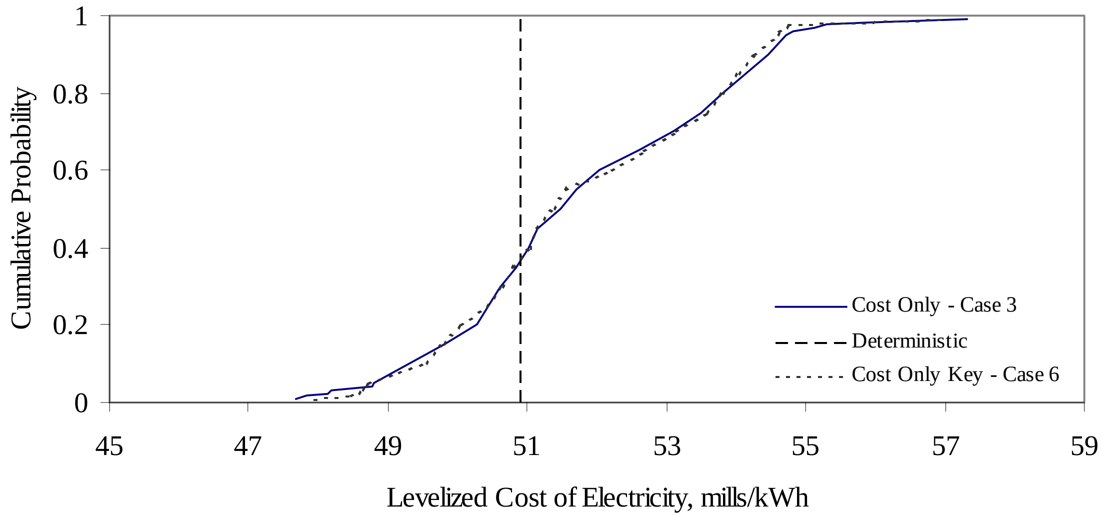

Figure 9.12 Comparison of Probabilistic Results for Cases 3 and 6 for the Levelized Cost of Electricity for Radiant and Convective Model

9.5.3.2 Coal-Fueled Texaco-based IGCC system with Total Quench Design

The uncertainty in the levelized cost of electricity is shown in Figures 9.13, 9.14, and 9.15. Figure 9.13 shows the CDF' s of cost of electricity of Case 1 and Case 4, Figure 9.14 for Case 2 and Case 5, and Figure 9.15 for Cases 3 and 6. The deterministic result of 47.67 mills/kWh is shown as a vertical dotted line in the all the above figures.

Cases 3 and 6 are cases with uncertainties only in cost input variables. In these cases the 90 percent confidence interval of cost of electricity distribution is from 46.19 mills/kWh to 50.89 mills/kWh. The probabilistic simulation with key cost uncertain variables (Case 6) gives a mean of 48.41 mills/kWh and median of 48.00 mills/kWh which is similar to those obtained from Case 3. Both the mean and the median are greater than the deterministic result of 47.67 mills/kWh. From the Figure 9.15 showing case 3 and 6, the probability that the total capital cost is greater than the deterministic value of 47.67 mills/kWh is 60 percent. Therefore, if only point estimates are used, then the cost of electricity is underestimated 60 percent of the time.

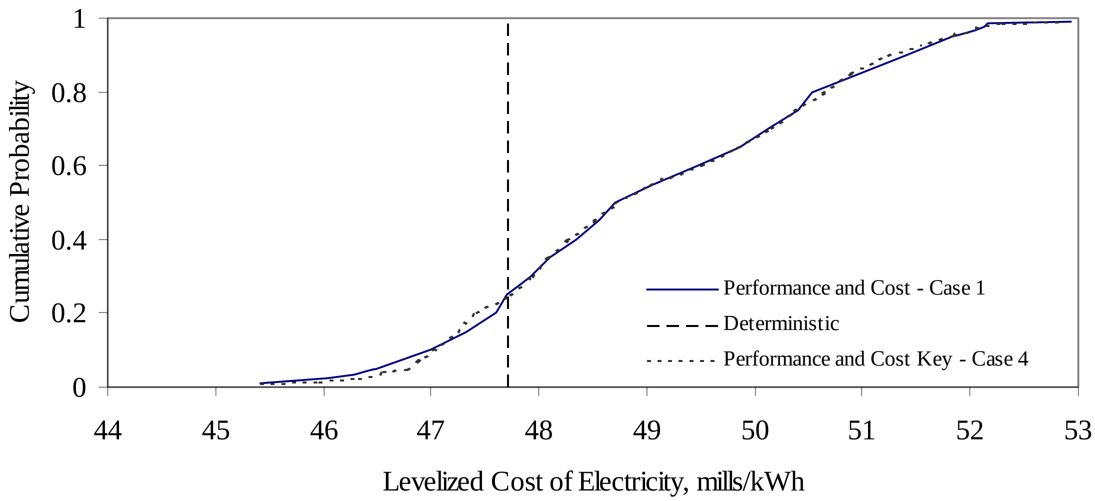

Figure 9.13 Comparison of Probabilistic Results for Cases 1 and 4 for the Levelized Cost of Electricity for Total Quench Coal Model

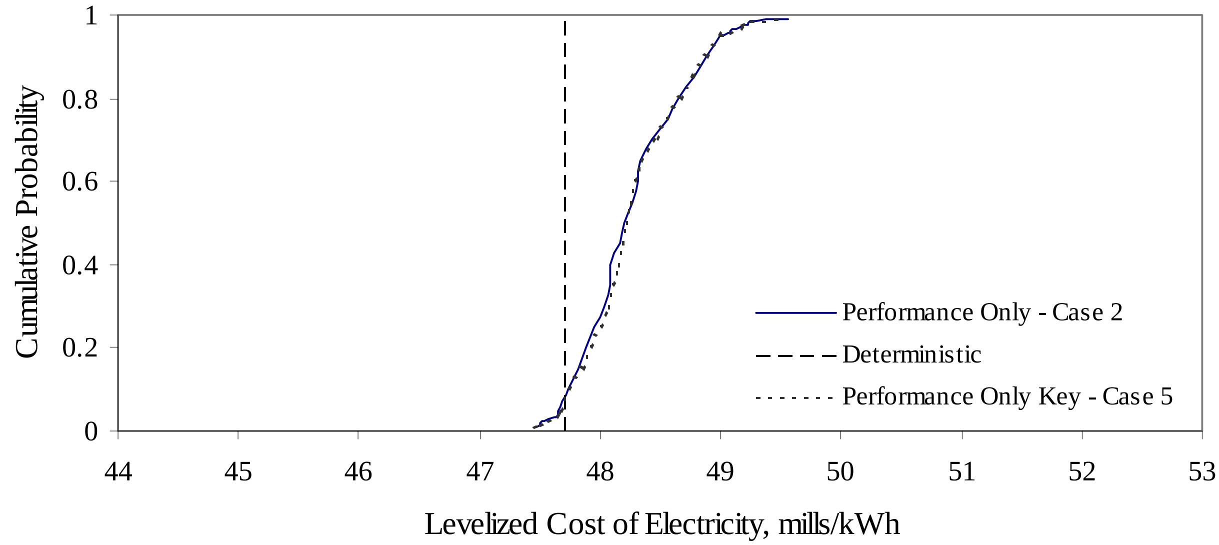

Figure 9.14 Comparison of Probabilistic Results for Cases 2 and 5 for the Levelized Cost of Electricity for Total Quench Coal Model

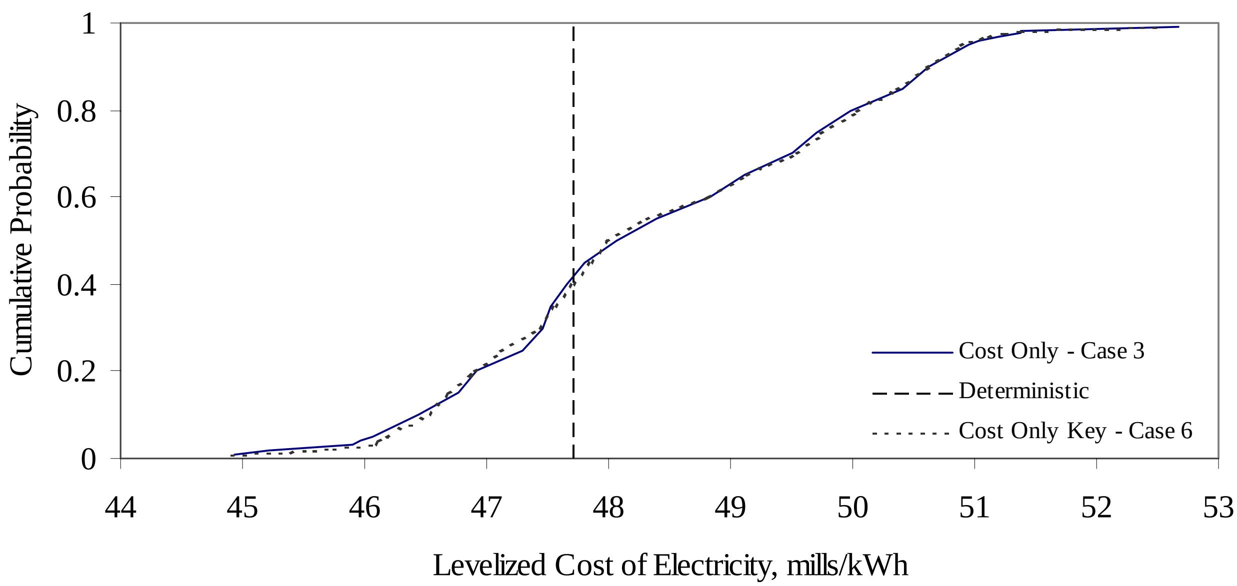

Figure 9.15 Comparison of Probabilistic Results for Cases 3 and 6 for the Levelized Cost of Electricity for Total Quench Coal Model

9.6 Discussion

The probabilistic analysis indicated that the deterministic case study overestimates the plant thermal efficiency 85 percent of the time in both the IGCC systems. The deterministic case study underestimates the cost of electricity 60 to 65 percent of the time.

The probabilistic analysis indicated that the range of the plant thermal efficiency of the radiant and convective coal-fueled model (38.0 - 39.5 percent) is higher than that of the total quench coal-fueled model (33.5 - 35.1 percent). However, the range of cost of electricity of radiant and convective coal-fueled model (45.4 - 55.6 mills/kWh) and that of total quench coal-fueled model (46.5 - 51.9 mills/kWh) are similar.

This demonstrates that the plant efficiency of a radiant and convective-based IGCC system is always higher than that of the coal-fueled total quench-based IGCC system. However, the cost of electricity of both the coal-fueled systems can be comparable and further analysis can be done to explore the possibilities of reducing the costs of the radiant and convective-based system.

10.0 CONCLUSIONS

This study documents the development of two Texaco gasifier-based models of Integrated Gasification Combined Cycle (IGCC) systems: (1) coal-fueled with radiant and convective high temperature gas cooling; and (2) coal-fueled with total quench high temperature gas cooling; using ASPEN.

In the first case, a new performance, emissions, and cost model was developed based upon refinements and modifications to a performance model previously developed by DOE/FETC. The new model incorporates more performance details regarding key process areas, such as the gas turbine and gasifier. New comprehensive capital, annual, and levelized cost models have been developed in this study. In addition, the new model includes additional features regarding flowsheet calculation sequencing and convergence schemes, as illustrated by the addition of a number of key design specifications to enhance the scope of important design assumptions and constraints. The new gas turbine performance model was calibrated to published data for operation on natural gas and also to data for operation on syngas. The other IGCC systems developed also contain the features included for the first model such as the new gas turbine model. The models are based upon properly sized gas turbines and deal with interactions among all process areas in order to properly capture differences due to fuel type and gas cooling design thereby facilitating the comparison of the different models..

The models of the Texaco-gasifier based IGCC systems are primarily based on the findings of a study sponsored by the Electric Power Research Institute (EPRI) (Matchak *et al*., 1984). The EPRI study provided extensive process designs which were modified as deemed appropriate for the development of the current models. The performance model for radiant and convective high temperature gas cooling design was adopted and modified from a previous model developed by K.R. Stone in 1985 for Federal Energy Technology Center (FETC). The performance model for the total quench high temperature gas cooling model is newly developed in the present study.

Cost models for each of the IGCC system models were developed using guidelines as suggested by the EPRI Technical Assessment Guide (EPRI, 1986), to estimate the capital, annual, and levelized costs of the IGCC systems. The cost models were developed as FORTRAN subroutines which are called by the performance models of the respective IGCC systems. The inputs to the cost models are provided by the ASPEN performance models in the form of values for key system variables, such as flowrates.

An example case study in each of these cases illustrates the type of results that may be obtained from the model regarding plant performance, emissions, and cost. The results indicate that of the two IGCC system models using coal as fuel to the gasifier, the IGCC system using the radiant and convective high temperature gas cooling has a higher plant efficiency of 39.4 percent, and higher cost of electricity of 50.88 mills/kWh than the

IGCC system using the total quench high temperature gas cooling design, which has a plant efficiency of 35.0 percent and cost of electricity of 47.67 mills/kWh.

The radiant and convective model has higher plant efficiency than that of the total quench coal-fueled model due to the additional steam generation in the gasifier process which in turn results in more power generation from the steam turbines. However, the radiant and convective coolers are expensive units to maintain resulting in higher costs in the case of the radiant and convective model than those of the total quench coal-fueled model.

IGCC is in the early stages of development. It has been demonstrated for only first-of-a-kind applications and there is not much history regarding the performance of these systems. Therefore, there are inherent uncertainties in the performance and cost parameter estimates. Therefore incorporation of uncertainties is critical for the design and evaluation of the IGCC systems.

The efficiency, total capital cost, and cost of electricity of an IGCC system depends on the values of key design and performance variables and cost parameter assumptions. The new IGCC systems developed in the present study were applied extensively in probabilistic case studies to evaluate the response of the models to changes in these parameters and to identify key uncertainties. The uncertainties in the IGCC systems were characterized by a systematic approach. In each of the three IGCC systems, a probabilistic model was developed to account for uncertainties in the performance parameters and cost parameters. The stochastic capability of ASPEN was used to perform the probabilistic analysis. The results from the probabilistic model simulations included possible ranges of values for the performance, emissions, and cost variables of the IGCC system. These results were used to identify the key uncertainties.

The total uncertain input variables initially assumed were 40. The total number of key uncertainty variables were atmost 16 in any of the three cases. This reduction in the number of uncertainties reduces the costs of conducting research in the less uncertain process areas. The key uncertain performance input variables include the gasifier temperature, the carbon conversion and the water-to-fuel ratio. The uncertainties in these parameters largely influence the plant thermal efficiency and net plant output. This indicates that significant research has to be done in the gasifier process area to reduce risks or poor plant efficiencies. Uncertainties in the engineering and home office fees, the project contingency, the indirect construction factor, and the fuel cost largely influence the capital, annual, and levelized cost of all the three IGCC systems.

The probabilistic analysis indicated that the deterministic case study overestimates the plant thermal efficiency 85 percent of the time in both the IGCC systems. The deterministic case study underestimates the cost of electricity 60 to 65 percent of the time.

The probabilistic analysis indicated that the range of the plant thermal efficiency of the radiant and convective coal-fueled model (38.0 - 39.5 percent) is higher than that of the total quench coal-fueled model (33.5 - 35.1 percent). However, the range of cost of electricity of radiant and convective coal-fueled model (45.4 - 55.6 mills/kWh) and that of total quench coal-fueled model (46.5 - 51.9 mills/kWh) are similar.

This demonstrates that the plant efficiency of a radiant and convective-based IGCC system is always higher than that of the coal-fueled total quench-based IGCC system. However, the cost of electricity of both the coal-fueled systems can be comparable and further analysis can be done to explore the possibilities of reducing the costs of the radiant and convective-based system.

The radiant and convective model has high costs compared to conventional power plants. However, it might be competitive in terms of high plant efficiencies and low emissions. The coal-fueled total quench model has higher costs and lower plant efficiencies than the conventional power plants. Therefore, it may not have any competitive edge in the United States except under stringent NO_x and SO_2 regulations.

The probabilistic analysis can be used to identify key process areas which have potential for further research, to possibly optimally configure the IGCC systems, and also to compare more comprehensively the trade-offs between the three technologies. The uncertainties in the costs can be reduced by a detailed cost estimate study. The models will be used in future work as a benchmark for comparison with more advanced and technologically-risky power generation system concepts.

11.0 REFERENCES

Agarwal, P, and H.C. Frey (1995), "Development and Application of Performance and Cost Models for the Externally-Fired Combined Cycle, " Task 1 Topical Report, Volume 2, DOE/MC/29094--5246-Vol.2, Prepared by North Carolina State University for Carnegie Mellon University and U.S. Department of Energy, Morgantown, West Virginia, July.

Agarwal, P. and H.C. Frey (1996), "Modeling and Assessment of the Externally-Fired Combined Cycle System," Prepared by North Carolina State University for U.S. Department of Energy, Morgantown, WV.

Anand, A.K., F.C. Jahnke, and R.R. Olson, Jr (1992), "High Efficiency Quench Gasification Combined Cycles with Integrated Air Separation," *Proceedings of Eleventh EPRI Conference on Gasification Power Plants,* Electric Power Research Institute, Inc., Palo Alto, CA, October.

Bharvirkar, R. and H.C. Frey (1998), "Development of Simplified Performance and Cost Models of Integrated Gasification Combined Cycle Systems, " Prepared by North Carolina State University for Carnegie Mellon University, Pittsburgh, Pennsylvania, July.

Bjorge, R and M. Jandrisevits (1996), “IGCC Technology for the 21st Century,” 1996 *Gasification Technologies Conference*, Electric Power Research Institute, Inc., Palo Alto, CA, October.

Buchanan, T. L., M.R. DeLallo, and J.S. White (1998), “Economic Evaluation of Advanced Coal Gasification Technologies for Power Generation,” 1998 *Gasification Technologies Conference*, Electric Power Research Institute, Inc., Palo Alto, CA, October.

DeMoss, T.B. (1995), “Gas, Coal Duke it out Over Repowering,” *Power Engineering*, 21-24, June.

Doering, E.L., and U. Mahagaokar (1992), “Benefits of Heat Recovery vs Water Quench in Coal or Petroleum Coke Gasification for Power Generation,” *Proceedings of Eleventh EPRI Conference on Gasification Power Plants*, Electric Power Research Institute, Inc., Palo Alto, CA, October.

EPRI, (1986), “TAG™ - Technical Assessment Guide, Volume1: Electricity Supply - 1986,” P-4463-SR, Electric Power Research Institute, Inc., Palo Alto, CA, December.

Eustis, F. H. and M.S. Johnson (1990), “Gas Turbine Effects on Integrated-Gasification-Combined-Cycle Power Plant Operations,” GS/ER-6770, Prepared by Stanford University for Electric Power Research Institute, Inc, Palo Alto, CA, March.

Farina, G.L., L. Bressan, and D. Todd (1998), “IGCC Capital Cost and Performance, ” 1998 *Gasification Technologies Conference,* Electric Power Research Institute, Inc., San Francisco, CA, October.

Farmer, R (1997); *Gas Turbine World*; Pequot Publishing Inc., Fairfield, CT; Vol 18., p 44.

Frey, H.C. and E.S. Rubin (1990), “Stochastic Modeling of Coal Gasification Combined Cycle Systems: Cost Models for Selected IGCC Systems,” Report No. DOE/MC/24248-2901 (NTIS No. DE90015345). June. Prepared by Carnegie Mellon University for U.S. Department of Energy, Morgantown, WV.

Frey, H.C. and E.S. Rubin (1991), “Development and Application of a Probabilistic Evaluation Method for Advanced Process Technologies,” DOE/MC 24248-3105 (DE91002095), Prepared by Carnegie Mellon University for U.S. Department of Energy, Morgantown, WV, April.

Frey, H.C., and E.S. Rubin (1992a), "Evaluate Uncertainties in Advanced Process Technologies," *Chemical Engineering Process*, p 63-70, May.

Frey, H.C., and E.S. Rubin (1992b), "Evaluation of Advanced Coal Gasification Combined-Cycle Systems under Uncertainty," *Ind. Eng. Chem. Res.*, 31: 1299-1307.

Frey, H.C., and E.S. Rubin (1992c), "Integration of Coal Utilization and Environmental Control in Integrated Gasification Combined Cycle Systems," *Environ. Sci. and Tech.*, 26(10):1982-1990.

Frey, H.C. (1994), "Development and Application of Performance and Cost Models for Gas Turbine-Based Selective Catalytic Reduction NOx Control," Task 1 Topical Report, Volume 1, Prepared by North Carolina State University for Carnegie Mellon University and U.S. Department of Energy, Morgantown, West Virginia, October.

Frey, H.C., E. Rubin, and U. Diwekar (1994), "Modeling Uncertainties in Advanced Technologies: Application to a Coal Gasification System with Hot Gas Cleanup," *Energy*, 19(4):449-463.

Frey, H.C., and R.B. Williams (1995), "Performance and Cost Models for the Direct Sulfur Recovery Process," Task 1 Topical Report, Volume 3, DOE/MC/29094--5246-Vol.3, Prepared by North Carolina State University for Carnegie Mellon University and U.S. Department of Energy, Morgantown, West Virginia, September.

Frey, H.C. (1998), "Quantitative Analysis of Variability and Uncertainty in Energy and Environmental Systems," Chapter 23 in *Uncertainty Modeling and Analysis in Civil Engineering*, B. M. Ayyub, ed., CRC Press: Boca Raton, FL.

Hager, R. L., and D. L. Heaven (1990), "Evaluation of a Dow-Based Gasification-Combined-Cycle Plant Using Bituminous Coal," GS-6904. Prepared by Flour Daniel, Inc for EPRI, Irvine, CA

Holt, N. (1998), " IGCC Power Plants - EPRI Design & Cost Studies," *Proceedings of EPRI/GTC Gasification Technologies Conference*, Electric Power Research Institute, Inc., San Francisco, CA, October.

Iman, R.L., and M.J. Shortencarier (1984), *A FORTRAN 7 Program and User's Guide for the Generation of Latin Hypercube and Random Samples for Use with Computer Models*, Report No. SAND83-2365, Sandia National Laboratories, Albuquerque, NM. March.

Iman, R.L., and M.J. Shortencarier and J. D. Johnson (1985), *A FORTRAN 7 Program and User's Guide for the Calculation of Partial Correlation and Standardized Regression Coefficients*, Report No. SAND85-0044, Sandia National Laboratories, Albuquerque, NM. June.

Kerkhof, F.P.J.M. and P.van Steerderen (1993), "Integration of Gas Turbine and Air Separation Unit for IGCC Power Plants," Netherlands Agency for Energy and the Environment, Amsterdam, Netherlands. August.

Matchak, T.A., A.D. Rao, V. Ramanathan et al. (1984), "Cost and Performance for Commercial Applications of Texaco-Based Gasification-Combined-Cycle Plants," AP-3486. Prepared by Flour Engineers, Inc for EPRI, Palo Alto, CA.

Merrow, E. W., Phillips, K. E., Myers, C. W. (1981), "Understanding Cost Growth and Performance Shortfalls in Pioneer Process Plants," Report No. R-2569-DOE, Rand Corporation, Santa Monica, CA, September.

McNamee, G.P., and G.A. White (1986), “Use of Lignite in Texaco Gasification-Based-Combined-Cycle Power Plants,” AP-4509. Prepared by Energy Conversion Systems, Inc for EPRI, Los Angeles, CA.

Mendez-Vigo, I., J. Pisa, J. Cortes et al. (1998), “The Puertollano IGCC plant: Status Update,” *1998 Gasification Technologies Conference*, Electric Power Research Institute, Inc., Palo Alto, CA, October.

Morgan, M.G., and M. Henrion (1990), *Uncertainty: A Guide to Dealing with Uncertainty in Quantitative Risk and Policy Analysis*, Cambridge University Press: New York.

Nowacki, P. (1981), *Coal Gasification Processes*, Noyes Data Corporation, Park Ridge, NJ, p 256.

Pietruszkiewicz, Milkavich, Booras et al. (1988), “An Evaluation of IGCC and PCFS plants, ” AP-5950. Prepared by Bechtel Group, Inc for EPRI, Palo Alto, CA.

Preston, W.E. (1996), "Texaco Gasification: Powering the 90's," 1996 *Gasification Technologies Conference*, Electric Power Research Institute, Inc., Palo Alto, CA, October.

Rocha, M. F., and H.C. Frey (1997), "Cost Modeling of a Texaco Coal Gasification Combined Cycle System, " Prepared by North Carolina State University for Carnegie Mellon University and U.S. Department of Energy, Morgantown, WV, August.

Sendin, U., W.Schellberg, W.Empsperger, et al. (1996), " The Pertollano IGCC Project, A 335 MW Demonstration Power Plant for the Electricity Companies in Europe," 1996 *Gasification Technologies Conference*, Electric Power Research Institute, Inc., Palo Alto, CA, October.

Simbeck D. R., R.L. Dickenson, and E.D. Oliver (1983), "Coal Gasification Systems: A Guide to Status, Applications, and Economics," AP-3109 Prepared by Synthetic Fuel Associates, Inc for Electric Power Research Institute, Palo Alto, California.

Simbeck, D. R. and R.L. Dickenson (1996), "The EPRI Coal Gasification Guidebook - A Review of the Recent Update," 1996 *Gasification Technologies Conference*, Electric Power Research Institute, Inc., Palo Alto, CA, October.

Smith, J., and D. Heaven (1992), "Evaluation of a 510-Mwe Destec GCC Power Plant Fueled With Illinois No. 6 Coal," TR-100319. Prepared by Flour Daniel, Inc for EPRI, Irvine, CA

Supp., E. (1990), How to Produce Methanol from Coal, Springer-Verlag, New York.

APPENDIX A: GLOSSARY OF ASPEN UNIT OPERATION BLOCKS AND BLOCK PARAMETERS

This appendix provides a summary of the ASPEN unit operation blocks and the associated block parameters. Table 1 lists the ASPEN unit operation block and a brief description of each block, and Table 2 lists the associated block parameters and a brief description of each of the parameters.

Table A.1 ASPEN Unit Operation Block Description

ASPEN MODEL NAME	DESCRIPTION
CLCHNG	This block is used to change the class of a stream. There must be only one inlet and outlet stream.
COMPR	The compressor block computes the work required for compression in a single-stage compressor or the work yielded by expansion in a single-stage turbine. The temperature, enthalpy, and phase condition of the outlet stream are also calculated. This block can simulate a centrifugal compressor, a positive displacement compressor, or an isoentropic turbine/compressor
DUPL	This block copies an inlet stream to any number of outlet streams. Material and energy balances are not satisfied by this block. All

	streams must be of the stream class.
FLASH2	This block determines the compositions and conditions of two outlet material streams (one vapor and one liquid) when any number of feed streams are mixed and flashed at specified conditions.
FSPLIT	The flow splitter block splits an inlet stream into one or more streams. All outlet streams have the same composition and intensive properties as the inlet stream. However, the extensive properties are a fraction of those of the inlet streams.
HEATER	This block calculates the physical equilibrium for a material stream at specified conditions and can be used to model heaters, coolers, valves, or pumps. There must be one material outlet stream for the block. The heat duty, if specified, may be supplied by an inlet information stream, or may be placed in an outlet information stream if calculated.
MIXER	This block simulates the mixing of two or more material and/or information streams. Every substream that appears in any outlet stream must be present in the inlet stream. The information stream must be class "HEAT". The user can specify the outlet pressure drop, the number of phases in the conventional substream, and the key phase.
PUMP	This block is used to raise the pressure of an inlet stream to a

	specified value and calculates the power requirement. Alternatively, PUMP will calculate the pressure of an outlet stream, given the inlet stream conditions and input work. This block can be used to model a centrifugal pump, a slurry pump, or a positive displacement pump.
RSTOIC	This stoichiometric block can be used to simulate a reactor when the stoichiometry is known, but the reaction kinetics are unknown or unimportant. The model may have any number of inlet material streams and one outlet material stream. This block can handle any number of reactions.
SEP2	This block simulates separation processes when the details of a separation process are not relevant or available. All streams must be of the same stream class. The first outlet is the top stream, and the second is the bottom stream.
SEP	This block separates an inlet stream into two or more outlet streams according to the split specified for each component. Two of the three properties, temperature, pressure, and vapor fraction, may be specified for each component.
RGIBBS	This block computes the phase and/or chemical equilibrium compositions at user-specified temperature and pressure when any number of feed streams are mixed. The output consists of up to one vapor phase, any number of liquid and solid phases. All

	materials must be of same class, and all information streams must be of the class "HEAT".
ABSBR	This block determines the overhead vapor and bottom liquid streams given at a set of inlet streams with specified inlet tray locations, number of stages and sidedraws. The model allows two to five material inlets and two to five material outlet. All material streams must be of the same stream class. All information streams must be of class "HEAT". The first material outlet is the top product stream. The second material outlet is the bottom product stream.

Table A.2 ASPEN Block Parameters Description

ASPEN Block Parameter	DESCRIPTION
ENT	Fraction of the liquid stream which is entrained in the vapor stream.
FRAC	It refers to the fraction of an inlet stream.
IDELT	It is a flag to indicate whether a temperature approach to chemical equilibrium is for an individual reaction (IDELT =1), or for the entire system (IDELT=0)
Isoentropic Efficiency	It refers to the isoentropic efficiency of s pump or compressor
MOLE-FLOW	It is used to specify the mole flow of a key in an outlet stream.
NAT	Number of atoms present in the system
NPHS	It is a flag to indicate whether a phase equilibrium calculation is desired. If 0, equilibrium phase distribution is determined. If 1, no phase equilibrium is calculated.
NPX	Maximum number of phases that may be present.
NR	Number of chemical reactions
NPK	Number of phases in the outlet stream for equilibrium calculations.
Q	Heat from the block. If 0 indicates that the block is adiabatic.
RFRAC	It is the fraction of the residue.
SYSOP3	Physical properties library containing values for vapor and liquid

	enthalpies and molar volumes and vapor-liquid K-values.
TYPE	It is the type of pump or compressor. For a pump type 1 refers to a centrifugal pump, type 2 refers to a slurry pump, and type 3 refers to a positive displacement pump. For a compressor type 1 refers to a centrifugal compressor, type 2 refers to a positive displacement compressor, and a type 3 refers to isoentropic turbine/compressor.
V	It refers to the vapor fraction in the outlet stream.

APPENDIX B: RESULTS OF THE PERFORMANCE AND COST MODELS OF THE IGCC SYSTEMS

COAL-FUELED TEXACO ENTRAINED FLOW IGCC POWER PLANT WITH RADIANT AND CONVECTIVE HIGH TEMPERATURE GAS COOLING: SYSTEM SUMMARY

```
                  ***  GASIFIER CONDITIONS  ***

             DRY COAL FLOW RATE:  0.584876E+06 LB/HR
               OXYGEN FLOW RATE:  0.539297E+06 LB/HR
                WATER FLOW RATE:  0.294777E+06 LB/HR
              GASIFIER PRESSURE:   615.0 PSIA
           GASIFIER TEMPERATURE:  2400.0 F


              ***  MS7000 GAS TURBINE CONDITIONS  ***

                FUEL FLOW RATE:  0.148088E+07 LB/HR
                 AIR FLOW RATE:  0.104475E+08 LB/HR
                      FUEL LHV:  3525.5 BTU/LB, 182.5 BTU/SCF
                      FUEL HHV:  3769.3 BTU/LB, 195.1 BTU/SCF
            FIRING TEMPERATURE:  2350.0 F
    COMBUSTOR EXIT TEMPERATURE:  2432.4 F
   TURBINE EXHAUST TEMPERATURE:  1113.4 F
      THERMAL EFFICIENCY (LHV):  0.3790
          GENERATOR EFFICIENCY:  0.9850


                 ***  STEAM TURBINE CONDITIONS  ***

   SUPERHEATED STEAM FLOW RATE:  0.225048E+07 LB/HR
 SUPERHEATED STEAM TEMPERATURE:   995.2 F
      REHEAT STEAM TEMPERATURE:   995.9 F
        EXPANDED STEAM QUALITY:  0.9539
          GENERATOR EFFICIENCY:  0.9850


                ***  POWER PRODUCTION SUMMARY  ***

               GAS TURBINE:    0.579898E+09  WATTS
             STEAM TURBINE:    0.401080E+09  WATTS
               COMPRESSORS:   -0.856458E+06  WATTS
                     PUMPS:   -0.559458E+07  WATTS
              OXYGEN PLANT:   -0.100250E+09  WATTS
               PLANT TOTAL:    0.874277E+09  WATTS

        ********************************************
        *  PLANT THERMAL EFFICIENCY (HHV) = 0.3993  *
        ********************************************
```

PERFORMANCE SUMMARY
Oxygen Blown Texaco-Based IGCC System with Cold Gas Cleanup

```
COST MODEL INPUT PERFORMANCE PARAMETERS  -------------------------------
                           Description       Value
                           -----------       -----
          Mass flow of coal to gasifier       584886. lb/hr
                    Ambient temperature           59. F
           Oxidant feedrate to gasifier      16708.63 lbmole/hr
                Oxygen flow to gasifier      15873.20 lbmole/hr
               Percent moisture in coal          0.00 percent
                    Percent ash in coal          0.10 percent
           Molar flow of syngas to LTGC      54318.40 lbmole/hr
              Syngas temperature in LTGC        101.00 F
                 Syngas pressure in LTGC        557.00 psia
               H2S entering Selexol unit        666.87 lbmole/hr
           Syngas entering Selexol unit      54318.40 lbmole/hr
          Molar flow of H2S out of Selx          7.07 lbmole/hr
         Mass flow of sulfur from Claus      20352.81 lb/hr
           Mass flow of sulfur from B/S       1119.88 lb/hr
                 Mass flow of raw water     492613.81 lb/hr
          Mass flow of polished water 1    2206982.41 lb/hr
          Mass flow of polished water 2      42579.02 lb/hr
          Mass flow of polished water 3      14986.12 lb/hr
         Mass flow of Scrubber Blowdown     174714.68 lb/hr
                    Gas Turbine Power 1 -0.446634E+09 Watts
                    Gas Turbine Power 2 -0.401891E+09 Watts
                    Gas Turbine Power 3 -0.317016E+09 Watts
               Gas Turbine Compressor 1  0.141826E+09 Watts
               Gas Turbine Compressor 2  0.187951E+09 Watts
               Gas Turbine Compressor 3  0.247430E+09 Watts
            Pressure of HP steam (HRSG)       1465.00 psia
           Mass flow of HP steam (HRSG)    2250518.15 lb/hr
                  Steam Turbine Power 1 -0.102356E+09 Watts
                  Steam Turbine Power 2 -0.940459E+08 Watts
                  Steam Turbine Power 3 -0.116760E+07 Watts
                  Steam Turbine Power 4 -0.209346E+09 Watts
                  Heating value of coal      12774.00 BTU/lb
                  Waste water flow rate     174714.68 lb/hr
             Steam Cycle Pump SLUR psia  0.326308E+06 Watts
             Steam Cycle Pump 1785 psia  0.514344E+07 Watts
              Steam Cycle Pump 565 psia  0.359577E+05 Watts
              Steam Cycle Pump 180 psia  0.296559E+05 Watts
               Steam Cycle Pump 65 psia  0.129703E+04 Watts
               Steam Cycle Pump 25 psia  0.541784E+05 Watts
                High pressure blowdown      69603.65 lb/hr
                 Claus boiler blowdown       1171.24 lb/hr
                 Low pressure blowdown       4223.90 lb/hr
                  CO2 from gas turbine      32037.41 lbmole/hr
                   CO from gas turbine          3.87 lbmole/hr
                  SO2 from gas turbine         25.80 lbmole/hr
                  COS from gas turbine          0.00 lbmole/hr
                   NO from gas turbine         19.97 lbmole/hr
                  NO2 from gas turbine          1.05 lbmole/hr
             CO2 from Beavon-Stretford       1316.29 lbmole/hr
           Actual heating value of coal      12782.65 BTU/lb
COST VARIABLE RESET - Variable PSNLTO value of        557.000
         in DCLT   reset to the upper limit of        435.000

COST VAR WARNING ---- Variable MHSH/N value of     750172.717
            in DCHR   above the upper limit of     640000.000
```

```
COST VAR WARNING ---- Variable MPW    value of    2264547.546
            in MSABFP above the upper limit of    2200000.000

COST VAR WARNING ---- Variable MCWI   value of       7816.940
            in M**CWI above the upper limit of       7700.000

                                 COST SUMMARY
           Oxygen Blown Texaco-Based IGCC System with Cold Gas Cleanup

A.  COST MODEL PARAMETERS  --------------------------------------------
            Plant Capacity Factor:   0.65       Cost Year:  January 1998
        General Facilities Factor:   0.17       Plant Cost Index:  388.0
            Indirect Construction:   0.20   Chemicals Cost Index:  446.8
                        Sales Tax:   0.05              Escalation:  0.00
           Engr & Home Office Fee:   0.10                Interest:  0.10
              Project Contingency:   0.17  Years of construction:     4
                 Number of Shifts:   4.25    Byproduct marketing:  0.10
              Fixed Charge Factor: 0.1034     Average Labor Rate:  19.70
Variable Levelization Cost Factor: 1.0000      Book Life (years):     30

B.  PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS  ------------------
                                                 Process     Maintenance
                           Plant Section     Contingency     Cost Factor
                           -------------     -----------     -----------
                           Coal Handling           0.050           0.030
                            Oxidant Feed           0.050           0.020
                            Gasification           0.150           0.045
             Low Temperature Gas Cooling           0.000           0.030
                                 Selexol           0.100           0.020
                             Claus Plant           0.050           0.020
                         Beavon-Stretford          0.100           0.020
              Boiler Feedwater Treatment           0.000           0.015
             Process Condensate Treatment          0.300           0.020
                             Gas Turbine           0.125           0.015
            Heat Recovery Steam Generator          0.025           0.015
                           Steam Turbine           0.025           0.015
                      General Facilities           0.050           0.015

C.  DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS ($1,000)  -------------
                               Number of Units       Direct      Process
                 Plant Section Operating Total Capital Cost  Contingency
                 -------------  --------- ----- ------------  -----------

                 Coal Handling          1     1       41963.        2867.
                  Oxidant Feed          2     2      108868.        7439.
                  Gasification          5     5      161082.       33018.
   Low Temperature Gas Cooling          2     2       16716.           0.
                       Selexol          2     2       26849.        3669.
                   Claus Plant          3     4       10055.         687.
              Beavon-Stretford          2     2        8720.        1192.
    Boiler Feedwater Treatment          1     1        5203.           0.
  Process Condensate Treatment          1     1        8495.        3483.
                   Gas Turbine          3     3      105969.       18101.
Heat Recovery Steam Generator           3     3       36995.        1264.
                 Steam Turbine          1     1       70214.        2399.
            General Facilities        N/A   N/A      102192.        6982.

D.  TOTAL CAPITAL REQUIREMENT ($1,000)  --------------------------------
```

Description	Annual Cost
Total Direct Cost	703322.
Indirect Construction Cost	140664.
Sales Tax	28836.
Engineering and Home Office Fees	87282.
Environmental Permitting	1000.
Total Indirect Costs	257783.
Total Process Contingencies	81100.
Project Contingency	182386.
Total Plant Cost	1224590.
AFDC	196241.
Total Plant Investment	1420831.
Preproduction (Startup) Costs	34522.
Inventory Capital	14564.
Initial Catalysts and Chemicals	7366.
Land	2745.
TOTAL CAPITAL REQUIREMENT ($1,000) -------->	1494237.

E. FIXED OPERATING COSTS ($/year) ------------------------------------

Description	Annual Cost
Operating Labor	7488364.
Maintenance Costs	30097578.
Administration and Supervision	5858219.
TOTAL FIXED OPERATING COST ($/year) -------------->	43444161.

F. VARIABLE OPERATING COSTS --

1. CONSUMABLES ($/year)

Description	Unit Cost	Material Requirement	Annual Operating Cost
Sulfuric Acid:	119.52 $/ton	1539.8 ton/yr	184039.
NaOH:	239.04 $/ton	317.7 ton/yr	75933.
Na2 HPO4:	0.76 $/lb	1230.7 lb/yr	936.
Hydrazine:	3.48 $/lb	5915.3 lb/yr	20567.
Morpholine:	1.41 $/lb	5492.2 lb/yr	7758.
Lime:	86.92 $/ton	709.7 ton/yr	61689.
Soda Ash:	173.85 $/ton	782.5 ton/yr	136032.
Corrosion Inh.:	2.06 $/lb	141601.1 lb/yr	292327.
Surfactant:	1.36 $/lb	141601.1 lb/yr	192321.
Chlorine:	271.64 $/ton	21.7 ton/yr	5886.
Biocide:	3.91 $/lb	24053.8 lb/yr	94088.
Selexol Solv.:	1.96 $/lb	55557.5 lb/yr	108659.
Claus Catalyst:	478.08 $/ton	12.7 ton/yr	6078.
Sul.. Acid Cat:	2.06 $/liter	0.0 liter/yr	0.
SCOT Catalyst:	249.91 $/ft3	0.0 ft3/yr	0.
SCOT Chemicals:	0.39 $/ft3	0.0 ft3/yr	0.
B/S Catalyst:	184.71 $/ft3	62.3 ft3/yr	11510.
B/S Chemicals:	N/A	N/A	134457.
Fuel Oil:	45.64 $/bbl	48949.5 bbl/yr	2233815.
Plant Air Ads.:	3.04 $/lb	3671.2 lb/yr	11169.
Raw Water:	0.79 $/Kgal	336233.6 Kgal/yr	266694.
Waste Water:	912.70 $/gpm ww	174714.7 lb/hr	207079.
LPG - Flare:	12.71 $/bbl	4283.1 bbl/yr	54449.
TOTAL CONSUMABLES ($/year) ---------------------->			4105485.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)

Coal:	1.26 $/MMBtu	584885.7 lb/hr	53619480.

```
        Ash Disposal:   10.87 $/ton        694.8 ton/day        1791195.
      Byprod. Credit:  135.82 $/ton         10.7 ton/hr      (
7472670.)

      TOTAL VARIABLE OPERATING COST  ($/year) ------------>    52043491.

G.  COST OF ELECTRICITY ----------------------------------------------
      Power Summary (MWe)                   Auxiliary Loads (MWe)
-----------------------------   ----------------------------------------
Gas Turbine Output     579.51   Coal Handling   7.30  Claus          0.43
Steam Turbine Output   400.81   Oxidant Feed   83.49  B/S            1.30
Total Auxiliary Loads  117.43   Gasification    1.16  Proc. Cond     0.59
-----------------------------   Low T Cool.     2.38  Steam Cycle    5.26
Net Electricity        862.89   Selexol         4.84  General Fac   10.68

                                      Capital Cost:    1731.66 $/kW
                              Fixed Operating Cost:      50.35 $/(kW-yr)
        Incremental Variable Costs:   1.20 mills/kWh
        Byproduct Credit:             1.52 mills/kWh
        Fuel Cost:                   10.91 mills/kWh
                           Variable Operating Cost:      10.59 mills/kWh
COST OF ELECTRICITY  --------------------------------->  50.88 mills/kWh

     Heat Rate is:  8664. BTU/kWh.  Efficiency is: 0.3941

H.  ENVIRONMENTAL SUMMARY ---------------------------------------------
               INPUTS:   Coal          0.678 lb/kWh
                         Water         0.571 lb/kWh
               OUTPUTS   Water         0.087 lb/kWh
                         Ash           0.067 lb/kWh
                         WstWater      0.202 lb/kWh
                         CO2           1.701 lb/kWh
                         CO            0.000 lb/kWh
                         SO2        0.220869 lb/MMBtu
                         NOx        0.129329 lb/MMBtu
                         COS        0.000000 lb/MMBtu
                         NH3
Summary of Output Variables for Stochastic Analysis
DIRECT CAPITAL COST: Coal Handling:       41963.16
                     Ox. Feed :          108867.72
                     Gasifiers:          161082.17
                     LowT Cool:           16715.73
                     Selexol  :           26849.19
                     Claus    :           10054.72
                     Beavon-St:            8720.04
                     Boiler FW:            5203.44
                     Proc Cond:            8495.16
                     Gas Turbi:          105968.71
                     Heat ReSG:           36995.42
                     Steam Tur:           70214.06
                     Gen Facil:          102192.02
 Total Direct Capital Cost               703321.53
 Cost of initial Catalyst and Chems      140664.31
 Cost of Taxes                            28836.18
 Engr and Home Office Fees                87282.20
 Indirect capital costs                  257782.69
 Project Contingency costs               182385.80
 Total plant cost                       1224590.37
 Allowance for funds during constr            1.16
```

Total process investment	1420830.97
Preproduction costs	34522.25
Initial chemicals	14564.45
TCICC	7366.21
Land costs	2744.53
Total capital requirements	1494236.72
New Power	862.89
OC Labor	7488364.00
OC Maintenance	30097578.11
OC Admin & Supervision	5858218.57
Fixed Operating Costs	43444160.69
Consumables Operating Cost	4105485.16
Byproduct Credit	579.51
Ash Disposal Cost	400.81
Variable Operating Cost	117.43
Fuel Cost	7472669.60
Capital Cost $/Kw	1791195.13
FOC, $/kW-yr	52043491.09
VOC, mills/kWh	53619480.39
Fuel Cost mills/kWh	1731.66
Byproduct Credit, mills/kWh	50.35
Incremental VOC, mills/kWh	10.59
Cost of Electricity, mills/kWh	10.91
Heat Rate	1.52
Efficiency	1.2001
CO2 Emissions	50.8802
CO Emissions	8664.3324
SO2 Emissions	0.3941
COS Emissions	1.7007
CH4 Emissions	0.0001
H2S Emissions	0.2209
NOx Emissions	0.0000
No of Op. Gasifiers	5
No of Total Gasifiers	5
No of Plant operators	43
Ash output	0.0671
Coal Inputs	0.6778
Water inputs	0.5709
Water outputs (blowdown)	0.0869
Sulfur outputs	0.0249
Fixed Charge Factor	0.1034
Variable Levelization Cost Factor	1.0000

Gasifier coal feed, lb/hr	584885.69
SO2 emissions, lbmole/hr	25.8016
COS emissions, lbmole/hr	0.0000
Percent Ash, fraction	0.0990
Percent Sulfur, %	3.8700
Coal sulfur inlet, lb/hr	22635.0761
Percent Sulfur Capture, fraction	0.9635

COAL-FUELED TEXACO ENTRAINED FLOW IGCC POWER PLANT WITH TOTAL QUENCH HIGH TEMPERATURE GAS COOLING: SYSTEM SUMMARY

```
                         ***  GASIFIER CONDITIONS  ***

                    DRY COAL FLOW RATE:  0.604729E+06 LB/HR
                      OXYGEN FLOW RATE:  0.557609E+06 LB/HR
                       WATER FLOW RATE:  0.304783E+06 LB/HR
                     GASIFIER PRESSURE:   615.0 PSIA
                  GASIFIER TEMPERATURE:  2400.0 F


                     ***  MS7000 GAS TURBINE CONDITIONS  ***

                        FUEL FLOW RATE:  0.183004E+07 LB/HR
                         AIR FLOW RATE:  0.999286E+07 LB/HR
                              FUEL LHV:  2948.7 BTU/LB, 150.4 BTU/SCF
                              FUEL HHV:  3152.6 BTU/LB, 160.8 BTU/SCF
                    FIRING TEMPERATURE:  2335.0 F
            COMBUSTOR EXIT TEMPERATURE:  2410.4 F
           TURBINE EXHAUST TEMPERATURE:  1123.7 F
              THERMAL EFFICIENCY (LHV):  0.3891
                  GENERATOR EFFICIENCY:  0.9850


                        ***  STEAM TURBINE CONDITIONS  ***

           SUPERHEATED STEAM FLOW RATE:  0.128138E+07 LB/HR
         SUPERHEATED STEAM TEMPERATURE:   992.9 F
              REHEAT STEAM TEMPERATURE:   993.1 F
                EXPANDED STEAM QUALITY:  0.9347
                  GENERATOR EFFICIENCY:  0.9850


                       ***  POWER PRODUCTION SUMMARY  ***

                      GAS TURBINE:    0.615370E+09  WATTS
                    STEAM TURBINE:    0.293852E+09  WATTS
                      COMPRESSORS:   -0.874967E+06  WATTS
                            PUMPS:   -0.620837E+07  WATTS
                     OXYGEN PLANT:   -0.103654E+09  WATTS
                      PLANT TOTAL:    0.798484E+09  WATTS

                 *********************************************
                 *  PLANT THERMAL EFFICIENCY (HHV) = 0.3527  *
                 *********************************************

                               PERFORMANCE SUMMARY
           Oxygen Blown Texaco-Based IGCC System with Cold Gas Cleanup

COST MODEL INPUT PERFORMANCE PARAMETERS  -------------------------------
                            Description       Value
                            -----------       -----
          Mass flow of coal to gasifier       604739. lb/hr
                    Ambient temperature           59. F
           Oxidant feedrate to gasifier      17275.98 lbmole/hr
               Oxygen flow to gasifier      16412.18 lbmole/hr
                Percent moisture in coal         0.00 percent
```

Percent ash in coal	0.10	percent
Molar flow of syngas to LTGC	56106.18	lbmole/hr
Syngas temperature in LTGC	101.00	F
Syngas pressure in LTGC	537.00	psia
H_2S entering Selexol unit	677.58	lbmole/hr
Syngas entering Selexol unit	56106.18	lbmole/hr
Molar flow of H_2S out of Selx	7.18	lbmole/hr
Mass flow of sulfur from Claus	20681.16	lb/hr
Mass flow of sulfur from B/S	1139.98	lb/hr
Mass flow of raw water	799465.69	lb/hr
Mass flow of polished water 1	1923748.71	lb/hr
Mass flow of polished water 2	173997.79	lb/hr
Mass flow of polished water 3	8002.93	lb/hr
Mass flow of Scrubber Blowdown	1379566.84	lb/hr
Gas Turbine Power 1	-0.450308E+09	Watts
Gas Turbine Power 2	-0.405647E+09	Watts
Gas Turbine Power 3	-0.320454E+09	Watts
Gas Turbine Compressor 1	0.135653E+09	Watts
Gas Turbine Compressor 2	0.179771E+09	Watts
Gas Turbine Compressor 3	0.236662E+09	Watts
Pressure of HP steam (HRSG)	1465.00	psia
Mass flow of HP steam (HRSG)	1281399.33	lb/hr
Steam Turbine Power 1	-0.581552E+08	Watts
Steam Turbine Power 2	-0.778951E+08	Watts
Steam Turbine Power 3	-0.649041E+07	Watts
Steam Turbine Power 4	-0.155587E+09	Watts
Heating value of coal	12774.00	BTU/lb
Waste water flow rate	1379566.84	lb/hr
Steam Cycle Pump SLUR psia	0.337384E+06	Watts
Steam Cycle Pump 1785 psia	0.439947E+07	Watts
Steam Cycle Pump 565 psia	0.363124E+05	Watts
Steam Cycle Pump 180 psia	0.753556E+05	Watts
Steam Cycle Pump 65 psia	0.131953E+04	Watts
Steam Cycle Pump 25 psia	0.475339E+05	Watts
Steam Cycle Pump 55 psia	0.705571E+04	Watts
High pressure blowdown	39630.91	lb/hr
Claus boiler blowdown	1187.74	lb/hr
Low pressure blowdown	12364.82	lb/hr
Intermed. pressure blowdown	8624.38	lb/hr
55 psia boiler blowdown	5628.89	lb/hr
CO_2 from gas turbine	33083.58	lbmole/hr
CO from gas turbine	4.00	lbmole/hr
SO_2 from gas turbine	26.43	lbmole/hr
COS from gas turbine	0.00	lbmole/hr
NO from gas turbine	19.10	lbmole/hr
NO_2 from gas turbine	1.01	lbmole/hr
CO_2 from Beavon-Stretford	1355.88	lbmole/hr
Actual heating value of coal	12782.65	BTU/lb

COST VAR WARNING ---- Variable MRW value of 799465.686
in DCBF above the upper limit of 614000.000

COST SUMMARY
Oxygen Blown Texaco-Based IGCC System with Cold Gas Cleanup

A. COST MODEL PARAMETERS --

Plant Capacity Factor:	0.65	Cost Year:	January 1998
General Facilities Factor:	0.17	Plant Cost Index:	388.0
Indirect Construction:	0.20	Chemicals Cost Index:	446.8
Sales Tax:	0.05	Escalation:	0.00

Engr & Home Office Fee:	0.10	Interest:	0.10
Project Contingency:	0.17	Years of construction:	4
Number of Shifts:	4.25	Byproduct marketing:	0.10
Fixed Charge Factor:	0.1034	Average Labor Rate:	19.70
Variable Levelization Cost Factor:	1.0000	Book Life (years):	30

B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Oxidant Feed	0.050	0.020
Gasification	0.150	0.045
Low Temperature Gas Cooling	0.000	0.030
Selexol	0.100	0.020
Claus Plant	0.050	0.020
Beavon-Stretford	0.100	0.020
Boiler Feedwater Treatment	0.000	0.015
Process Condensate Treatment	0.300	0.020
Gas Turbine	0.125	0.015
Heat Recovery Steam Generator	0.025	0.015
Steam Turbine	0.025	0.015
General Facilities	0.050	0.015

C. DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS ($1,000)

Plant Section	Number of Units Operating	Total	Direct Capital Cost	Process Contingency
Coal Handling	1	1	50626.	3460.
Oxidant Feed	2	2	112009.	7655.
Gasification	5	5	55192.	11316.
Low Temperature Gas Cooling	2	2	32793.	0.
Selexol	2	2	18120.	2477.
Claus Plant	3	4	10163.	695.
Beavon-Stretford	2	2	8820.	1206.
Boiler Feedwater Treatment	1	1	5849.	0.
Process Condensate Treatment	1	1	11219.	4600.
Gas Turbine	3	3	105969.	18105.
Heat Recovery Steam Generator	3	3	31527.	1077.
Steam Turbine	1	1	51442.	1758.
General Facilities	N/A	N/A	83934.	5736.

D. TOTAL CAPITAL REQUIREMENT ($1,000)

Description	Annual Cost
Total Direct Cost	577664.
Indirect Construction Cost	115533.
Sales Tax	23684.
Engineering and Home Office Fees	71688.
Environmental Permitting	1000.
Total Indirect Costs	211905.
Total Process Contingencies	58084.
Project Contingency	148339.
Total Plant Cost	995992.
AFDC	159608.
Total Plant Investment	1155600.
Preproduction (Startup) Costs	28658.
Inventory Capital	15383.

Initial Catalysts and Chemicals	6952.
Land	2860.
TOTAL CAPITAL REQUIREMENT ($1,000) -------->	1221009.

E. FIXED OPERATING COSTS ($/year) ----------------------------------

Description	Annual Cost
Operating Labor	7488364.
Maintenance Costs	21435941.
Administration and Supervision	4818822.
TOTAL FIXED OPERATING COST ($/year) -------------->	33743127.

F. VARIABLE OPERATING COSTS --

1. CONSUMABLES ($/year)

Description	Unit Cost	Material Requirement	Annual Operating Cost
Sulfuric Acid:	119.52 $/ton	1888.5 ton/yr	225708.
NaOH:	239.04 $/ton	390.4 ton/yr	93318.
Na2 HPO4:	0.76 $/lb	1950.7 lb/yr	1484.
Hydrazine:	3.48 $/lb	9385.8 lb/yr	32634.
Morpholine:	1.41 $/lb	8743.3 lb/yr	12350.
Lime:	86.92 $/ton	648.7 ton/yr	56391.
Soda Ash:	173.85 $/ton	716.8 ton/yr	124623.
Corrosion Inh.:	2.06 $/lb	129328.0 lb/yr	266990.
Surfactant:	1.36 $/lb	129328.0 lb/yr	175651.
Chlorine:	271.64 $/ton	20.1 ton/yr	5453.
Biocide:	3.91 $/lb	22298.0 lb/yr	87220.
Selexol Solv.:	1.96 $/lb	57393.5 lb/yr	112250.
Claus Catalyst:	478.08 $/ton	12.9 ton/yr	6176.
Sul.. Acid Cat:	2.06 $/liter	0.0 liter/yr	0.
SCOT Catalyst:	249.91 $/ft3	0.0 ft3/yr	0.
SCOT Chemicals:	0.39 $/ft3	0.0 ft3/yr	0.
B/S Catalyst:	184.71 $/ft3	63.4 ft3/yr	11716.
B/S Chemicals:	N/A	N/A	136871.
Fuel Oil:	45.64 $/bbl	44983.0 bbl/yr	2052802.
Plant Air Ads.:	3.04 $/lb	3373.7 lb/yr	10264.
Raw Water:	0.79 $/Kgal	545675.3 Kgal/yr	432819.
Waste Water:	912.70 $/gpm ww	1379566.8 lb/hr	1635123.
LPG - Flare:	12.71 $/bbl	3936.0 bbl/yr	50037.
TOTAL CONSUMABLES ($/year) ---------------------->			5529880.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)

Coal:	1.26 $/MMBtu	604739.0 lb/hr	55439532.
Ash Disposal:	10.87 $/ton	718.4 ton/day	1851995.
Byprod. Credit:	135.82 $/ton	10.9 ton/hr	(7593932.)

TOTAL VARIABLE OPERATING COST ($/year) ------------> 55227476.

G. COST OF ELECTRICITY ---

Power Summary (MWe)		Auxiliary Loads (MWe)			
Gas Turbine Output	614.96	Coal Handling	7.55	Claus	0.29
Steam Turbine Output	293.66	Oxidant Feed	86.33	B/S	1.32
Total Auxiliary Loads	115.64	Gasification	0.81	Proc. Cond	1.31
		Low T Cool.	1.80	Steam Cycle	4.57
Net Electricity	792.97	Selexol	1.16	General Fac	10.51

```
                                   Capital Cost:    1539.79 $/kW
                           Fixed Operating Cost:      42.55 $/(kW-yr)
        Incremental Variable Costs:   1.63 mills/kWh
        Byproduct Credit:             1.68 mills/kWh
        Fuel Cost:                   12.28 mills/kWh
                        Variable Operating Cost:      12.23 mills/kWh
COST OF ELECTRICITY  ---------------------------------> 47.67 mills/kWh

     Heat Rate is:  9748. BTU/kWh.  Efficiency is: 0.3503

H.  ENVIRONMENTAL SUMMARY  -------------------------------------------
               INPUTS:   Coal           0.763 lb/kWh
                         Water          1.008 lb/kWh
               OUTPUTS   Water          0.085 lb/kWh
                         Ash            0.075 lb/kWh
                         WstWater       1.740 lb/kWh
                         CO2            1.911 lb/kWh
                         CO             0.000 lb/kWh
                         SO2         0.218806 lb/MMBtu
                         NOx         0.119657 lb/MMBtu
                         COS         0.000000 lb/MMBtu
                         NH3
Summary of Output Variables for Stochastic Analysis
DIRECT CAPITAL COST: Coal Handling:        50625.89
                     Ox. Feed :           112009.44
                     Gasifiers:            55192.03
                     LowT Cool:            32792.63
                     Selexol  :            18120.28
                     Claus    :            10162.79
                     Beavon-St:             8819.95
                     Boiler FW:             5849.46
                     Proc Cond:            11219.07
                     Gas Turbi:           105968.71
                     Heat ReSG:            31527.18
                     Steam Tur:            51442.40
                     Gen Facil:            83934.07
 Total Direct Capital Cost                577663.91
 Cost of initial Catalyst and Chems       115532.78
 Cost of Taxes                             23684.22
 Engr and Home Office Fees                 71688.09
 Indirect capital costs                   211905.10
 Project Contingency costs                148339.30
 Total plant cost                         995992.47
 Allowance for funds during constr             1.16
 Total process investment                1155600.26
 Preproduction costs                       28657.66
 Initial chemicals                         15383.05
 TCICC                                      6952.34
 Land costs                                 2859.75
 Total capital requirements              1221009.08
 New Power                                   792.97
 OC Labor                                7488364.00
 OC Maintenance                         21435941.32
 OC Admin & Supervision                  4818822.16
 Fixed Operating Costs                  33743127.48
 Consumables Operating Cost              5529880.06
 Byproduct Credit                        7593931.52
 Ash Disposal Cost                       1851995.19
 Variable Operating Cost                55227475.87
```

Fuel Cost	55439532.14
Capital Cost $/Kw	1539.79
FOC, $/kW-yr	42.55
VOC, mills/kWh	12.23
Fuel Cost mills/kWh	12.28
Byproduct Credit, mills/kWh	1.68
Incremental VOC, mills/kWh	1.63
Cost of Electricity, mills/kWh	47.67
Heat Rate	9748.37
Efficiency	0.3503
CO2 Emissions	1.9110
CO Emissions	0.0001
SO2 Emissions	0.2188
COS Emissions	0.0000
CH4 Emissions	0.0000
H2S Emissions	0.0000
NOx Emissions	0.1197
No of Op. Gasifiers	5
No of Total Gasifiers	5
No of Plant operators	43
Ash output	0.0755
Coal Inputs	0.7626
Water inputs	1.0082
Water outputs (blowdown)	0.0850
Sulfur outputs	0.0275
Fixed Charge Factor	0.1034
Variable Levelization Cost Factor	1.0000

Gasifier coal feed, lb/hr	604738.96
SO2 emissions, lbmole/hr	26.4282
COS emissions, lbmole/hr	0.0000
Percent Ash, fraction	0.0990

HEAVY RESIDUAL OIL-FUELED TEXACO ENTRAINED FLOW IGCC POWER PLANT WITH TOTAL QUENCH HIGH TEMPERATURE GAS COOLING: SYSTEM SUMMARY

```
                 ***  GASIFIER CONDITIONS  ***

            DRY OIL FLOW RATE:  0.368072E+06 LB/HR
              OXYGEN FLOW RATE:  0.377643E+06 LB/HR
               WATER FLOW RATE:  0.169313E+06 LB/HR
             GASIFIER PRESSURE:   615.0 PSIA
          GASIFIER TEMPERATURE:  2400.0 F


             ***  MS7000 GAS TURBINE CONDITIONS  ***

               FUEL FLOW RATE:  0.134825E+07 LB/HR
                AIR FLOW RATE:  0.104519E+08 LB/HR
                     FUEL LHV:  3970.9 BTU/LB, 182.4 BTU/SCF
                     FUEL HHV:  4283.7 BTU/LB, 196.7 BTU/SCF
           FIRING TEMPERATURE:  2361.7 F
   COMBUSTOR EXIT TEMPERATURE:  2444.4 F
  TURBINE EXHAUST TEMPERATURE:  1122.8 F
     THERMAL EFFICIENCY (LHV):  0.3770
         GENERATOR EFFICIENCY:  0.9850


                ***  STEAM TURBINE CONDITIONS  ***

   SUPERHEATED STEAM FLOW RATE:  0.131168E+07 LB/HR
 SUPERHEATED STEAM TEMPERATURE:   993.4 F
      REHEAT STEAM TEMPERATURE:   993.6 F
       EXPANDED STEAM QUALITY:  0.9344
         GENERATOR EFFICIENCY:  0.9850


               ***  POWER PRODUCTION SUMMARY  ***

            GAS TURBINE:    0.591583E+09  WATTS
          STEAM TURBINE:    0.295136E+09  WATTS
            COMPRESSORS:   -0.310176E+06  WATTS
                  PUMPS:   -0.555983E+07  WATTS
           OXYGEN PLANT:   -0.702001E+08  WATTS
            PLANT TOTAL:    0.810649E+09  WATTS

       *********************************************
       *  PLANT THERMAL EFFICIENCY (HHV) = 0.3927  *
       *********************************************

                   PERFORMANCE SUMMARY
     Oxygen Blown Texaco-Based IGCC System with Cold Gas Cleanup

COST MODEL INPUT PERFORMANCE PARAMETERS  -------------------------------
                             Description       Value
                             -----------       -----
          Mass flow of oil to gasifier       368078. lb/hr
                   Ambient temperature           59. F
           Oxidant feedrate to gasifier      11700.22 lbmole/hr
               Oxygen flow to gasifier      11115.21 lbmole/hr
             Percent moisture in oil          0.00 percent
```

```
              Percent ash in oil          0.10 percent
         Molar flow of syngas to LTGC      48157.23 lbmole/hr
           Syngas temperature in LTGC        101.00 F
              Syngas pressure in LTGC        537.00 psia
            H2S entering Selexol unit        150.75 lbmole/hr
         Syngas entering Selexol unit      48157.23 lbmole/hr
        Molar flow of H2S out of Selx          1.60 lbmole/hr
       Mass flow of sulfur from Claus       3199.48 lb/hr
         Mass flow of sulfur from B/S       1642.91 lb/hr
               Mass flow of raw water     602394.98 lb/hr
        Mass flow of polished water 1    1926828.76 lb/hr
        Mass flow of polished water 2      56578.35 lb/hr
        Mass flow of polished water 3       2831.92 lb/hr
       Mass flow of Scrubber Blowdown     895040.48 lb/hr
                 Gas Turbine Power 1 -0.451592E+09 Watts
                 Gas Turbine Power 2 -0.406009E+09 Watts
                 Gas Turbine Power 3 -0.320037E+09 Watts
            Gas Turbine Compressor 1  0.141885E+09 Watts
            Gas Turbine Compressor 2  0.188030E+09 Watts
            Gas Turbine Compressor 3  0.247534E+09 Watts
          Pressure of HP steam (HRSG)       1465.00 psia
         Mass flow of HP steam (HRSG)    1311700.00 lb/hr
               Steam Turbine Power 1 -0.595578E+08 Watts
               Steam Turbine Power 2 -0.776550E+08 Watts
               Steam Turbine Power 3 -0.649604E+07 Watts
               Steam Turbine Power 4 -0.155721E+09 Watts
               Heating value of oil      19135.00 BTU/lb
               Waste water flow rate     895040.48 lb/hr
          Steam Cycle Pump SLUR psia  0.187424E+06 Watts
          Steam Cycle Pump 1785 psia  0.410285E+07 Watts
           Steam Cycle Pump 565 psia  0.497124E+04 Watts
           Steam Cycle Pump 180 psia  0.737179E+05 Watts
            Steam Cycle Pump 65 psia  0.147067E+04 Watts
            Steam Cycle Pump 25 psia  0.476062E+05 Watts
            Steam Cycle Pump 55 psia  0.240154E+04 Watts
             High pressure blowdown      40568.04 lb/hr
              Claus boiler blowdown         86.37 lb/hr
              Low pressure blowdown      12064.53 lb/hr
         Intermed. pressure blowdown      8528.62 lb/hr
            55 psia boiler blowdown       1837.43 lb/hr
               CO2 from gas turbine      26093.82 lbmole/hr
                CO from gas turbine          3.47 lbmole/hr
               SO2 from gas turbine          5.14 lbmole/hr
               COS from gas turbine          0.00 lbmole/hr
                NO from gas turbine         19.96 lbmole/hr
               NO2 from gas turbine          1.05 lbmole/hr
          CO2 from Beavon-Stretford        463.66 lbmole/hr
         Actual heating value of oil     19147.96 BTU/lb
COST VARIABLE RESET - Variable ETAO2  value of          0.950
         in DCOF   reset to the lower limit of          0.950

COST VAR WARNING ---- Variable MSBS/N value of       1642.913
            in DCBS   above the upper limit of       1200.000

COST VAR WARNING ---- Variable FH2S   value of          0.003
            in CHEMIS below the lower limit of          0.004

                                   COST SUMMARY
           Oxygen Blown Texaco-Based IGCC System with Cold Gas Cleanup
```

A. COST MODEL PARAMETERS

Plant Capacity Factor:	0.65	Cost Year:	January 1998
General Facilities Factor:	0.17	Plant Cost Index:	388.0
Indirect Construction:	0.20	Chemicals Cost Index:	446.8
Sales Tax:	0.05	Escalation:	0.00
Engr & Home Office Fee:	0.10	Interest:	0.10
Project Contingency:	0.17	Years of construction:	4
Number of Shifts:	4.25	Byproduct marketing:	0.10
Fixed Charge Factor:	0.1034	Average Labor Rate:	19.70
Variable Levelization Cost Factor:	1.0000	Book Life (years):	30

B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS

Plant Section	Process Contingency	Maintenance Cost Factor
Oil Handling	0.050	0.030
Oxidant Feed	0.050	0.020
Gasification	0.150	0.045
Low Temperature Gas Cooling	0.000	0.030
Selexol	0.100	0.020
Claus Plant	0.050	0.020
Beavon-Stretford	0.100	0.020
Boiler Feedwater Treatment	0.000	0.015
Process Condensate Treatment	0.300	0.020
Gas Turbine	0.125	0.015
Heat Recovery Steam Generator	0.025	0.015
Steam Turbine	0.025	0.015
General Facilities	0.050	0.015

C. DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS ($1,000)

Plant Section	Number of Units Operating	Total	Direct Capital Cost	Process Contingency
Oil Handling	1	1	0.	0.
Oxidant Feed	2	2	80363.	5494.
Gasification	3	3	33115.	6792.
Low Temperature Gas Cooling	2	2	29064.	0.
Selexol	2	2	15601.	2133.
Claus Plant	1	2	3043.	208.
Beavon-Stretford	1	1	8730.	1194.
Boiler Feedwater Treatment	1	1	5228.	0.
Process Condensate Treatment	1	1	8768.	3597.
Gas Turbine	3	3	105969.	18113.
Heat Recovery Steam Generator	3	3	31740.	1085.
Steam Turbine	1	1	51667.	1766.
General Facilities	N/A	N/A	63459.	4339.

D. TOTAL CAPITAL REQUIREMENT ($1,000)

Description	Annual Cost
Total Direct Cost	436746.
Indirect Construction Cost	87349.
Sales Tax	17907.
Engineering and Home Office Fees	54200.
Environmental Permitting	1000.
Total Indirect Costs	160456.
Total Process Contingencies	44721.

Project Contingency	112336.
Total Plant Cost	754259.
AFDC	120870.
Total Plant Investment	875129.
Preproduction (Startup) Costs	20394.
Inventory Capital	1237.
Initial Catalysts and Chemicals	7143.
Land	2550.
TOTAL CAPITAL REQUIREMENT ($1,000) -------->	915204.

E. FIXED OPERATING COSTS ($/year) -----------------------------------

Description	Annual Cost
Operating Labor	6791772.
Maintenance Costs	14849387.
Administration and Supervision	3819458.
TOTAL FIXED OPERATING COST ($/year) -------------->	25460617.

F. VARIABLE OPERATING COSTS ---

1. CONSUMABLES ($/year)

Description	Unit Cost	Material Requirement	Annual Operating Cost
Sulfuric Acid:	119.52 $/ton	1632.5 ton/yr	195112.
NaOH:	239.04 $/ton	328.3 ton/yr	78470.
Na2 HPO4:	0.76 $/lb	1488.3 lb/yr	1132.
Hydrazine:	3.48 $/lb	7156.9 lb/yr	24884.
Morpholine:	1.41 $/lb	6655.4 lb/yr	9401.
Lime:	86.92 $/ton	664.2 ton/yr	57738.
Soda Ash:	173.85 $/ton	733.5 ton/yr	127523.
Corrosion Inh.:	2.06 $/lb	132447.7 lb/yr	273431.
Surfactant:	1.36 $/lb	132447.7 lb/yr	179889.
Chlorine:	271.64 $/ton	20.5 ton/yr	5563.
Biocide:	3.91 $/lb	22744.3 lb/yr	88966.
Selexol Solv.:	1.96 $/lb	49230.0 lb/yr	96283.
Claus Catalyst:	478.08 $/ton	2.0 ton/yr	955.
Sul.. Acid Cat:	2.06 $/liter	0.0 liter/yr	0.
SCOT Catalyst:	249.91 $/ft3	0.0 ft3/yr	0.
SCOT Chemicals:	0.39 $/ft3	0.0 ft3/yr	0.
B/S Catalyst:	184.71 $/ft3	91.4 ft3/yr	16885.
B/S Chemicals:	N/A	N/A	197254.
Fuel Oil:	45.64 $/bbl	45991.3 bbl/yr	2098815.
Plant Air Ads.:	3.04 $/lb	3449.3 lb/yr	10494.
Raw Water:	0.79 $/Kgal	411164.7 Kgal/yr	326128.
Waste Water:	912.70 $/gpm ww	895040.5 lb/hr	1060841.
LPG - Flare:	12.71 $/bbl	4024.2 bbl/yr	51159.
TOTAL CONSUMABLES ($/year) ---------------------->			4900923.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)

Oil:	0.00 $/MMBtu	368078.3 lb/hr	0.
Ash Disposal:	10.87 $/ton	437.3 ton/day	1127229.
Byprod. Credit:	135.82 $/ton	2.4 ton/hr	(1685192.)

TOTAL VARIABLE OPERATING COST ($/year) ------------>	4342960.

G. COST OF ELECTRICITY --

Power Summary (MWe)	Auxiliary Loads (MWe)

```
Gas Turbine Output     591.19   Oil Handling   0.00  Claus         0.04
Steam Turbine Output   294.94   Oxidant Feed   58.47  B/S           1.88
Total Auxiliary Loads   75.38   Gasification    0.49  Proc. Cond    0.87
-----------------------------   Low T Cool.     1.55  Steam Cycle   4.23
Net Electricity        810.74   Selexol         1.00  General Fac   6.85

                                      Capital Cost:    1128.85 $/kW
                              Fixed Operating Cost:      31.40 $/(kW-yr)
        Incremental Variable Costs:   1.31 mills/kWh
        Byproduct Credit:             0.37 mills/kWh
        Fuel Cost:                    0.00 mills/kWh
                           Variable Operating Cost:       0.94 mills/kWh
COST OF ELECTRICITY  -------------------------------->  26.96 mills/kWh

     Heat Rate is:  8693. BTU/kWh.  Efficiency is: 0.3928

H.  ENVIRONMENTAL SUMMARY  -------------------------------------------
               INPUTS:   Oil          0.454 lb/kWh
                         Water         0.743 lb/kWh
               OUTPUTS   Water         0.078 lb/kWh
                         Ash           0.045 lb/kWh
                         WstWater      1.104 lb/kWh
                         CO2           1.441 lb/kWh
                         CO            0.000 lb/kWh
                         SO2        0.046658 lb/MMBtu
                         NOx        0.137122 lb/MMBtu
                         COS        0.000000 lb/MMBtu
                         NH3
Summary of Output Variables for Stochastic Analysis
DIRECT CAPITAL COST: Oil Handling:           0.00
                     Ox. Feed :          80362.74
                     Gasifiers:          33115.22
                     LowT Cool:          29064.33
                     Selexol  :          15600.65
                     Claus    :           3042.91
                     Beavon-St:           8729.81
                     Boiler FW:           5228.07
                     Proc Cond:           8767.88
                     Gas Turbi:         105968.71
                     Heat ReSG:          31739.58
                     Steam Tur:          51667.24
                     Gen Facil:          63458.81
 Total Direct Capital Cost              436745.95
 Cost of initial Catalyst and Chems      87349.19
 Cost of Taxes                           17906.58
 Engr and Home Office Fees               54200.17
 Indirect capital costs                 160455.95
 Project Contingency costs              112336.46
 Total plant cost                       754259.08
 Allowance for funds during constr           1.16
 Total process investment               875129.09
 Preproduction costs                     20394.05
 Initial chemicals                        1236.54
 TCICC                                    7143.39
 Land costs                               2550.00
 Total capital requirements             915204.36
 New Power                                 810.74
 OC Labor                              6791772.00
 OC Maintenance                       14849386.99
```

OC Admin & Supervision	3819458.04
Fixed Operating Costs	25460617.03
Consumables Operating Cost	4900922.93
Byproduct Credit	1685191.73
Ash Disposal Cost	1127228.88
Variable Operating Cost	4342960.08
Fuel Cost	0.00
Capital Cost $/Kw	1128.85
FOC, $/kW-yr	31.40
VOC, mills/kWh	0.94
Fuel Cost mills/kWh	0.00
Byproduct Credit, mills/kWh	0.37
Incremental VOC, mills/kWh	1.31
Cost of Electricity, mills/kWh	26.96
Heat Rate	8693.19
Efficiency	0.3928
CO2 Emissions	1.4413
CO Emissions	0.0001
SO2 Emissions	0.0467
COS Emissions	0.0000
CH4 Emissions	0.0000
H2S Emissions	0.0000
NOx Emissions	0.1371
No of Op. Gasifiers	3
No of Total Gasifiers	3
No of Plant operators	39
Ash output	0.0449
Oil Inputs	0.4540
Water inputs	0.7430
Water outputs (blowdown)	0.0778
Sulfur outputs	0.0060
Fixed Charge Factor	0.1034
Variable Levelization Cost Factor	1.0000

Gasifier oil feed, lb/hr	368078.29
SO2 emissions, lbmole/hr	5.1382
COS emissions, lbmole/hr	0.0000
Percent Ash, fraction	0.0990
Percent Sulfur, %	1.4000
Oil sulfur inlet, lb/hr	5153.0961
Percent Sulfur Capture, fraction	0.9681

APPENDIX C: PARTIAL RANK CORRELATION COEFFICIENTS FOR THE IGCC SYSTEMS