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Development and Application of Optimal Design Capability for Coal Gasification Systems - Task 1 (Volumes 1, 2, and 3)

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July 1995

September 1995

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For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia

MASTER

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Development and Application of Performance and Cost Models for the Externally-Fired Combined Cycle - Task 1, Volume 2

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For
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NOMENCLATURE

English Letter Symbols

A_j	=	Heat transfer surface of equipment or plant section j (ft^2)
A_c	=	Critical area of the gas turbine inlet nozzle (ft^2)
A_L	=	Area of land required (acres)
AF	=	Allowance for funds during construction (fraction)
ALR	=	Average labor rate ($/hour)
C_{elec}	=	Cost of electricity produced by plant (mills/kWh)
c_f	=	Plant annual capacity factor (fraction)
C_i	=	Cost of item i ($)
$C_{PC,i}$	=	Process contingency cost for plant section i ($)
$C_{PJ,i}$	=	Project contingency cost for plant section i ($)
$CHEM_{I,J}$	=	Initial chemical requirement for chemical i in plant section j (mass units)
CTP	=	Coal throughput per combustor (lb/hr/combustor)
d	=	Diameter (ft)
d_{min}	=	Minimum diameter (ft)
$D_{i.j.k}$	=	Density of species i at plant section j inlet or outlet k (lb/ft^3)
$DC_{j,l}$	=	Direct capital cost of plant section j, component l ($1,000 January 1989)
e_a	=	Annual escalation rate for plant equipment such as escalation rate (fraction)
E(x)	=	Expected (mean) value of variable x
f	=	Combustor fuel-to-air ratio
f_{ash}	=	Fraction of ash in coal (fraction)
f_{CaCO3}	=	Purity of limestone (fraction)
f_{cr}	=	Capital recovery factor (fraction)
$f_{e,j}$	=	Scaling factor for auxiliary electrical consumption of process section j
f_{EHO}	=	Engineering and home office cost factor (fraction)
f_{ICC}	=	Indirect capital cost factor (fraction)
$f_{M,i}$	=	Maintenance cost factor for plant section i (fraction)
$f_{PC,i}$	=	Process contingency for plant section i (fraction)
f_{PJ}	=	Project contingency (fraction)
f_s	=	Fraction of sulfur in coal (fraction)
f_{SO2}	=	Fraction of off-gas volume flow rate that is sulfur dioxide

f_{vclf}	=	Variable cost levelization factor (ratio)
FOC	=	Fixed operating cost ($/year)
$G_{i,j,k}$	=	Volumetric flow rate of species i, at plant section j, inlet or outlet k (acf/hr)
$h_{i,l}$	=	Height of equipment i, component l (ft)
HV_{coal}	=	Heating value of coal (BTU/lb)
HHV	=	Higher Heating Value of coal (Btu/lb)
i	=	Interest cost for spent funds (fraction)
I_{PCI}	=	*Chemical Engineering* plant cost index
IC	=	Inventory capital ($1,000)
IDC_i	=	Indirect capital cost of plant section i ($1,000)
INT(x)	=	Nearest integer value of x
IC	=	Inventory capital ($1,000)
k	=	Constant of proportionality
L_s	=	Sulfur loading in sorbent (weight fraction)
$m_{i,j,k}$	=	Mass flow rate of species i at the plant section j inlet or outlet k (lb/hr in all cases except for coal, where the units are tons/day)
$mw_{i,j,k}$	=	Molecular weight of species i at plant section j inlet or outlet k
$M_{i,j,k}$	=	Molar flow rate of species i at the plant section j inlet or outlet k (lbmole/hr)
MW_j	=	Electrical output of plant section j (megawatts)
n	=	Number of data points used in a regression analysis (integer)
N	=	Construction period (years)
N_B	=	Number of filter bags in the Fabric Filter unit (integer)
$N_{O,j}$	=	Number or operating trains of plant section j (integer)
$N_{s,j}$	=	Number of spare trains of plant section j (integer)
$N_{T,j}$	=	Number of total trains of plant section j (integer)
O_i	=	Number of operators in process area i per shift
OC_i	=	Operating cost for category i ($/year)
OC_M	=	Total maintenance cost
OC_{MM}	=	Maintenance materials cost
OC_{AS}	=	Administrative and support labor cost
OC_{ML}	=	Maintenance labor cost
$P_{j,k}$	=	Pressure at the plant section j inlet or outlet k (psia)
PCI	=	Plant cost index

PFC	=	Process facility capital cost
PP_i	=	Preproduction cost of category i ($1,000)
PPC	=	Total preproduction cost ($1,000)
PR	=	Pressure ratio
q	=	Heat flux (Btu/hr)
Q_j	=	Heat duty of process equipment or plant section j (Btu/hr)
Q_{coal}	=	Energy flow of coal (MMBtu/hr)
$R_{A/C}$	=	Air-to-cloth ratio of the fabric filter (acfm/ft^2)
$R_{Ca/S}$	=	Calcium-to-sulfur molar ratio (lbmole Ca/lbmole S)
$R_{L/G}$	=	Liquid to gas ratio for the FGD unit (gpm/Kacfm)
R^2	=	Coefficient of determination (decimal)
$r_{i,l}$	=	Radius of equipment i, component l (ft)
r_{tax}	=	Sales tax (fraction)
s	=	Sample standard deviation estimated from a data set
$SA_{i,l}$	=	Surface area of equipment i, component l (ft^2)
SF	=	Shift factor (labor shifts/day)
$t_{i,l}$	=	Thickness of equipment i, component l (ft)
$T_{i,j,k}$	=	Temperature of species i at the plant section j inlet or outlet k (°F)
TCR	=	Total capital requirement ($1,000)
TDC	=	Total direct cost ($1,000)
TIC	=	Total indirect cost ($1,000)
TPC	=	Total process capital ($1,000)
TPI	=	Total plant investment ($1,000)
U	=	Universal heat transfer coefficient (Btu/ft^2-hr-°F)
UC_l	=	Unit cost of component l ($/mass unit)
$V_{i,j,k}$	=	Velocity of species i at plant section j inlet or outlet k (ft/sec)
V_s	=	Superficial velocity (ft/sec)
Var(x)	=	Variance of variable x
VOC	=	Total variable operating cost ($/year)
$W_{e,j}$	=	Electricity requirement for plant section j (kW)
$W_{s,j}$	=	Shaft work for plant section j
Y_i	=	Mole fraction of species i (decimal)

Greek Letter Symbols

η	=	Efficiency (decimal)
Δp_j	=	Pressure drop across plant section j (psi)
ΔT_{lm}	=	Log mean temperature difference (^oF)
ΔT_h	=	Temperature difference at the hot side of the heat exchanger (^oF)
ΔT_c	=	Temperature difference at the cold side of the heat exchanger (^oF)
ε	=	Heat exchanger effectiveness (fraction)
τ	=	Residence time (sec)

Subscripts

a	=	Ambient
aux	=	Auxiliary
AS	=	Administrative and support labor
base	=	Represents the base case corresponding to the particular equipment or process area
cer	=	Ceramic material
cons	=	Consumables
elec	=	Electricity
EHO	=	Engineering and home office cost
EP	=	Environmental permitting
FC	=	Fixed operating cost
Fuel	=	Fuel
i	=	Inlet
IC&C	=	Initial catalyst and chemicals
ICC	=	Indirect construction cost
L	=	Land
m	=	Miscellaneous features
met	=	Metallic material
M	=	Maintenance
ML	=	Maintenance labor
MM	=	Maintenance materials
OC	=	Variable operating cost

plant	=	EFCC plant
PC	=	Process Contingency
PJ	=	Project Contingency
PR	=	Prepaid royalties
o	=	Outlet
r	=	Refractory
ref	=	Reference gas
s	=	Structural supports
SPI	=	Spare parts inventory
t	=	Total
tax	=	Sales tax
v	=	Vessel

Species

a	=	Air
aa	=	Auxiliary air
ads	=	Plant instrument air adsorbent
ash	=	Ash
cf	=	Coal feed
chem	=	Chemical
cm	=	Coal feed, moisture and ash free basis
cons	=	Consumables
fg	=	Flue gas
FO	=	Fuel oil
hy	=	Hydrazine
lime	=	Lime
L	=	Limestone
LPG	=	Liquified petroleum gasoline
mo	=	Morpholine
NO_x	=	Oxides of nitrogen
NO	=	Nitrogen oxide
NO_2	=	Nitrogen Dioxide
O	=	Oxygen

Ox = Oxidant

pw = Polished water

rw = Raw water

s = Sulfur

sa = Sulfuric acid

sh = Sodium hydroxide

slag = Slag from combustor

sludge = Sludge from FGD unit (solids and water)

SO_2 = Sulfur dioxide

sp = Sodium phosphate

Equipment/Plant Sections

C = Coal combustor

CH = Coal handling

PA = Primary air fan

SS = Slag screen

HX = Ceramic Heat Exchanger

GT = Gas Turbine

HR = Heat Recovery Steam Generator

BF = Boiler feed water system

ST = Steam turbine

FGD = Flue gas desulfurization

FF = Fabric filter

ID = Induced draft fan

GF = General facilities

1.0 INTRODUCTION

Increasing restrictions on emission of pollutants from conventional pulverized coal fired steam (PCFS) plants generating electrical power is raising capital and operating cost of these plants and at the same time lowering plant efficiency (Simbeck et.al, 1983). This is creating a need for alternative technologies which result in lower emissions of regulated pollutants and which are thermally more efficient. Natural gas-fired combined cycle power generation systems have lower capital cost and higher efficiencies than conventional coal fired steam plants, and at this time they are the leading contender for new power plant construction in the U.S. But the intermediate and long term cost of these fuels is high and there is uncertainty regarding their long-term price and availability. Coal is a relatively low cost fuel which will be abundantly available in the long term. This has motivated the development of advanced technologies for power production from coal which will have advantages of other fuels. The Externally Fired Combined Cycle (EFCC) is one such technology.

The Department of Energy (DOE) Morgantown Energy Technology Center (METC) has developed a performance model for an EFCC system based on a Hague International (HI) conceptual design of the system. This report documents the modifications to the METC performance model, and new cost models developed for the EFCC system. The objective is to obtain the output values of process performance parameters for a given set of input parameters, and to obtain capital and operating cost models that can be integrated with the EFCC process performance model. These models are implemented in ASPEN, a FORTRAN-based process simulator. A probabilistic modeling capability has been added to the ASPEN simulator, facilitating analysis of uncertainties in new process performance and cost (Diwekar and Rubin, 1989). One application of the probabilistic modeling capability is to explicitly characterize uncertainties in capital and annual costs, replacing the traditional approach of incorporating uncertainty via a "contingency factor."

The cost model includes both capital and annual costs. The capital cost models estimate the costs of each major plant section as a function of key performance and design parameters. A standard cost method based on Electric Power Research Institute (EPRI) Technical Assessment Guide (1986) was adopted. The annual cost models are based on operating and maintenance labor requirements, maintenance material requirements, and the costs of utilities and reagent consumption. Uncertainties in cost parameters are identified for both capital and operating cost models.

1.1 Overview of EFCC Systems

Environmental regulations are spurring the development of new technologies for electric power generation using coal. Current federal New Source Performance Standards (NSPS) applicable to coal-fired power plants require up to 90 percent sulfur dioxide (SO_2) removal, over 99 percent particulate matter (PM) removal, and moderate (about 50 percent) reduction of nitrogen oxides (NO_x) emissions. A conventional emission control system for a new pulverized coal-fired power plant typically consists of a wet limestone flue gas desulfurization (FGD) system for SO_2 control, an electrostatic precipitator (ESP) for PM removal, and combustion controls for NO_x reduction. These systems are all commercially available and well-demonstrated. However, recent commercial experience in Japan and Germany with selective catalytic reduction (SCR) indicates that 80 to 90 percent NO_x removal may be feasible, although SCR has been applied to date to only a small number of U.S. plants (Frey and Rubin, 1990; Cho et al, 1995). Furthermore, DOE and others have supported development of more advanced alternatives for control of SO_2 and NO_x emissions from pulverized coal (PC) power plants.

Externally Fired combined cycle (EFCC) systems are an alternative to conventional pulverized coal (PC) combustion. EFCC systems are capable of significantly higher thermal efficiencies and comparable emissions of CO_2, SO_2, NO_x, and particulate matter compared to a conventional PC fired steam (PCFS) plant. Therefore the emissions of air pollutants from an EFCC plant is expected to be lower than that for a conventional PCFS plant per unit of electrcity generated. Long term concerns over acid rain (for which SO_2 and NO_x are precursors) may lead to increasingly stringent emission regulations of conventional PC plants. Therefore, there is incentive to develop technology options to reliably achieve stringent emission reductions at minimum cost in a timely fashion. As a result, there has been significant research and development of EFCC and other advanced coal-based power systems.

Natural gas- and oil-fired systems based on gas turbine combined cycle technology have high efficiencies, but consume expensive premium fuels. In a combined cycle plant, fuel is burned in a gas turbine, and the hot exhaust gas is used to generate steam for a steam cycle. Electric generators on both the gas turbine and steam turbine generate electricity. By substituting a ceramic heat exchanger for a combustor where high pressure air can be heated by coal combustion flue gases, a coal-fueled gas turbine combined cycle power plant results. By integrating the steam cycle to generate steam from the high temperature coal combustion process, the overall thermal efficiency can be optimized. Advantages of EFCC plants over PC plants include higher thermal efficiency, reduced cooling water requirements (because gas turbines, rather than boiler/furnaces, generate a large portion of the power), and reduced land requirements.

1.2 Modeling Methodology for Performance Model

The EFCC model was developed as an ASPEN (Advanced System for Process ENgineering) 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 operation blocks, ASPEN 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 the sequential modular approach to flowsheet convergence. In this approach, individual unit operations models compute outlet streams and block results, given inlet streams and block input variables. When there are recycle loops in an ASPEN flowsheet, stream and block variables have to be manipulated in order to converge the flowsheet sequence. This can be done by three different mechanisms:

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

2. *FORTRAN blocks.* 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.

3. *Transfer Blocks.* Transfer blocks are used for information transfer. The user can transfer the values of a stream, sub-stream, attribute, or scalar flowsheet variable directly from one part of the flowsheet to another.

The METC performance model is used to calculate mass and energy balances for the EFCC system, to track environmental species, to conduct sensitivity analyses of performance parameters, and to evaluate design modifications. While the bulk of the model is comprised of generalized unit operation blocks, there are a number of FORTRAN blocks and design specifications which are specific to the EFCC system or to the flowsheet. There are also user models to handle coal properties, and there is a Fortran block used as a summary report writer to concisely present plant performance results. The flowsheet was developed using a modular approach to allow sections to be "borrowed" from other flowsheets, substantially reducing the development time of future EFCC simulation models.

1.3 Modeling Philosophy for Cost Estimation

There are a variety of approaches to developing cost estimates for process plants. These approaches differ in the level of detail with which costs are disaggregated into separate line items, 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. EPRI (1986) has defined four types of cost estimates: simplified, preliminary, detailed, and finalized. The cost estimates developed here are best described as "preliminary." The differences between different types of cost estimates are briefly described below.

A *simplified* cost estimate is based on information about major stream flow rates and design parameters from a simple process flow diagram. The cost information used in a simplified estimate typically includes published cost curves or scaling relationships for generic process areas or for the plant as a whole. A simplified cost estimate may also be based on adjusting costs from similar published or in-house work on the basis of a single performance parameter. A simplified estimate is thus sensitive to only one (or a few) major performance parameter(s), such as the coal feed rate or the plant electrical output.

A *preliminary* cost estimate is based on a more disaggregated consideration of the costs of specific process areas and specific equipment items. A preliminary estimate also includes the use of ratio or scaling relationships to adjust costs for a variety of operating conditions. The preliminary estimate is sensitive to a larger number of performance parameters (perhaps a few dozen) than the simplified estimate.

Detailed and *finalized* cost estimates are generally developed only for site-specific projects that are intended for construction. For a large process plant, these types of estimates may cost millions of dollars to prepare. They are based on vendor quotations for specific equipment costs in response to specifications developed by an architect/engineering firm.

For the purposes of evaluating alternative technologies, and for research planning, preliminary cost estimates are the most appropriate. Preliminary cost estimates are sensitive to the performance and design parameters that are most influential in affecting costs. Thus, the goal of this study is to develop preliminary cost estimates for the EFCC system under study.

A major constraint on cost model development is the availability of data from which to develop cost versus performance relationships for specific process areas or for major equipment items. Data from published studies can be used to develop cost models for specific process areas using regression analysis. Alternatively, cost models for process areas consisting of only one major equipment item can be based on published equipment cost curves, either in place of or as a supplement to regression analysis.

1.4 Integration of Performance and Cost Models

The cost models must be compatible with the ASPEN EFCC flowsheet simulations. The performance models must include the key parameters that are required to determine capital and operating costs. Generally, the performance models estimate all the parameters that are required for the cost models with a few exceptions. In the cases where the ASPEN models contain insufficient information for the cost models, regression models representing operating requirements are developed for the purpose of estimating annual costs.

1.5 Scope and Organization of This Report

The performance and cost models presented here represent initial versions based on currently available information for the processes selected. The costs for each plant section may be updated or refined as appropriate, depending on modeling needs and on the availability of additional data.

This report includes separate chapters describing the modification of the METC performance model and the models developed for capital and annual cost of the EFCC system. Chapter 2 outlines the technical background and design basis of the EFCC system. Chapter 3 documents the METC performance model and modifications made to the model. Chapter 4 describes the models developed for estimating the auxiliary power consumption more realistically. Chapter 5 describes the direct capital cost models and Chapter 6 then describes how the total capital cost of the EFCC system is calculated. Chapter 7 describes annual operating cost calculations. Chapter 8 describes the total annualized cost model used to calculate the cost of electricity. Chapter 9 presents a summary of the applications of the performance and cost model to various case studies and describes implementation of the cost model with the ASPEN performance models, sample results, priorities for further performance and cost model development, and recommendations for further work. The computer code for the newly implemented performance and cost model and sample outputs from these models are given in the appendices.

2.0 TECHNICAL BACKGROUND AND DESIGN BASIS FOR THE EXTERNALLY FIRED COMBINED CYCLE

2.1 Introduction

The Externally Fired Combined Cycle (EFCC) is an advanced power generation concept. It is intended to utilize coal or other ash bearing fuels such as peat, wood waste, and bitumen emulsion (LaHaye, Zabolotny, and Vivenzio, no date), with a potential for higher thermal efficiency and a lower cost than conventional coal-fired power generating systems. The concept incorporates the highly efficient gas turbine combined cycle technology for use with coal. In contrast to "direct fired" combined cycle technologies, the EFCC concept does not require any special fuel preparation, coal beneficiation, or coal conversion.

The EFCC concept has yet to be demonstrated commercially. The concept is under development by Hague International, Inc. (HI) of South Portland, Maine, and the U.S Department of Energy (DOE). The main obstacle in the development of a viable EFCC system has been the unavailability of a suitable heat exchanger which would allow for a sufficiently high gas turbine inlet temperature. Recent developments in ceramic heat exchanger (CerHx) technology for use in the EFCC process have been promising enough to warrant further development of this system.

This project involves development of engineering performance and cost models of the EFCC technology based on available performance and cost information. The purpose of these models is to enable evaluation of the EFCC based upon alternative assumptions regarding factors affecting process design, performance emissions, and cost. Thus, the models may be employed to compare alternative designs, evaluate plant efficiency, emissions, and cost, and compare the EFCC to other technologies. As part of previous work, performance, emissions, and cost models of conventional and advanced coal-based power generation systems have been developed (e.g., Frey and Rubin, 1990 & 1991; Rubin et al., 1986, 1991). These systems have included coal-fired power plants and Integrated Gasification Combined Cycle (IGCC). Concurrent work is focussing on development of models for PFBC (Diwekar, Rubin,and Frey, 1994).

In this report, the newly developed models for the EFCC are applied in a series of illustrative sensitivity analysis case studies. Since the EFCC is in the early stages of development, there are inherent uncertainties in the performance and cost parameter estimates. As part of future work to be reported in a separate topical report, probabilistic modeling capability will be incorporated to account for the uncertainties in the performance and cost parameters. Details of previous work on probabilistic modeling can be found in Frey and Rubin (1990, 1992 a,b,c). The results obtained from the EFCC model simulation will include the possible ranges of values for the performance, environmental emissions, and cost parameters, and information about the probability of obtaining these results. These results will be used to characterize key uncertainties; optimize the flowsheet configuration and parameter values; identify process areas for further research; and probabilistically compare competing advanced coal based power generating technologies, such as IGCC systems, with the EFCC technology in order to identify the risks and potential payoffs of the EFCC technology relative to the other technologies.

The potential advantages of the EFCC concept are (LaHaye, Zabolotny, and Vivenzio, no date, LaHaye and Bary, 1994):

1. Conversion of fuel to electricity in modular unit sizes, ranging from 50 MW to 300 MW, with high efficiencies, typically near 40 percent.

2. Operation on a wide variety of ash-bearing fuels, with acceptable atmospheric particulate matter emissions when coupled with available pollution control technologies (e.g., fabric filter).

3. Factory assembly of modules for low initial-cost and short lead-time from order to initial operation, with minimum field erection cost.

4. Adaptability to commercially available gas turbines including aircraft derivatives, and it can accommodate a variety of turbine improvements such as high pressure ratios, higher firing temperatures, and steam injection.

5. Avoidance of the problem of corrosion, erosion, and deposition of turbine components by using air rather than fuel combustion flue gas as the working fluid for the turbine .

6. For a pre-exisiting steam cycle, such as in the case of repowering of existing conventional coal-fired steam plants, the division of the available energy between the topping gas turbine and the bottoming steam cycle could be set by adjusting the CerHx flue gas side inlet temperature and effectiveness. This allows for a means of obtaining the steam conditions and steam flow required to satisfy the existing steam cycle. For a greenfield EFCC plant, which does not have a pre-existing steam cycle, the most efficient bottoming cycle steam conditions can be selected to obtain optimum EFCC performance.

7. Adaptability for repowering of existing steam-cycle power plants, resulting in increased output and overall plant thermal efficiency.

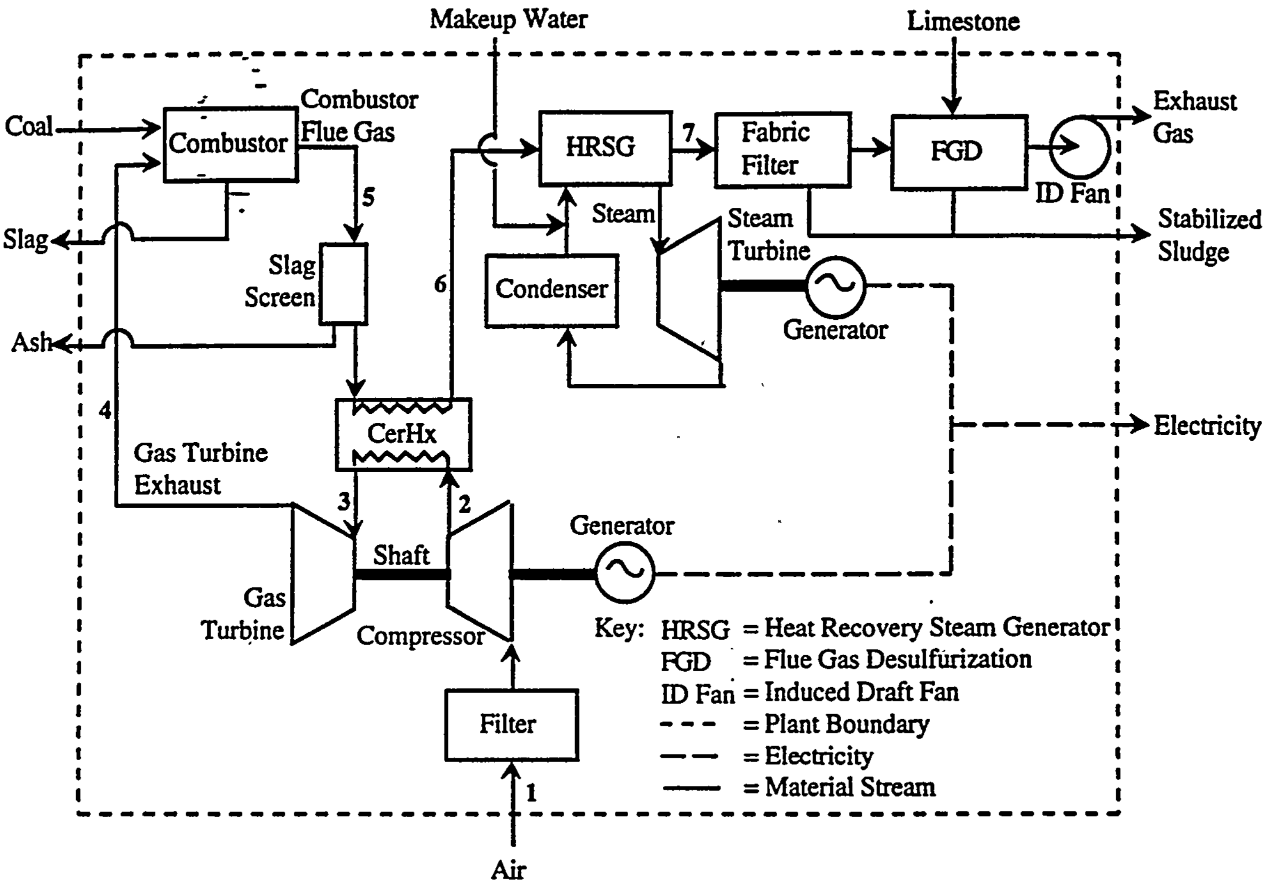


Figure 2.1. Process Flow Diagram of an EFCC Conceptual Model (Orozco and Seger, 1993)

Table 2.1. Typical Stream Process Conditions for an EFCC System[a,b]

Stream #	1	2	3	4	5	6	7
Stream Description	Inlet Air to Compressor	Compressed Air	Gas Turbine Inlet Air	Gas Turbine Exhaust	Combustor Exhaust Flue Gas	Heat Exchanger Exhaust	HRSG Exhaust
Flow Rate (lb/s)	918	918	872	872	1066	1074	1089
Temperature (°F)	59	685	2183	1087	2701	1564	122
Pressure (psia)	14.7	196.9	193	15.1	14.9	13.9	13.6

[a] Steam Cycle: 2600 psia/1050 °F/1050 °F reheat
[b] Vandervort and Seger (1991)

2.2 Process Description

A conceptual process model was developed by Hague International (HI) to estimate the performance of a 300 MW EFCC system (Vandervort and Orozco, 1992). Figure 2.1 shows the flow diagram for this process.

In the EFCC process a coal combustor and a CerHx, along with auxiliary components, replaces the gas turbine combustor of a natural gas-fired gas turbine. Flue gas from the coal combustor is used in the CerHx to heat high pressure air from the gas turbine compressor discharge. Hot, high pressure air from the CerHx is expanded in the turbine to generate electricity. The exhaust air from the gas turbine, which is at approximately 1000 ^{o}F and contaminant-free compared to conventional combustion turbine designs, is used as the combustion air for the coal combustor. Hence, this process is also called Exhaust-Fired Combined Cycle. Since the temperature of the exhaust flue gas from the CerHx is around 1500 ^{o}F, it can be used to supply heat to a bottoming steam cycle for further power generation.

Coal is combusted in an atmospheric pressure coal combustor. The combustor exhaust gases pass through a slag screen to remove large (> 12 micron) particles and then enter the shell-side of a ceramic heat exchanger. Clean filtered air is pressurized in a compressor before it enters the tube-side of the heat exchanger. In the heat exchanger, the thermal energy of the combustion flue gas is transferred to the high pressure air flowing in the tube-side. The temperature of the air is raised to the desired turbine inlet temperature. Internally insulated, high pressure piping is used to transport the hot, compressed air to the turbine, where it is expanded to provide the power to drive the compressor and the electric generator. Turbine exhaust air exits at a pressure slightly above one atmosphere, and enters the coal combustion chamber. Flue gas exiting the shell-side of the ceramic heat exchanger is at a high enough temperature to fire the bottoming steam cycle through a Heat Recovery Steam Generator (HRSG). Table 2.1 lists typical process conditions for the stream numbers shown in Figure 2.1. These data are based on EFCC design studies reported by HI (Vandervort and Orozco, 1992).

The gas turbine inlet air mass flow is less than that for the compressed air because of compressor air extraction for turbine blade cooling. Leakage from the high pressure air stream in the CerHx to the low pressure flue gas stream further reduces the mass flow rate of the inlet air stream to the gas turbine. Heat exchanger air leakage leads to an increase in the mass flow rate of the flue gas flowing through the HRSG. Furthermore, possible leakage of steam from the HRSG into the flue gas stream further increases the flue gas mass flow rate, although it is not expected to be signifcant (Jarvis, 1995). An Induced Draft (ID) fan is located downstream of the fabric filter to overcome flue gas pressure drops in the slag screen, CerHx, HRSG, FGD unit, and the fabric filter.

2.3 Major Process Equipment

The following description of the major process equipment in an EFCC process is based upon the HI conceptual 300 MW EFCC design (Vandervort and Orozco, 1992). The process conditions and process parameter values are drawn from this design study except where indicated.

2.3.1 Coal Combustor

Pulverized coal is combusted in a slagging coal combustor to provide the heat input to the turbine via the CerHx. Use of a slagging combustor minimizes the ash loading on the CerHx. The combustion chamber is vertically-oriented and down-fired. Slag accumulates on the walls, and then flows downward to a slag tap and quench tank at the combustor base.

Typically in a conventional coal-fired furnace the combustor walls are water-cooled, but the combustor design for an EFCC system incorporates an air-cooled wall in addition to the water walls. Such a design enables high flue gas temperatures to direct maximum heat to the gas turbine, thereby maximizing plant thermal efficiency. In one design, only 1.5 percent of the heat is transferred to the water walls and the bottoming steam plant (Vandervort and Orozco, 1992). The desired inlet temperature to the shell-side of the ceramic heat exchanger can be achieved by increasing the coal flow to the combustor to compensate for the heat absorbed by the combustor walls. Due to the high combustor temperatures, radiative heat losses from the furnace may be significantly higher than for conventional coal-fired boilers. The combustor excess air ratio is approximately 90 percent and the flue gas outlet temperature from the combustor is approximately 2700 ^{o}F (Vandervort and Orozco, 1992). The flue gas temperature is high enough to heat the high pressure air in the CerHx to turbine inlet temperatures typical of current state-of-the-art heavy duty commercial gas turbines (e.g., 2,300 ^{o}F - 2,350 ^{o}F).

2.3.2 Slag Screen

Ash deposition in the EFCC system components, such as the ceramic heat exchanger, is primarily controlled by the use of a slag screen The slag screen consists of a staggered array of refractory tubes, supported by a hexagonal lattice of refractory blocks. The tubes act as impact separators. Particle laden gases exit the combustor with a relatively high velocity and enter the slag screen, where all particles greater than 12 microns impact the slag screen tubes, and can be potentially collected and removed. At the slag screen exit, the flue gas is directed into an ash collection area, before being decelerated for entry in the heat exchanger. Within the CerHx, the low gas velocity and larger tube diameters significantly lower the particle collection efficiency.

Only particles larger than 30 microns in diameter are predicted to impact the heat exchanger tubes. However, such particles are expected to be collected in the slag screen upstream of the CerHx. Particles less than 12 microns are collected by a back-end exhaust clean-up system such as a baghouse. Although nearly all particles between 12 and 30 microns which impact the slag screen are expected to be collected by it, some may not be collected. These particles are swept around the heat exchanger tubes and collected by the back-end particulate matter control system (Vandervort and Orozco, 1992).

Since the slag screen tubes obstruct the flow of combustion flue gas, these are a source of pressure drop. For optimal efficiency of the EFCC, pressure loss between the combustor and the heat exchanger must be kept as low as possible. Optimal cycle efficiency requires that pressure drop across the slag screen be maintained at less than 10 in.w.g (0.356 psi) (Orozco and Seger, 1993). There is a tradeoff between slag screen performance and the pressure drop across it.

2.3.3 Ceramic Heat Exchanger

The ceramic heat exchanger, also known as the ceramic air heater, is a multi-pass shell and tube heat exchanger, consisting of a large number parallel tube strings. This design is reported to be the most cost-effective and efficient for application to the EFCC system as compared to other designs (Orozco and Seger, 1993). Most of the heat exchanger components are made of ceramics to accommodate material temperatures well in excess of 1900 ^{O}F and to resist corrosion from the combustion gas. Metal alloys can be utilized in the low temperature end of the heat exchanger, which operates in the range of 600 ^{O}F to 1400 ^{O}F. Each tube string consists of multiple (two or four) ceramic tubes mounted end-to-end. High pressure air from the gas turbine compressor enters the ceramic tubes of the heat exchanger through the inlet header. The product gases of coal combustion flow through the shell-side of the heat exchanger, and the heat energy is transferred from the shell-side to the high pressure tube-side air.

The ceramic tubes are held vertically in compression, supported by a spring pack and bellows assembly which provide the compressive forces. Figure 2.2 shows a two pass heat exchanger with a center "hex sleeve" that separates the two passes and heat exchanger adapter assembly. The spring pack and bellows assembly are located on the cold-end of the tube-string. These components are never exposed to temperatures above 750 ^{O}F (Vandervort, Bary, Stoddard, and Higging, 1993). The maximum shell-side temperature is kept below 2,750 ^{O}F because of material considerations (Vandervort and Orozco, 1992). The tubes are internally enhanced, resulting in a tube-side convective resistance that is approximately equal to the shell-side convective

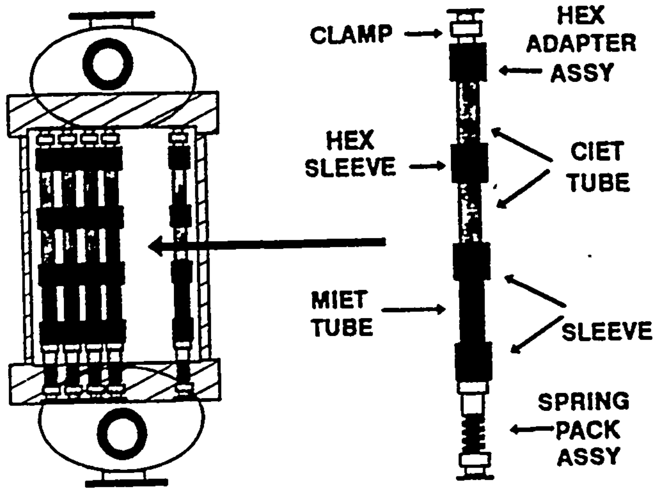


Figure 2.2. Ceramic Heat Exchanger and Tube-String (Vandervort and Seger, 1991)

resistance. This results in an overall heat transfer coefficient of approximately 10 Btu/hr-ft^2-^{o}F, and a heat transfer effectiveness of 0.85.

Heat transfer effectiveness of a two fluid heat exchanger is a dimensionless measure of the quantity of heat actually being transferred from the hot stream to the cold stream (Q_{act}) compared to the theoretical maximum heat transfer (Q_{max}). It is defined as Œ = Q_{act}/Q_{max}, and it is a unique measure of thermal performance. Q_{act} can be expressed either in terms of the hot stream parameters or the cold stream parameters. Q_{max} is the enthalpy change of the cold stream undergoing the maximum possible temperature change of ($T_{hot,in}$ - $T_{cold,in}$) without any losses. The heat transfer effectiveness can then be obtained from the outlet temperatures of either stream and is given by equation (2.1) (Shah, Kraus, and Metzger, 1990):

$$\varepsilon = (T_{cold,out} - T_{cold,in})/(T_{hot,in} - T_{cold,in}) = (T_{hot,in} - T_{hot,out})/(T_{hot,in} - T_{cold,in}) \qquad (2.1)$$

An important issue in the operation of the ceramic heat exchanger is the air leakage from the high pressure tube-side to the low pressure shell-side. Leakage of air results in reduction of pressure and mass throughput to the turbine, which reduces the turbine efficiency. At three percent leakage, the approximate penalty in the net cycle efficiency is reported to be one percentage point (Vandervort and Seger, 1991). Test results have shown air leakage to be as low as 0.5 percent of flow, and usually well under one percent (Staff Report, no date). High temperature gaskets are used at each heat exchanger tube joint to prevent leakage. Another issue in the development of ceramic tube material for the high pressure air heater is the possibility of high velocity fragments from a failed tube, initiating a chain reaction of tube failures. Hague International has developed a proprietary tube design which is reported to result in "graceful failure", wherein the failed tube does not damage adjacent tubes (Staff Report, no date).

Fouling of the outer walls of the heat exchanger tubes due to ash buildup is a potential problem because it reduces heat transfer and could eventually clog the entire heat exchanger. Therefore, there is a need for an upstream ash removal system. Ash particles larger than 30 microns have been found to significantly affect the heat exchanger tubes and the overall performance. Ash deposition in the heat exchanger can be prevented by the use of a slag screen. Ash deposits can be removed by soot blowers.

Soot blowers provide a means of removing particles that have collected on the heat exchanger tubes. Air or steam may be used for soot removal from the CerHx tubes.

The successful performance of the ceramic heat exchanger is strongly influenced by carbon burnout and the slag and ash capture capabilities of the coal combustor and slag screen (Orozco and Seger, 1993).

2.3.4 Gas Turbine

The gas turbine chosen for the process flow scheme described is a General Electric MS 7001F ("7F") turbine. At the time that previous design studies were done, this turbine had a simple cycle ISO rating of 150 MW, a heat rate of 9,880 Btu/kWh on natural gas fuel, and a turbine inlet temperature of 2,300 ^{o}F. Since that time the 7F has been upgraded to a 2,350 ^{o}F firing temperature. However, due to operating temperature limitations of the CerHx assumed in the design studies (Vandervort and Orozco, 1992), the turbine inlet temperature was assumed to be 2,180 ^{o}F. Ongoing ceramic material developments are expected to enable higher inlet gas turbine temperatures in the future. According to HI, increasing the firing temperature to a value of 2,300 ^{o}F would increase the system thermal efficiency by one percentage point. An increase to 2,350 ^{o}F would result in additional efficiency gains.

Currently available gas turbines would require modification to be employed in an EFCC system. The fuel valve and combustor of a gas turbine would not be needed with an EFCC system and would have to be removed. The gas turbine requires adaptation for appropriate interfaces with the CerHx. High temperature air piping is needed to and from the CerHx at the locations where the combustor would normally be located. Dual concentric piping is used in which the inner pipe carries the hot air to the turbine, and the outer pipe carries the compressor discharge air to the CerHx. Turbine control valves are incoporated in this piping. The power output of the gas turbine is regulated by these fast response valves, which allow a portion of the compressor discharge air to bypass the CerHx and mix with the hot discharge air to control the turbine inlet temperature (Staff Report, no date). These valves are the primary means of controlling the gas turbine power output. In this way, the CerHx can be maintained as a constant temperature heat reservoir. The slower responding portion of the load control is obtained by modulating the coal firing rate and the combustion air to the combustor.

A portion of the steam from the intermediate pressure steam turbine exhaust may be injected into the high pressure air inlet stream going from the compressor to the CerHx to increase the mass throughput and the heat capacity of the inlet stream to the gas turbine. This results in a higher gas turbine net power output. The overall plant efficiency tends to decrease slightly with increasing steam injection rate.

Since steam injection increases the mass flow rate to the gas turbine over and above its design basis, modifications have to be made to the gas turbine to accommodate the increased mass flow rate. Typically, the mass flow rate at the turbine inlet nozzle is limited by choking, i.e. the Mach number of the gas stream is unity. The mass flow rate into the expander is then proportional to the critical area, which is the area of the portion of the turbine inlet nozzle where the sonic velocity is reached, and total pressure, and is inversely proportional to the square root of the temperature at the turbine inlet. An equation representing this relationship is given by (Eustis and Johnson, 1990):

$$m_{air}\,(1+f) = k\,\frac{P_t}{\sqrt{T_t}}\,A_c \tag{2.2}$$

Conceivably the turbine inlet nozzle critical area can be increased to accommodate the higher mass flow rate, but this would involve a major turbine redesign and would not be economically practical unless a large market develops for EFCC. Other strategies which can be employed for this purpose are (Eustis and Johnson, 1990, Frey and Rubin, 1991):

1. Increasing the compressor pressure ratio, which would increase the density of the turbine inlet flow and hence the maximum mass flow rate;

2. Reducing compressor mass flow by partially closing the variable inlet guide vanes (IGV) at the compressor entrance;

3. Reducing the turbine inlet temperature which would allow an increased mass flow through the turbine inlet, although this strategy would impose an efficiency penalty on the system; and

4. Bleeding air from the compressor and adding it to the combustor inlet air stream.

Substantial gas turbine shaft power is consumed by the compressor in compressing the inlet air to the turbine. A reduction in compressor inlet air mass flow, which could be compensated by steam injection, would therefore represent savings in power usage within the gas turbine. Strategy two would be the most reasonable one for the purpose of modified gas turbine operation with steam injection.

2.3.5 Steam Cycle

The steam cycle for the base case process configuration outlined here uses a 2,600 psia/1050 ^{o}F/ 1050 ^{o}F reheat steam system (Vandervort and Seger, 1991). The option of using a super-critical bottoming cycle is available; however, in the 150 MW steam power range it is not expected to be economical.

2.3.6 Heat Recovery Steam Generator (HRSG)

An advantage of the EFCC system is that the steam bottoming cycle can be operated at a much higher temperature than in a conventional natural gas fired combined cycle. In a conventional system, the exhaust gas temperature is typically 1000 ^{o}F to 1,100 ^{o}F. In the EFCC system the bottoming cycle inlet gas temperature is typically 1,500 ^{o}F and may be higher, which enables a much higher temperature and pressure design of the steam system.

The steam system features a dual pressure unit that includes a superheater, low and high temperature evaporators, economizers, and reheater. The low pressure evaporator supplies a steam blanket for the deaerating feed tank (DFT). The DFT receives condensate and make-up water from the condenser hotwell. A feedwater pump supplies the high pressure evaporator following passage through the high pressure economizer section. Under the given steam conditions, the pinch point is 30 ^{o}F.

2.3.7 Steam Turbine

The steam turbine consists of three stages: (1) high pressure (HP); (2) intermediate pressure (IP); and (3) low pressure (LP). The HP turbine exhausts at 697 ^{o}F and 700 psia. The steam is then reheated to 1050 ^{o}F for expansion in the IP turbine. The IP turbine exhausts at 747 ^{o}F and 200 psia. A portion of the exit stream from the IP turbine is directed to the air inlet of the CerHx for steam injection into the gas turbine. In the base case design, the steam injection rate is 65.0 lb/s. The remaining steam flow is directed to the LP turbine, which exhausts at a pressure of one psia. Higher rates of injection will increase the mass flow and specific heat capacity of the gas turbine inlet air. A 20 percent higher mass flow of the turbine inlet gas can increase the power output of the gas turbine by 50 percent, however the plant efficiency is slightly reduced (Vandervort, Barry, Stoddard, and Higgins, 1993).

2.3.8 Environmental Control

Environmental emissions control for SO_2 and particulate matter in the EFCC system are based upon "back end" flue gas treatment technologies in the HI conceptual designs. NO_x control in the EFCC system is expected to be achieved by air staging of the combustion process. Environmental emissions control strategies for the EFCC system are discussed in the following three sections.

2.3.8.1 Sulfur Control

In an EFCC system sulfur control would be accomplished by back-end systems. An innovative sulfur control system based on an amine solvent was proposed for this process configuration. The potential advantages of this system are reported to be its relatively small size, low pressure drop, reduced auxiliary power consumption, and production of sulfur or sulfuric acid, both of which are marketable products. A disadvantage of this system is that the flue gas inlet temperature must be maintained below 130 ^{o}F, which requires the use of a special corrosion resistant heat exchanger in the HRSG economizer section to cool the sulfur laden gas to this temperature, which may be below the acid dewpoint. This system lowers the SO_2 emission rate to below 0.15 lb/MMBtu (Vandervort and Seger, 1991).

Conventional sulfur cleanup using a wet limestone scrubber or similar technology can also be employed for sulfur control. Sulfur dioxide in the flue gas is removed before venting to the atmosphere by the process of Flue Gas Desulfurization (FGD). SO_2 from the flue gas is removed by transfer of SO_2 from the gas to a slurry containing a calcium based material such as limestone. The SO_2 dissolves into the slurry and reacts with the limestone to form calcium sulfite. In a forced oxidation FGD system, the calcium sulfite is oxidized to calcium sulfate, which can be dried and more easily handled. High performance wet lime/limestone and lime spray dryer FGD systems exist which are capable of removing 90-98 percent SO_2 from the flue gas. The single loop countercurrent spray tower is the most commonly used device for the removal of SO_2 (Kalagnanam and Rubin, 1993). The design of spray towers for high efficiency without additives is achieved by using high liquid-to-gas (L/G) ratios and improving gas/liquid contact by spray nozzle design. The L/G ratio is the mass ratio of recirculating slurry and the flue gas. Organic acids such as dibasic acid or adipic acid can be added as buffers to improve performance. In the base case EFCC system, a wet limestone with forced oxidation FGD system has been assumed as the basis for model development. The main advantages of this process compared to the conventional systems are: (1) easy dewatering of the "sludge" containing spent sorbent; (2) more economical disposal of scrubber products; and (3) decreased scaling on tower walls.

Sulfur sorbent injection into the combustor may also be employed for sulfur removal, but this process would increase the mass of non-combustibles carried into the slag screen and ceramic heat exchanger. Furthermore, sorbent injection sets limits on the operating temperatures of the coal combustor (Vandervort and Orozco, 1992).

2.3.8.2 NO_X Control

Sources of NO_X from pulverized coal burning boilers can be grouped into three categories. The first is "thermal NO_X", which is generated at high flame temperatures (greater than 2300 ^oF) by thermal fixation of atmospheric nitrogen and oxygen. The second is the conversion of fuel-bound or organic nitrogen in the volatile fraction of coal. The third is the conversion of fuel-bound nitrogen in the coal char fraction. The latter two mechanisms depend on the fuel nitrogen content and its distribution between the volatile and the char fractions. Due to these different sources of NO_X formation, the emission levels of NO_X for different types of coal in the same combustion system can vary by a factor of two (Orozco and Seger, 1993).

Air staging (rich-lean) of the combustion process may be effective for NO_X control. It is estimated that rich/lean combustion may reduce NO_X emissions to below 0.1 lb/million BTU (Vandervort and Seger, 1991). In this process, the first stage operates fuel-rich, with fuel-air equivalence ratio from 1.67 to 1.0. Fuel-air equivalence ratio is define as the actual fuel-to-air ratio divided by the stoichiometric fuel-to-air ratio. The remaining combustion air is added in the second stage, where combustion is completed. The advantage of operating fuel-rich in the first stage is the lowered conversion of fuel-bound nitrogen to NO_X. Oxygen in the combustion air reacts more readily with the hydrocarbon fuel than with nitrogen, so under oxygen-starved conditions the nitrogen can recombine as molecular N_2.

2.3.8.3 Particulate Matter Control

One method of removing particulate matter from flue gas before venting it to the atmosphere is by means of fabric filters. These units are capable of removal efficiencies of greater than 99.9 percent and of producing clear stack plumes with less than one percent opacity. Fabric filters consist of a large number of long tubular filter bags arranged in parallel flow paths. As the ash-laden flue gas passes through these filters, almost all of the particulate matter is removed. Ash that accumulates on the bags is removed periodically by cleaning.

The key design variable for fabric filters is the air-to-cloth ratio, which is defined as the volumetric flow rate of flue gas divided by the total bag cloth area. It is expressed in units of acfm/ft^2. The air-to-cloth ratio is determined based on the bag cleaning method, which controls the residual amount of ash remaining on the bags. This in turn affects the resistance to gas flow and the pressure drop across the fabric filter unit.

There are four bag cleaning methods. These are Reverse Gas cleaning, Reverse Gas/Sonic cleaning, Shake/Deflate cleaning, and Pulse-Jet cleaning. The choice of bag cleaning method is usually base upon the type of coal being used (which determines the filterability of the ash) and the historical experience with filtering the particular kind of ash. Although reverse gas cleaning has been primarily used in the past, it is not as effective as other methods in achieving low capital and O&M costs (Kalagnanam and Rubin, 1994). Reverse gas and shaker systems impart more cleaning energy to the bags, which results in thinner residual dustcakes and lower pressure drop. Such systems can operate at a higher air-to-cloth ratio, which reduces costs. In recent years reverse gas with sonic cleaning has been emerging as the most acceptable method for full scale utility baghouses on pulverized coal-fired boilers. Because conventional reverse gas fabric filters are the most widely applied type of fabric filter for coal based applications, this design is assumed for the EFCC system.

2.4 Auxiliary Power Losses

In the HI conceptual EFCC design, the gas and steam turbine power outputs are 152.5 MW and 183.0 MW respectively, for a total of 335.5 MW. Auxiliary power losses total 25.9 MW, corresponding to 7.7 percent of gross output. Auxiliaries include units within the plant such as pumps, fans, and coal pulverizers. Therefore, taking into account the auxiliary power losses, the net power output is 309.1 MW. The net heat rate is 7780 Btu/kWh on a Higher Heating Value (HHV) basis, corresponding to a thermal efficiency of 44 percent. Increasing the steam injection rate can raise the total power output at no significant capital cost increase, but with a slight penalty on plant efficiency. The base case HI design does not account for water injection to gas turbine, flue gas reheat, and pressure drop across the fabric filter (Vandervort and Orozco, 1992).

2.5 Performance Data Needs

Table 2.2 list the typical process parameters for the EFCC process found in literature. Typical values, or ranges of values, for these parameters are given followed by a brief description of their implications for the EFCC system. A review of conceptual EFCC designs described in the literature reveals lack of reported information on several process areas.

Table 2.2. Process Parameter Values

Parameter	Typical Value	Description
Heat Transfer to Combustor Water Walls	1.5 percent [a]	Heat transferred to the water walls of the combustor which is directed to the bottoming steam plant. Some EFCC designs feature an air cooled combustor.
Overall Heat Transfer Coefficient of the CerHx	5.3[b]-10.0[a] Btu/hr-ft^2-^{o}F	The range of numbers reported is based on two values found in literature. It is not evident from literature whether the design value is based on clean tubes or fouled tubes.
Heat Transfer Effectiveness of the CerHx	0.85[a]	Definition of Heat Transfer Effectiveness is given in the text in section 2.3.3.
Maximum Shell Side Temperature of the CerHx	2750 ^{o}F[a]	This is the maximum temperature of the coal combustion flue gas exiting the combustor due to material considerations regarding the CerHx.
Maximum Tube Side Temperature of the CerHx	2450 ^{o}F	This temperature is limited by the ceramic heat exchanger material considerations.
Air Leakage Rate from Tube Side to Shell Side of the CerHx	0.5-1 percent[c]	This value represents the maximum expected leakage rate.
Pinch Points of Steam Cycle	30 ^{o}F[a]	Pinch point is the difference between the flue gas temperature leaving the evaporator and the saturation temperature corresponding to the steam pressure in the evaporator. It is assumed to be the same for all components of the steam cycle (Vandervort and Orozco, 1992).
FGD Efficiency	90-99 percent	Range of efficiency values for currently available wet limestone commercial units.
Combustor pressure drop	5" water[a]	
Slag Screen pressure drop	5" water[a]	Ceramic tubes in the slag screen obstruct the flow of the flue gas causing a pressure drop. There is a tradeoff between slag screen performance and the pressure drop. The optimal pressure drop across the slag screen should be < 10" water (Vandervort and Seger, 1991).
CerHx shell-side pressure drop	10" water[a]	Depends on ash deposition on the outside of the CerHx tubes, and is therefore dependent on the slag screen efficiency.

(continued)

turbine, ceramic heat exchanger, slag screen, and a 7.4 MW_t (25×10^6 Btu/hr) combustion system. The heat exchanger will function under actual gas turbine operating conditions. Therefore, the tubes will be pressurized and thermally cycled under gas turbine operating conditions. The primary objectives of this phase are to demonstrate that the ceramic heat exchanger can be reliably pressurized up to 165 psia and that it can withstand exposure to coal combustion products. Work on this phase began in December, 1991 and was completed in mid-1994 (LaHaye and Bary, 1994). The results of the test facility are not available publicly at this time.

Phase III will involve work on a "Prototype Externally Fired Combined Cycle", and phase IV will involve a "Commercial EFCC Demonstration" (LaHaye, Zabolotny, and Vivenzio, no date).

Phase III, or the demonstration phase, of the program was selected under Round V of the U.S Clean Coal Technology Program in May 1993. The objective of this phase is to repower an existing coal-fired plant in the Pennsylvania Electric Company (Penelec) system at Warren, Pennsylvania. To demonstrate the advantages of repowering an existing coal fired steam plant with an EFCC, Hague International developed a conceptual design for repowering an actual plant. This design formed the basis for the Warren plant repowering (Staff Report, no date). The estimated start up date for this project is November 15, 1996 (LaHaye and Bary, 1994).

2.7 EFCC Cost

HI developed a cost estimate for a conceptual EFCC system. For the purpose of developing the estimate, the EFCC plant was divided into two segments; the Power Island and the Balance-of-Plant. The major components of the power island are the coal combustor, slag screen, CerHx, gas turbine, HRSG, high temperature piping and turbine control valve, and controls and instrumentation. The Balance-of-Plant components include the FGD unit, baghouse, chimney, lime and scrubber waste silos, transformers, coal bunkers, and access roads (LaHaye and Bary, 1994). A cost estimate for an EFCC plant takes into account the capital cost of these items as well as the operation and maintenance cost.

2.7.1 Capital Cost

An assessment of capital cost estimates of coal-fired plants for a nominal 250 MW design is shown in Table 2.3 (LaHaye, Zabolotny, and Vivenzio, no date). The comparisions of EFCC to other systems are based on data reported by LaHaye et al (no date). There is a considerable variability in the performance and cost of different IGCC and PCFS power plants. Therefore, comparisions based on single point estimates can be misleading. For example, IGCC system costs as reported by others, vary from $1,100/kW to $2,000/kW (Frey and Rubin, 1992b and 1992c).

Table 2.3. **Summary of Capital Cost Comparisons of 250 MW Coal-Fired Plant Designs as reported by Hague International**[a,b]

(Million 1985 Dollars)

Cost	Technology			
	IGCC	PC	AFB	EFCC
Total Direct Cost	314.0	286.0	238.6	201.2
Indirect Cost	31.3	32.5	26.1	24.2
Project Contingency	51.8	47.9	39.7	33.8
Process Contingency	34.5	9.6	26.5	31.1
Total Present Day Cost	431.6	376.8	330.9	303.1
AFUDC	19.0	16.5	14.6	12.8
Total Plant Investment	450.6	393.3	345.5	290.3
Dollars/KW	1,800	1,600	1,400	1,200

[a]PC - Pulverized Coal; IGCC - Integrated Gasification Combined Cycle; AFB - Atmospheric Fluidized Bed

[b] LaHaye, Zabolotny, and Vivenzio (no date)

Project contingency represents the expected increase in the capital cost estimate that would result from a more detailed estimate of the project. These estimates are based on the costs of major equipment represented in the process flow diagrams of the major plant sections. Process contingency represents the expected increase in the capital cost estimate of a technology due to the state of technological maturity. Process contingencies for new technologies are intended to represent the expected cost of a commercialized (e.g., fifth of a kind) plant. As commercial experience with the technology is gained the process contingency decreases, and may become zero for a fully commercialized process. The process contingency for a technology in the early stages of development is expected to be high (EPRI, 1986).

Process contingency factors reported in different studies differ considerably. The traditional contingency factor approach for new processes are known to underestimate actual capital costs (Milanese, 1987). Adjusting the capital cost of a technology by the conventional contingency factor method uses deterministic point estimates of contingency factors. This method is inadequate for quantifying the risk of cost increase for an advanced process technology. Probabilistic analysis can be used to overcome this shortcoming. In the probabilistic approach the disaggregated and quantitative consideration of uncertainties for specific

process parameters in the engineering model are taken into account. Uncertainty of a process parameter is quantified using probability distributions instead of point estimates (Frey and Rubin, 1990). Such an approach will be employed in future case studies.

The project contingency cost as a percentage of the total direct cost (TDC) is approximately the same for all the processes, whereas the process contingency cost as a percentage of the TDC is highest for the EFCC plant. This is expected because the EFCC technology is at an earlier stage of development compared to other technologies. The total cost, total plant investment, and cost per KW of electricity generated is reported to be the lowest for the EFCC technology. However, the details of the cost assumptions underlying this comparison were not reported.

2.7.2 Operating and Maintenance Cost

Operating and maintenance (O&M) costs of the EFCC technology and three other coal-fueled alternatives were developed by HI in a preliminary assessment of the EFCC technology. Considerations were given to fixed O&M costs, variable O&M costs, and consumables consisting of ceramic tubes, limestone, makeup water, liquid wastes, ash and slag disposal, sludge disposal, and sulfur scale. The results of this analysis are shown in Table 2.4.

Fuel cost for a power generating technology depends on the heat rate or plant thermal efficiency. Since the EFCC technology is estimated to have a higher thermal efficiency than other technologies, the fuel cost is lowest for an EFCC plant, assuming that the coal price and characteristics are identical for each technology. The EFCC technology non-fuel O&M costs are estimated to be 20-30 percent lower than other alternatives.

2.7.3 Cost Data Needs

The reported cost data are for the entire EFCC process. Cost estimates for individual process areas in the Power Island and the Balance-of-Plant are needed. These process areas are coal combustor, slag screen, CerHx, gas turbine, HRSG, steam turbines, FGD unit, and fabric filter unit. Cost models for selected process areas are available from Frey and Rubin (1990). Cost models for selected components of the environmental control section are available form Rubin et al. (1986, 1990) and Kalagnanam and Rubin (1993), and will be adapted for the EFCC process.

Table 2.4. Summary of O&M Cost Comparisions of 250 MW Coal-Fired Plant Designs as Reported by Hague International[a]

(Million 1985 Dollars)

Cost	Technology			
	IGCC	PC	AFB	EFCC
Fixed O&M	6.4	6.1	4.9	4.5
Variable O&M	2.7	2.6	2.1	1.9
Consumables	1.3	2.4	3.9	2.0
Fuel	15.2	15.8	15.5	12.9
Total O&M Cost (mills/kWh)	25.6	26.8	26.4	21.3

[a] LaHaye, Żabolotny, and Vivenzio (no date)

2.8 Discussion of Uncertainties

The development and analysis of conceptual process designs for the EFCC power generation technology has shown the potential of high thermal efficiency and low cost, with acceptable environmental emissions. However, no "fifth-of-a-kind" or commercial EFCC plant is operational as yet. Therefore, making predictions regarding the mature commercial scale performance and cost of an EFCC plant involves uncertainties.

The CerHx for EFCC application has not yet been fully developed and tested under actual operating conditions. Most of the work done on the development of the CerHx has been related to modifying a low pressure recuperator for EFCC application. The performance of the CerHx under high pressure and in a corrosive coal combustion flue gas environment is still uncertain.

The inlet temperature to the bottoming steam plant is much higher than that experienced in a conventional coal-fired power plant, which introduces uncertainty regarding its performance. Several modifications have to be made to a commercially available gas turbine for use in an EFCC plant. Such modifications, although conceptually possible, have yet to be proven feasible on a commercial scale. Innovative sulfur removal technology based on an amine solvent has been proposed to control sulfur emissions, but such a system has not yet been proven successful on a commercial scale. Instead, a wet limestone system may be employed which has been commercially demonstrated and has established performance and cost.

A wet bottom, vertically oriented, slagging combustor has been proposed for the EFCC system. A full scale commercially viable combustor for the EFCC system has not yet been demonstrated. Similarly, a full scale version of the slag screen, ceramic heat exchanger, and a modified gas turbine for the EFCC process have also not been commercially proven. Therefore, the cost of these process equipment areas are uncertain due to uncertainties in their performance and availability.

2.9 Scope of this Report

In this report, a previously developed performance model of the EFCC is enhanced and new cost models are developed. Thus, the scope of the report includes:

1. Documentation of a performance model previously developed by U.S. DOE.
2. Revision of the existing performance model.
3. Development of a new cost model.
4. Integration of the new cost model with the performance model.
5. Sample model applications.

3.0 DOCUMENTATION AND MODIFICATION OF THE METC PERFORMANCE MODEL FOR THE EXTERNALLY FIRED COMBINED CYCLE

3.1 Process Description

The Department of Energy (DOE) Morgantown Technology Energy Center (METC) has developed a performance model for a 264 MW_{net} EFCC system based on a Hague International (HI) conceptual design of the system.

The METC model represents a modified HI EFCC design. It consists of a slagging combustor fueled by Illinois No. 6 coal, a ceramic heat exchanger (CerHx), a 2300°F turbine inlet temperature gas turbine, a heat recovery steam generation (HRSG) system, a 1785 psia, 1050°F superheater, 1050°F reheater steam cycle, and a flue gas desulfurization (FGD) unit. The flue gas exiting the combustor passes through the CerHx and HRSG, and is then treated in a wet limestone FGD scrubber to remove sulfur dioxide. The CerHx indirectly heats the gas turbine expansion stream to the turbine's inlet temperature.

In developing the ASPEN performance model, METC made two simplifications to the HI base case design:

1. The intermediate pressure steam from the steam cycle, which is mixed with the compressed air before entering the CerHx, is set to zero as a default. This mixing is commonly referred to as steam injection. HI had set this flow to 8% of the hot air flow. The user may adjust the level of steam injection.

2. The water walls from the slagging combustor, which were included in the base case HI design for cooling purposes, were not incorporated into the METC model. If the heat loss from the coal combustor were to be significant then the water walls may be used to generate steam for the steam cycle steam (Micheli, 1995). A fraction of the economizer outlet water stream is directed to the water walls, and the steam generated is mixed with the inlet stream to the high pressure boiler. This would increase the mass flow of steam in the steam cycle, resulting in an increased power output of the steam turbine.

Several modifications have been made to the existing METC model. Modifications were made to the following process sections: coal handling, coal combustor, ceramic heat exchanger, gas turbine, steam side of the heat recovery steam generator, and the environmental control section. The details of these modifications are presented in Section 3.5.

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

Proximate Analysis, wt-%, run-of-mine basis	
Moisture	12.0
Fixed Carbon	54.3
Volatile Matter	35.7
Ash	10.0
Ultimate Analysis, wt-%, run-of-mine basis	
Carbon	69.53
Hydrogen	5.33
Nitrogen	1.25
Chlorine	0.0
Sulfur	3.86
Oxygen	10.03
Ash	10.00
Ash Fusion Temperature, ^oF	2,300

In the following section we briefly describe the EFCC process performance model previously developed by METC. Details regarding the METC model are given in Appendix A. We then discuss the major changes and updates that were made to the model as part of this work.

3.2 Major Process Sections In The METC EFCC Model

Each major flowsheet section is described below. Detailed flow diagrams and tables summarizing the ASPEN model are given in Appendix A.

3.2.1 Coal Combustor

In the METC model, as-received Illinois No. 6 bituminous coal is fed pneumatically to a slagging combustor. Table 3.1 lists the default coal properties used for this simulation. FORTRAN block SET-NCRX defines the coal stoichiometry. Auxiliary air is required for transporting coal to the combustor. The mass flow rate of coal and auxiliary air is initialized in the ASPEN input file. The higher heating value of the coal on an as-fired basis is 12,774 Btu/lb, and on a dry basis is 11,241 Btu/lb. The exhaust from the gas turbine, at approximately 1,050 ^oF, is used as the oxidant stream to the coal combustor. The combustion reaction temperature is calculated in the ASPEN flowsheet simulation and is based upon a specific turbine inlet temperature (T_i). The combustor flue gas exit temperature needs to be high enough to raise the temperature of compressed air to the turbine inlet temperature given heat losses and limited heat exchanger effectiveness. For the base case simulation a turbine inlet temperature of 2,300 ^{o}F was assumed by METC and the simulated combustor flue gas exit temperature is 2,775 ^{o}F. In the model the heat loss from the combustor was combined with the loss from the CerHx, and the total was assumed to be two percent of the CerHx heat duty.

In the METC model, the auxiliary air flow rate was not scaled to the coal flow rate. Instead, it was specified as a constant mass flow rate independent of the coal flow rate. The flow rate of auxiliary air is proportional to the coal flow rate, and is expected to increase with an increase in the coal flow rate. The combustor heat loss was incorporated with the ceramic heat exchanger heat loss in the METC. The combustor heat loss depends on the combustor design and is estimated as a percentage of the combustor heat duty. Therefore, it is necessary to separate the combustor heat loss from the CerHx heat loss. The METC model did not incorporate combustor water walls for cooling of combustor walls and steam generation in the steam cycle, whereas a number of design studies have included combustor water walls (Parsons and Bechtel, 1991; Vandervort and Orozco, 1992; and LaHaye and Bary, 1994)

3.2.2 Ceramic Heat Exchanger

The simulation model was based on HI's tubular ceramic heat exchanger. The pressure drops for both the flue gas and turbine sides were taken from HI's EFCC design. The temperature of the flue gas in the combustor is calculated based on the CerHx effectiveness and the inlet and exit CerHx air temperatures. The CerHx effectiveness is assumed to be 85 percent. The heat loss from the CerHx was estimated as a percentage of the heat exchanger heat duty. The heat required to raise the temperature of the air exiting the CerHx for a specified turbine inlet temperature was estimated after taking into account the heat loss. This heat is transferred from the flue gas flowing in the hot side of the CerHx.

A fraction of the high pressure air flowing though the CerHx tubes is expected to leak and mix with the flue gas flowing through the shell side. Air leakage would affect the gas turbine and steam turbine power output, and hence the net plant efficiency. The METC model did not account for the CerHx air leakage.

3.2.3 Gas Turbine

Atmospheric air, at the ambient conditions of 59 ^{o}F, 14.7 psia, and 60 percent relative humidity, enters the gas turbine compressor. The compressor pressure ratio was assumed to be 13.5, which is typical of the original performance specifications for heavy duty gas turbines such as the GE 7F. Thus, the compressor outlet pressure was 198.5 psia. The pressure drop across the compressor inlet air filter was assumed to be 0.5 psi. The compression was modeled using three stages of unit operation blocks. The outlet pressures in the first, second, and third stages were specified to be 52.6 psia, 81.3 psia, and 198.5 psia respectively. The isentropic efficiency for the compression section was 0.92. Based upon these assumptions, the simulation modeled yielded a 688 ^{o}F outlet discharge air temperature (Chen and Jarvis, 1993).

The compressed air was modeled as being indirectly heated to the 2,300 ^{o}F turbine inlet temperature by the CerHx. The expansion of hot high pressure air was modeled in three stages. The outlet pressure in the first, second, and third stages are specified to be 83.0 psia, 54.0 psia, and 15.2 psia respectively. An isentropic efficiency of 0.913 was assumed for each of the three expansion stages.

The exhaust gas from the gas turbine was fed to the slagging combustor. The model result for the net power produced by the gas turbine was 134 MWe. The gas turbine model accounted for pressure losses in the inlet air filter and across the CerHx. It also accounted for the energy penalty associated with the cooling air requirements for high temperature turbine rotor blade cooling. The METC model addressed cooling air flow using a two-step approach. A portion of the compressor discharge air was assumed to be diverted from the compression discharge for use in combustor lining and turbine cooling. Of the total cooling flow, 42.44 percent was assumed to add to the power output of the gas turbine as if it had been introduced through the gas turbine nozzle, and is referred to as "non-chargeable". The rest of the cooling flow was assumed to not contribute to the gas turbine power output and is referred to as "chargeable". The chargeable cooling flow is injected at the gas turbine exhaust to represent pressure drops and other efficiency losses associated with the cooling circuitry. The mass flow rates of ambient air inlet to the gas turbine compressor and steam injection from the steam cycle are initialized in the ASPEN input file.

In the METC gas turbine model the air mass flow was specified as a constant at the compressor inlet. The mass flow rate at the expander inlet was calculated based upon the compressor discharge and the chargeable cooling air. Typically, the gas turbine air mass flow is limited at the expander inlet and not the compressor inlet. Thus, the assumption of constant compressor mass flow would not be correct for model applications involving steam or water injection. In a gas turbine, the cooling air flows are extracted at multiple points within the compressor to better match the pressure requirements at various stages in the turbine. However, the METC model considered the cooling flow split only at the compressor discharge outlet.

3.2.4 Heat Recovery Steam Generator (HRSG)

The HRSG was modeled as consisting of a 1,785 psia and 1,050 °F superheater, a 1,050 °F reheater, an economizer, a high pressure boiler, a low pressure boiler, and a condenser operating at 2" Hg (0.98 psia). The inlet steam to the high pressure economizer and the make-up water for steam generation was initialized in the ASPEN input file. The low pressure boiler was used to produce steam for the deaerator from the flue gas leaving the economizer. The flue gas exits the low pressure boiler at 251 °F. The HRSG process area accounted for heat losses in each heat exchanger.

3.2.5 Steam Turbine

Steam generated in the HRSG was modeled as being expanded in four stages. These stages included a high pressure turbine, followed by an intermediate pressure turbine, followed by two low pressure turbines.

3.2.6 Flue Gas Desulfurization (FGD)

A wet limestone FGD process with forced oxidation was represented in the METC flowsheet. Ninety five percent of the SO_2 was assumed to be converted to calcium sulfate using a 10 weight percent Greer Limestone slurry. The limestone used for the flue gas desulfurization consists of 98.8 percent calcium carbonate and 1.2 percent silicon dioxide by weight. The mass and composition of the limestone slurry, and the work requirement by the FGD unit, were initialized in the ASPEN input file.

The limestone slurry feed was initialized in the METC ASPEN model. FORTRAN block SETFGD is used to reset the flow rate of limestone slurry to the FGD unit based on a user specified molar ratio of calcium-to-sulfur and the molar flow rate of SO_2 in the flue gas entering the FGD unit. A molar calcium-to-sulfur ratio of 1.05 was specified in the METC model and the molar flow rate of SO_2 was calculated directly from the ASPEN model simulation.

Power consumed by the FGD unit was initialized using a dummy stream, and was reset in the FORTRAN block FGDWORK. The FGD power requirement was calculated using a linear regression equation, which is based on the SO_2 concentration in the flue gas and the total flue gas mass flow rate. The estimated auxiliary power consumption by the FGD units was intended to be conservative, and the actual value of the power consumed was expected to be lower than that estimated by the METC model (Jarvis, 1994). Several shortcomings to the representation of the FGD system in the METC model have been identified.

Sludge from the FGD process typically contains 40-60 percent water by weight (Cooper and Alley, 1986; EPRI, 1991). However, in the METC model, the sludge from the FGD unit was assumed to be dry. The portion of the liquid water coming out of the FGD reactor vessels, which does not exit the FGD unit with the sludge, is usually recirculated to the limestone slurry feed. However, in the METC model the liquid water exiting the FGD reactor was flashed to water vapor with the flue gas exiting the FGD reactor. The nominal temperature of the flue gas exiting the FGD unit for a limestone forced oxidation process has been reported to be approximately 127 ^{o}F (EPRI, 1991), which is not high enough for buoyancy of stack gas and prevention of acid condensation. The temperature of the flue gas is usually raised by heating it in a reheater using steam from the steam cycle. Steam usage from the steam cycle for FGD flue gas reheat reduces the gross power output of the steam cycle and therefore imposes a penalty on the net plant efficiency. In the METC model the temperature of the flue gas exiting the FGD unit was 118 ^{o}F and was raised to 127 ^{o}F by flashing the flue gas stream. The heat required for raising the temperature of the flue gas was not accounted for in the METC performance model. Furthermore, the model did not consider reheat of the flue gas from 127 ^{o}F to a temperature high enough for sufficient plume buoyancy. Such temperature are typically 150 to 200 ^{o}F.

3.2.7 Plant Energy Balance

There are four main groups of energy balance calculations. The first is for estimating the gas turbine section power output. The second is to estimate the total gross power output of the steam turbine. The third is to estimate the total power consumption of auxiliary loads that are explicitly modeled in the flowsheet. The last is to estimate other auxiliary loads using a simple multiplier. The energy input to the EFCC system was estimated as a product of the coal flow rate and the higher heating value of the coal.

In the METC performance model the auxiliary loads have been assumed to be three percent of the net electrical power output. The auxiliary power requirement is a function of system performance variables and is expected to change with a change in their values. A change in auxiliary power affects the net plant power output and plant efficiency. Therefore, a better estimate of the plant auxiliary power requirement is needed. New auxiliary power requirement models have been developed based on process variables and integrated with the ASPEN performance model. These models are documented in Chapter 4.

3.3 Process Parameter Assumptions

The assumed process parameters for the EFCC base case simulation are listed in Table 3.2. Several process parameters, such as the particle removal efficiency of slag screen, CerHx, and the fabric filter, particle size distribution of ash in the flue gas from the combustor, FGD reheat, and others, have not been considered in the original METC flowsheet.

3.4 Convergence Sequence

The convergence sequence for the METC performance model simulation is based on five design specifications one and FORTRAN block. A design specification, CONTIT, is used to calculate the required temperature of the hot air leaving the ceramic heat exchanger to produce the specified turbine inlet temperature after a gas turbine cooling bleed stream is assumed. The turbine inlet temperature can be set by the user in this design specification block in the input file. The FORTRAN block, EFFECT, sets the desired temperature of the flue gas exiting the slagging combustor based on the temperature calculated by CONTIT and an assumed ceramic heat exchanger effectiveness. The ceramic heat exchanger effectiveness can be set by the user in this block. Design specification CONCOALF adjusts the coal flow rate to the slagging combustor to achieve the required temperature set by FORTRAN block EFFECT. The coal flow rate is adjusted by setting the heat loss from the combustor to a user specified value. In the METC model the combustor heat loss was set to zero.

The steam cycle convergence is based on the temperature of the flue gas stream leaving the ceramic heat exchanger and entering the HRSG, and a 251 °F HRSG exit temperature. First, the required steam cycle circulation rate is calculated, by design specification CIRC-RAT, based on the temperature of the flue gas entering the HRSG and an assumed flue gas temperature leaving the high pressure boiler. This temperature is constrained by an assumed minimum pinch temperature difference of 30 °F. Design specification LPBOIL then calculates the fraction of low pressure boiler water leaving the deaerator that must be used in the low pressure boiler to generate steam. The flue gas from the economizer is cooled to 251 °F and the resulting low quality heat is utilized to generate steam in the low pressure boiler. The low pressure boiler produces steam for the deaerator. Finally, design specification CONDEAER calculates the additional amount of low pressure steam which must be extracted from the low pressure turbine to heat the deaerator. This is done by setting the heat loss around the deaerator to a user specified value. In the METC model the deaerator heat loss was set to zero.

Table 3.2. Assumptions for METC Base Case Simulation

Process Unit Parameter	Units	Value
Gas Turbine		
Pressure drop across the compressor inlet ducting	psi	0.5
Compressor pressure ratio		13.7
Compressor isentropic efficiency	%	92.0
Expander isentropic efficiency	%	91.3
Generator efficiency	%	98.6
Turbine Inlet Temperature	°F	2,300
Total Cooling requirement	% compressor outlet air	19.6
Non-chargeable cooling air (cooling used around Gas Turbine inlet)	% of total cooling requirement	42.22
Chargeable cooling flow (cooling flow fraction which does not contribute to power generation in the gas turbine)	% of total cooling requirement	57.78
Coal Combustor		
Higher Heating Value of coal, dry basis	Btu/lb	12,774
Auxiliary Air flow rate	lb/lb coal	0.79
Pressure drop	psi	0.11
Ceramic Heat Exchanger		
Effectiveness		0.85
Heat loss	%	2.0
Flue Gas side pressure drop	psi	0.76
Turbine Air side pressure drop	psi	3.50

(continued)

Table 3.2. (continued)

Process Unit Parameter	Units	Value
Steam Cycle (1785 psia/1050 °F/1050 °F)		
Steam Turbine Generator efficiency	%	98.6
Deaerator Vent loss	%	1.0
Deaerator pressure	psia	17.5
Bottoming Cycle heat loss	%	1.0
Superheater steam side pressure drop	psi	30
Reheater steam side pressure drop	psi	30
Blowdown, HP evaporator	%	2.0
Blowdown, LP evaporator	%	1.0
High pressure turbine isentropic efficiency		0.920
Intermediate pressure turbine isentropic efficiency		0.920
Low pressure turbine isentropic efficiency		0.890
Make up water temperature	°F	59
Condenser pressure	in. Hg	2.0
HRSG		
Minimum HRSG pinch point	°F	20
HRSG flue gas pressure drop	in. H_2O	7.4
Flue gas temperature leaving HRSG	°F	251
FGD		Greer Limestone used
Limestone Ca/S molar ratio		1.05:1

(continued)

Table 3.2. (continued)

Process Unit-Parameter	**Units**	**Value**
Sulfur removal by wet scrubber FGD	%	95
Limestone slurry solids concentration	wt.%	10
Plant auxiliary power loss	%	3

3.5 Modifications to the METC Model

Several needs for modifications to the existing METC model were identified based upon the review of the model in Section 3.2. These needs are:

- Accounting for the combustor heat loss.
- Incorporating water walls in the combustor.
- Specifying the carbon conversion in the coal combustor.
- Scaling the auxiliary air requirement to the flow rate of coal.
- Accounting for the air leakage in the CerHx.
- Gas turbine compressor and expander outlet pressures and efficiencies.
- Gas turbine cooling air flow circuitry.
- Gas turbine inlet mass flow to represent choked conditions at the expander inlet.
- Steam Injection to Gas turbine.
- Water Injection to Gas Turbine.
- Estimations of NO_x, SO_2, and particulate matter emissions.
- Estimation of auxiliary power consumption based on performance parameters.
- Addition of a slag screen and fabric filter to account for the pressure drops across these units.
- Accounting for net plant efficiency penalty associated with the reheat of flue gas from the FGD unit.
- Incorporating FGD recirculation water and water in the sludge.

In this section, modifications and improvements to the performance model are described.

3.5.1 Auxiliary Air Requirement

Auxiliary air is required to deliver the pulverized coal from the coal handling system to the coal combustor. Therefore, the auxiliary air requirement is proportional to the coal feed rate to the combustor. In the METC model the auxiliary air flow rate is initialized at 144,000 lb/hr, and does not change with a change in the coal feed rate.

Auxiliary air flow rate per lb of coal has been introduced as a new parameter that can be set by the user. The auxiliary air requirement is estimated by the following equation:

$$m_{a,PA,i} = f_{aa}\ m_{cf,CH,i} \tag{3.1}$$

An increase in the value of this parameter would increase the excess air ratio. In the METC base case model, 144,000 lb/hr of auxiliary air was assumed to be required to deliver 181,623 lb/hr of coal. Therefore, 0.79 lb auxiliary air per lb of coal feed rate was required based on the auxiliary air flow and the coal flow rate in the original METC model.

Design specification AUXADJ was added to scale the auxiliary air requirement to the coal feed rate. The initial value of the auxiliary air mass flow rate was varied to achieve the user specified ratio of auxiliary air per lb of coal.

3.5.2 Carbon Conversion in the Combustor

Incomplete combustion of coal in the combustor causes a loss of ignition and a reduction in the carbon converted to products of combustion. In the METC model the carbon conversion was specified in the ASPEN stoichiometric reactor unit operation block COMBSTR which represents the coal combustor. The extent of carbon conversion depends on the combustor design characteristics, and is expected to be different for different combustor designs. Since the coal combustor for the EFCC system is in a developmental stage, the carbon conversion was accessed in a new FORTRAN block called STCTAIL so that it can be specified by the user.

3.5.3 Combustor Radiative Heat Loss

Since the temperature of combustion in the coal combustor is high, radiative heat losses would be expected from the combustor. In the METC model, the heat loss from the combustor was set to zero in the ASPEN design specification CONCOALF. This design specification was modified so that the user can specify the combustor heat loss as a percentage of the combustor heat duty (or the heat of reaction of coal). The mass flow of coal and the higher heating value of coal are accessed in CONCOALF and the combustor heat duty is estimated as their product . The combustor heat loss is then estimated as the user specified fraction of the combustor heat duty.

3.5.4 Combustor Water Walls

A water wall may be incorporated in the coal combustor to cool the combustor walls. A fraction of the water from the economizer outlet in the steam side of the HRSG would be directed through the water walls. Heat is transferred to the water in the water walls, which is then mixed with the economizer outlet stream which enters the high pressure evaporator.

Since the combustor water wall is used to transfer heat from the combustor flue gas to the economizer water, it should be modeled as a HEATER ASPEN unit operation block. In the METC model the water wall was represented as an ASPEN stream class changer block HPBOILC, and therefore did not accurately represent the combustor water wall. The HPBOILC block was changed to an ASPEN HEATER unit operation block. The fraction of economizer outlet water stream which is directed to the combustor water walls is specified by the user in the new ASPEN flow splitter block ST-SPLT4 and its value can be reset by the user in the FORTRAN block STCTAIL. The amount of heat directed to the water walls is specified in the ASPEN flow splitter block HCLAWW and can be reset by the user in the FORTRAN block STCTAIL as a percentage of the combustor heat duty.

The modifications for representing combustor water walls are shown in Figure 3.1 and 3.2, and the unit operation block descriptions are given in Table 3.3 and 3.4.

3.5.5 Air Leakage in the Ceramic Heat Exchanger

A portion of high pressure air leaks from the tube side to the shell side of the CerHx through the tube joint gaskets and due to tube string misalignments as a result of assembly. Previous cycle studies have indicated that for a three percent air leakage rate there is a resulting 1.5 percent loss in cycle efficiency (Orozco and Seger, 1993). Hague International has developed a gasket design specifically for the high temperature operating conditions of the CerHx which reduces the air leakage. Initial tests of prototype tube strings by Hague International showed less than three percent leakage. Tests with production components have shown leakage rates of one percent and in some cases as low as 0.5 percent (Orozco and Seger, 1993). Due to the air leakage in the CerHx, the high pressure air mass throughput to the turbine is reduced, which results in a lower net work output from the gas turbine. The air which leaks into the shell side reduces the temperature of the flue gas entering the HRSG and increases its mass flow rate, which affects the efficiency of the steam cycle and increases the ID fan power requirement. Therefore, air leakage in the CerHx imposes an energy penalty and reduces the plant efficiency.

Air leakage from the CerHx tubes is simulated in the performance model by splitting the high pressure air flow from the compressor using ASPEN flow splitter unit operation blocks. The air leakage stream and the flue gas stream are mixed using ASPEN mixer unit operation blocks. As an approximation to air leakage occurring internally in the CerHx, fifty percent of the leakage is assumed to occur from the high pressure air stream entering the CerHx, which mixes with the flue gas exit stream exiting the CerHx. The rest of the leakage is assumed to occur from the high pressure exit air stream from the CerHx, which mixes with the flue gas inlet stream to the CerHx before entering the CerHx.

These modifications are represented in Figures 3.1 and 3.3, and Tables 3.3 and 3.5 lists the associated unit operation block descriptions.

3.5.6 Gas Turbine Compressor and Expander Outlet Pressures and Efficiencies

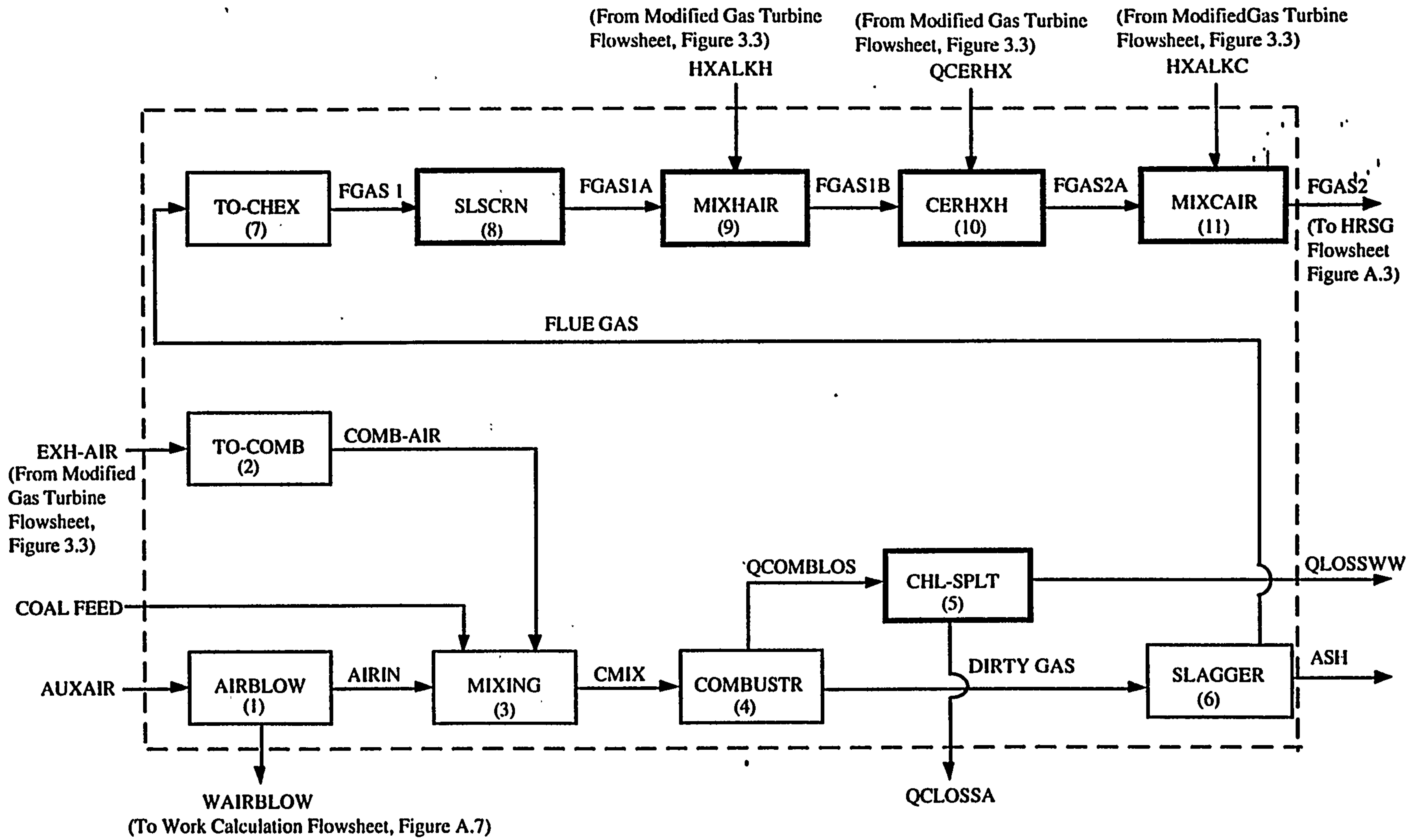
The gas turbine model has been calibrated to a GE Frame 7F firing natural gas. Atmospheric air at ambient conditions enters the gas turbine compressor. The pressure drop across the compressor inlet is specified to be 0.145 psi. The air pressure at the first stage compressor inlet is obtained by subtracting the pressure drop from the ambient air pressure. The outlet pressure at each 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. The individual compressor stage outlet pressures for the first, second, and third stages are estimated by the following relationships respectively:

$$P_{outlet} = P_{inlet}\ (PR)^{0.33} \tag{3.2}$$

$$P_{outlet} = P_{inlet}\ (PR)^{0.667} \tag{3.3}$$

$$P_{outlet} = P_{inlet}\ (PR) \tag{3.4}$$

The outlet stream pressure of each turbine expander stage is also estimated in the FORTRAN block STCTAIL. The outlet pressure of first stage turbine expander is set equal to the second stage compressor discharge pressure. The outlet pressure of second stage turbine expander is set equal to the first stage compressor discharge pressure. The third stage turbine expander outlet pressure is set to 15.2 psia. The compressor isentropic efficiencies are specified as 88 percent and the expander isentropic efficiencies are specified as 89 percent. These values can be reset by the user in the FORTRAN block STCTAIL. Figure 3.3 shows the process flow diagram of the modified gas turbine and Table 3.6 lists the associated unit operation block descriptions.



Note: Boxes outlined in bold indicate newly added unit operation blocks.

Figure 3.1. Modified Combustor Flowsheet

Table 3.3. Modified Combustor Section Unit Operation Block Description[a]

No[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
1	AIRBLOW (COMPR)	TYPE = 3 Pressure = 16.4 psia Isentropic Efficiency = 0.72	Air blower for the auxiliary air required for pneumatic transport of coal to combustor. It increases the pressure of the air from 14.7 psia to 16.4 psia.
2	TO-COMB (CLCHNG)		Changes turbine exhaust air stream class from CONVENTIONAL to MIXNC.
3	MIXING (MIXER)	Pressure = 15.13 psia	Sets the pressure for all combustor feed streams to 15.13 psia. Mixes auxiliary air, combustion air, and coal.
4	COMBUSTR (RSTOIC)	Pressure = -0.11 psia Temperature = 2750 °F NPK = 1	The stoichiometric combustor simulates stoichiometric combustion reaction of coal. The combustion temperature is initially set at 2,750 °F, and is later reset by the FORTRAN block EFFECT based on the temperature of the flue gas required to produce the specified turbine inlet temperature T_i.
5	CHL-SPLT (FSPLIT)	FRAC QCLOSSWW = 0.5	The fraction of combustor heat duty which is directed to the water walls is set in this block as a fraction of the combustor heat duty.
6	SLAGGER (SEP 2)	SUBS-FRAC MIXED FLUEGAS 1.0 / NC FLUEGAS 0.0 / NC ASH 1.0	Used to simulate the separation of ash from the flue gas. The fracton of gas in the combustor exit stream is specified to be 1.0 in the flue gas stream and the fractions of ash in the combustor exit stream is specified to be 1.0 in the ash stream.
7	TO-CHEX (CLCHNG)		Changes the flue gas stream class from MIXNC to CONVENTIONAL.
8	SLSCRN (MIXER)	Pressure = -0.356 psi NPK = 1 KPH = 1	Simulates the fluegas pressure drop across the slag screen.

(continued)

Table 3.3. (continued)

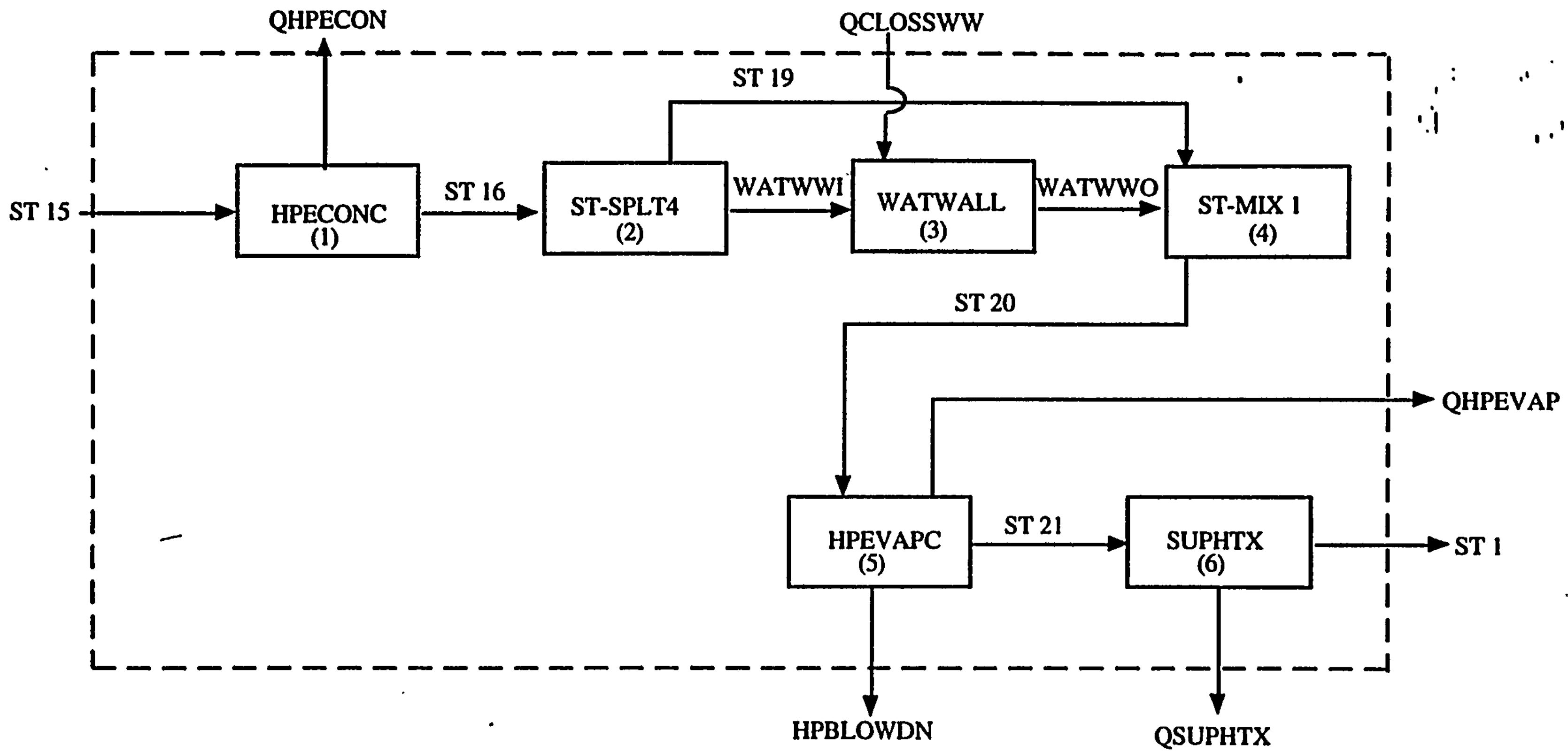
NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
9	MIXHAIR (MIXER)	Pres = -0.0 NPK = 1 KPH = 1	Mixes the air leakage from the hot end of the high pressure tube side of the CerHx with the fluegas.
10	CERHXH (HEATER)	Pressure = -0.76 psia NPK = 1 KPH = 1	Hot side of the ceramic heat exchanger, which transfers heat to the incoming compressed cool air stream. The pressure drop is specified to be 0.76 psi. The heat duty is estimated from the cold side of the ceramic heat exchanger block (CERHXC) in the gas turbine flowsheet.
11	MIXCAIR (MIXER)	Pres = -0.0 NPK = 1 KPH = 1	Mixes the airleak from the cold end of the tube of the CerHx with the fluegas.

[a] Block number 5 has been added to the original METC performance model to represent combustor water wall, block number 8 has been added to represent the slag screen, and blocks 9 and 11 have been added to represent ceramic heat exchanger air leakage.

[b] Numbers for each unit operation block correspond to those in Figure 3.1.

[c] 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 B.1 in Appendix B.

[d] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.



Note: Boxes outlined in bold indicate newly added unit operation blocks.

Figure 3.2. Modified HRSG (steam side) Flowsheet

Table 3.4. Modified Heat Recovery Steam Generator (Steam-Side) Section Unit Operation Block Description[a]

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
1	HPECONC (HEATER)	Pressure = -85.0 psia Temperature = 563.2 °F	The pressure drop and the outlet temperature in the cold (steam) side of the high pressure economizer are specified to be 85 psi and 563.2 °F respectively.
2	ST-SPLT4 (FSPLIT)	FRAC WATWWI = 0.3	Splits the water from the high pressure economizer. A negligible fraction is specified to be directed to combustor water walls the value of which is reset in the FORTRAN block STCTAIL.
3	WATWALL (HEATER)	PARAM = -0.0	Represents the water walls of the coal combustor. Water from economizer is heated in the coal combustor water walls.
4	ST-MIX1 (MIXER)	Pressure = 0.0 psia	Combines the stream from the combustor water wall and the high pressure economizer.
5	HPEVAPC (FLASH 2)	Pressure = -100.0 psia V = 0.98	Steam side of the high pressure evaporator. Flashes the inlet liquid stream. 98% of the outlet stream is specified to be vapor which goes to the superheater, and the rest is blowdown. The pressure drop is specified to be 100 psi.
6	SUPHTXC (HEATER)	Pressure = -30.0 psia Temperature 1050 °F NPK = 1 KPH = 1	Steam side of the superheater. The outlet temperature and the pressure drop are specified to be 1050 °F and 30 psi respectively.

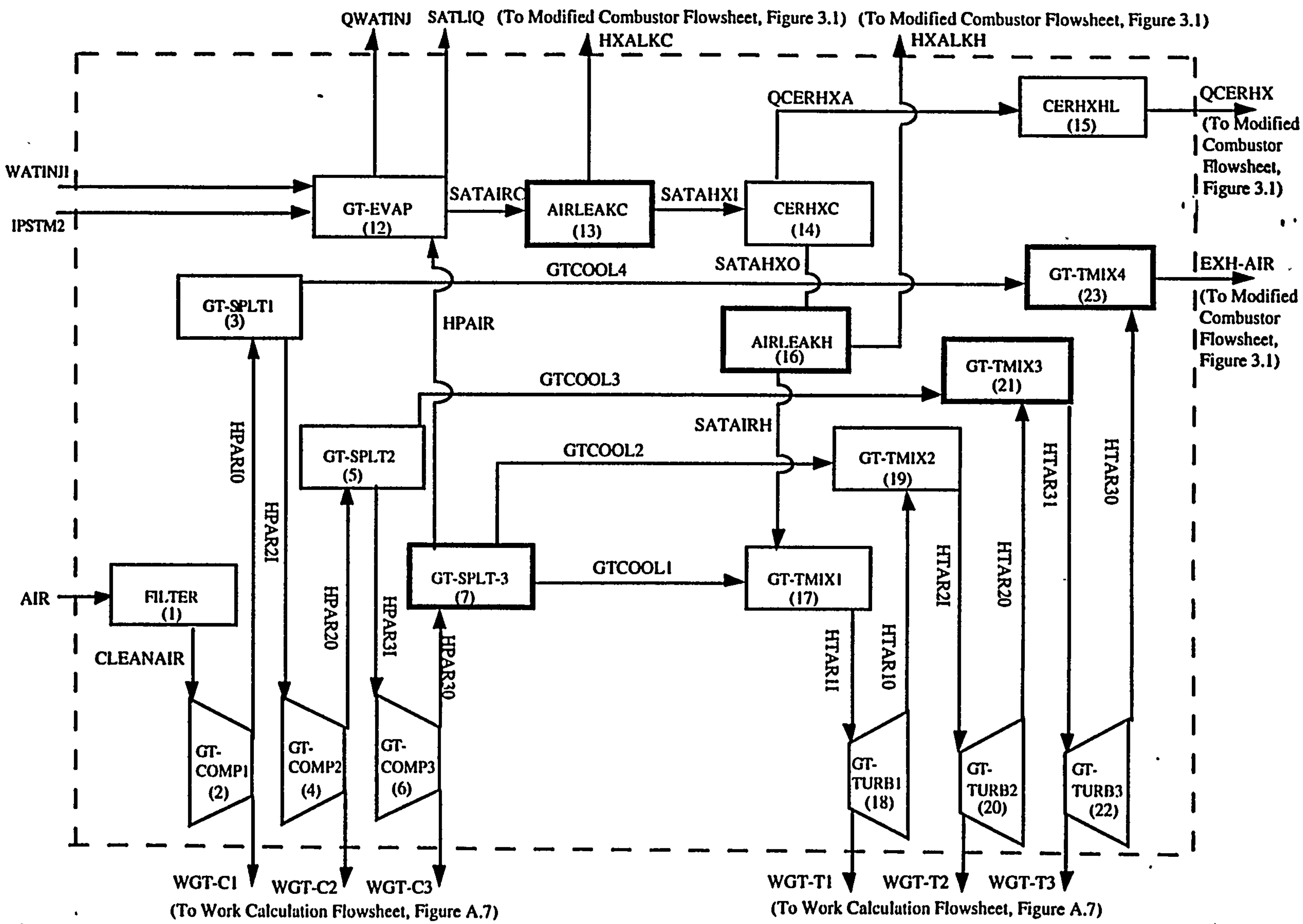
(continued)

[a] Block number 4 has been modified to represent the combustor water wall.

[b] Numbers for each unit operation block correspond to those in Figure 3.2.

[c] 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 B.1 in Appendix B.

[d] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.



Note: Boxes outlined in bold indicate newly added unit operation blocks.

Figure 3.3(a). Modified Gas Turbine Flowsheet

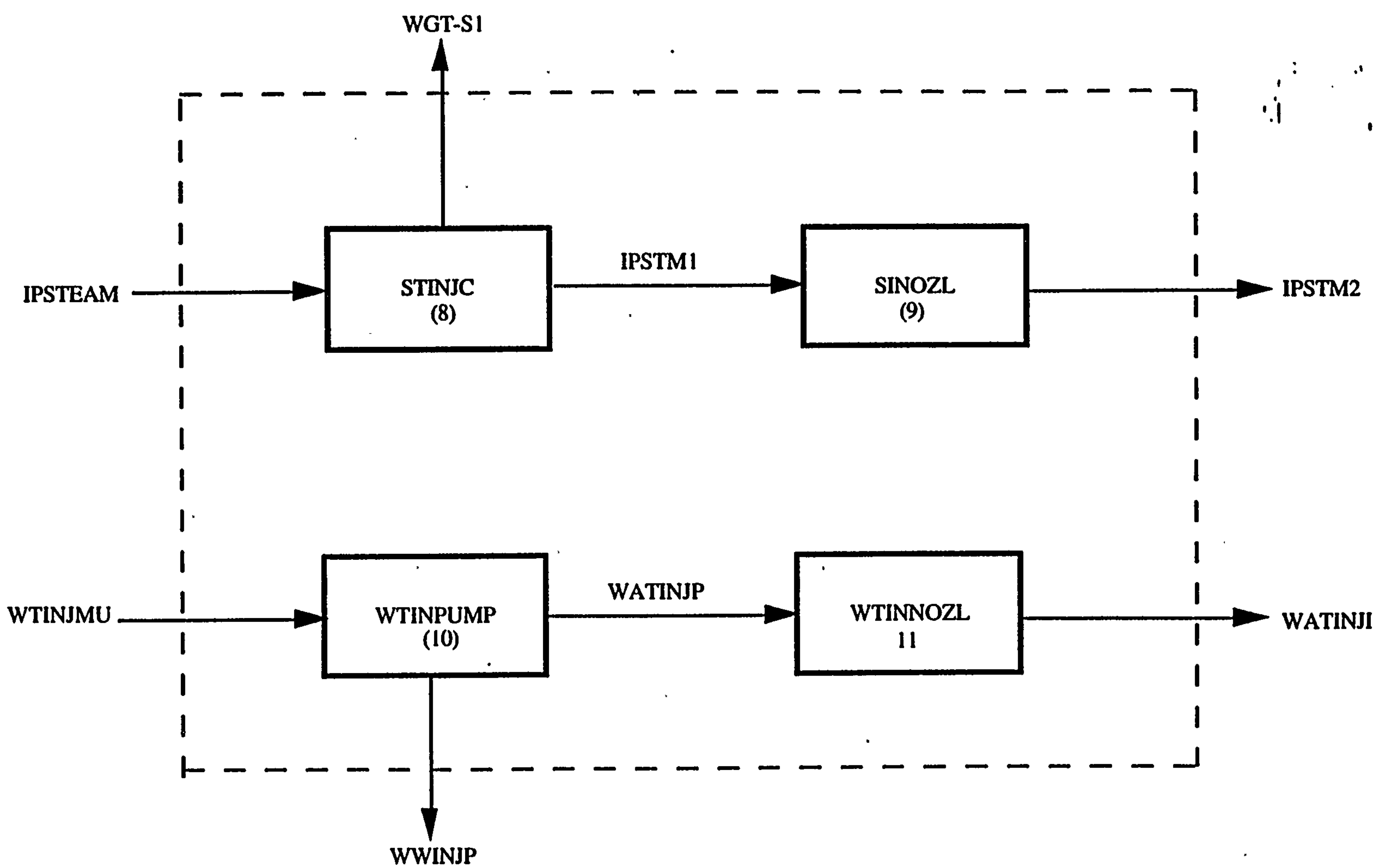


Figure 3.3(b). Modified Gas Turbine Flowsheet

Table 3.5. Modified Gas Turbine Section Unit Operation Block Description[a]

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
1	FILTER (MIXER)	Pressure = -0.5 psi NPK = 1 KPH = 1	Simulates the pressure drop across the air filter which is specified to be 0.5 psi.
2	GT-COMP 1 (COMPR)	TYPE = 3 Pressure = 52.6 psia Isentropic Efficiency = 0.92 GFLAG = 1	Simulates an isentropic compressor. The outlet pressure is specified to be 52.5 psia. The outlet pressure and isentropic efficiency is reset in the FORTRAN block STCTAIL.
3	GT-SPLT1 (FSPLIT)	FRAC GTCOOL4 0.03	Specifies the fraction of the cooling flow split GTCOOL4 to be three percent of the inlet air flow.
4	GT-COMP 2 (COMPR)	TYPE = 3 Pressure = 81.3 psia Isentropic Efficiency = 0.92 GFLAG = 1	Simulates an isentropic compressor. The outlet pressure is specified to be 81.3 psia. The outlet pressure and isentropic efficiency is reset in the FORTRAN block STCTAIL.
5	GT-SPLT2 (FSPLIT)	FRAC GTCOOL3 0.03	Specifies the fraction of the cooling flow split GTCOOL3 to be three percent of the inlet air flow.
6	GT-COMP-3 (COMPR)	TYPE = 3 Pressure = 198.5 Isentropic Efficiency = 0.92 GFLAG = 1	Simulates an isentropic compressor. The outlet pressure is specified to be 198.5 psia. The outlet pressure and isentropic efficiency is reset in the FORTRAN block STCTAIL.
7	GT-SPLT3 (FSPLIT)	FRAC GTCOOL2 0.06 FRAC GTCOOL1 0.06	Specifies the fraction of the cooling flow split GTCOOL2 and GTCOOL1 to be six percent of the inlet air flow each.

(continued)

Table 3.5. (continued)

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
8	STINJC (COMPR)	TYPE = 3 Pressure = 220.0 psia Isentropic Efficiency = 0.88 GFLAG = 1	Simulates an isentropic compressor. The outlet pressure and the isentropic efficiency are specifed to be 220.0 psia and 88 percent, and are reset in the FORTRAN block STCTAIL.
9	SINOZL (MIXER)	PARAM PRES=-0.0	This block is used to specify the pressure drop in the steam flow across the GT-EVAP flash block. The pressure drop is reset in the FORTRAN block STCTAIL.
10	WTINPUMP (PUMP)	PARAM PRES=300.0 TYPE=1	This block is used to pump water at ambient conditions after treatment for water injection into the gas turbine.
11	WTINNOZL (MIXER)	PARAM PRES=-100.0	This block represents the pressure drop across the inlet nozzle for water injection into the gas turbine evaporator.
12	GT-EVAP (FLASH2)	Pressure = -0.0	This block is used to add steam and water to the compressor discharge air stream before it enters the ceramic heat exchanger. The steam and the water flow rate can be specified by the user to represent different levels of steam and water injection for power augmentation. The pressure drop is assumed to be 0 psi.
13	AIRLEAKC (SSPLIT)	FRAC MIXED HXALKC 0.0025	This block simulates the airleak in the CerHx at the cold end of the tube side. 0.25 percent air is assumed to leak from this end. The fraction of air leakage can be reset in the FORTRAN block STCTAIL.

(continued)

Table 3.5. (continued)

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
14	CERHXC (HEATER)	Pressure = -3.5 psia Temperature = 2435 °F	Cold side of the ceramic heat exchanger, where compressed air is heated for expansion in gas turbine. The pressure drop is specified to be 3.5 psi. The outlet air temperature is specified to be 2435 °F, which is later reset by the DESIGN-SPEC CONTIT to achieve the desired T_i.
15	CERHXHL (MULT)	Factor = 1.02	Accounts for the assumed two percent ceramic heat exchanger heat loss based on the ceramic heat exchanger heat duty. The actual heat loss is speciifed in the FORTRAN block STCTAIL.
16	AIRLEAKH (SSPLIT)	FRAC MIXED HXALKH 0.0025	This block simulates the airleak in the CerHx at the hot end of the tube side. 0.25 percent air is assumed to leak from this end. The fraction of air leakage can be reset in the FORTRAN block STCTAIL.
17	GT-TMIX 1 (MIXER)	Pressure = -0 psia	Mixes the first stage turbine inlet air with the cooling air flow GTCOOL1. The pressure drop across this block is assumed to be zero.
18	GT-TURB 1 (COMPR)	TYPE = 3 Pressure = 83.0 psia Isentropic Efficiency = 0.913 GFLAG = 1	Simulates an isentropic turbine expander. The outlet pressure is specified to be 83.0 psia. The isentropic efficiency is set in the FORTRAN block STCTAIL.
19	GT-TMIX2 (MIXER)	Pressure = -0 psia	Mixes the first stage turbine outlet air with the cooling air flow GTCOOL2. The pressure drop across this block is assumed to be zero.

(continued)

Table 3.5. (continued)

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
20	GT-TURB 2 (COMPR)	TYPE = 3 Pressure = 54.0 psia Isentropic Efficiency = 0.913 GFLAG = 1	Simulates an isentropic turbine expander. The outlet pressure is specified to be 54.0 psia. The isentropic efficiency is set in the FORTRAN block STCTAIL.
21	GT-TMIX3 (MIXER)	Pressure = -0 psia	Mixes the second stage turbine outlet air with the cooling air flow GTCOOL3. The pressure drop across this block is assumed to be zero.
22	GT-TURB 3 (COMPR)	TYPE = 3 Pressure = 15.2 psia Isentropic Efficiency = 0.913 GFLAG = 1	Simulates an isentropic turbine expander. The outlet pressure is specified to be 15.2 psia. The isentropic efficiency is set in the FORTRAN block STCTAIL.
23	GT-TMIX 4 (MIXER)	Pressure = 15.13 psia	Mixes the turbine exhaust and the cooling air flow GTCOOL4. The combined stream forms the oxidation air stream to the coal combustor. The outlet pressure is specified to be 15.13 psia.

[a] Block numbers 3, 5, 7, 17, 19, 21 and 23 have been added to the original METC performance model to represent the.turbine cooling flow circuitry; block numbers 8 and9 have been added and block number 12 has been modified to represent steam injection; block numbers 10 and 11 have been added to represent water injection; and block numbers 11 and 14 have been added to represent ceramic heat exchanger air leakage.

[a] Numbers for each unit operation block correspond to those in Figure 3.3a and Figure 3.3b.

[c] 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 B.1 in Appendix B.

[d] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

3.5.7 Gas Turbine Cooling Air

In a gas turbine the cooling flows would be extracted from the discharge at multiple compresser stages (Frey and Rubin, 1991). Therefore, the METC model has been modified to simulate cooling extraction streams from multiple compressor stage discharges which in turn leads to improved characterization of the energy penalty associated with cooling air. As indicated in Figure 3.3 and Table 3.5, a portion of the total inlet air flow to the compressor is directed to the first and second stage turbine inlets from the third stage compressor discharge. In addition, a percentage of the inlet air flow to the compressor is directed to the third stage turbine inlet from the second stage compressor discharge, and a portion of the the compressor air bypasses the turbine and mixes with the turbine exhaust. The values given in Table 3.5 for the cooling air percentages were estimated by calibrating the model to the overall efficiency and output specifications for a typical heavy duty gas turbine. ASPEN flow splitter unit operations blocks were used to split the compressor discharge flows and ASPEN mixer unit operation blocks were used to mix the inlet streams to the turbines. The cooling air stream fractions are specified in the FORTRAN block AIRCOOL.

3.5.8 Turbine Inlet Mass Flow

The METC gas turbine model was developed based upon an assumed constant compressor inlet air mass flow, irrespective of changes in gas turbine mass flow associated with steam injection. However, the overall mass flow in a gas turbine is more typically limited by the turbine inlet nozzle and not by the compressor inlet, as previously discussed in Section 2.3.4. Therefore, the gas turbine model was modified to account for the mass flow constraint at the turbine inlet, due to choking. This modification enables the model to more accurately respond to the effect of the steam injection. As steam is injected, the compressor mass flow must be reduced to compensate for the constraint on overall turbine mass flow.

The flow at the inlet of the gas turbine expander is choked (i.e., the Mach number of the gas stream is unity). This is true regardless of the fuel type because the pressure ratio across the first stage turbine nozzle is large enough to cause choking (Eustis and Johnson, 1990). An equation to estimate the mass flow of a hot air at the gas turbine expander inlet at choked condition based on a reference fuel is given by:

$$m_{a,GT,i,choked} = m_{ref,GT,i} \frac{\frac{P_{a,GT,i}}{P_{ref,GT,i}} \sqrt{\frac{mw_{a,GT,i}}{mw_{ref,GT,i}}}}{\sqrt{\frac{T_{a,GT,i}}{T_{ref,GT,i}}}} \tag{3.5}$$

A mixture of air and natural gas was considered as the reference fuel based upon the performance specifications for heavy duty gas turbines. Design specification TCHOKE was added to set the flow of hot air at the turbine inlet nozzle corresponding to choked flow conditions by varying the compressor inlet air flow.

3.5.9 Steam Injection to Gas turbine

In the METC model steam was injected as a fraction of the exit steam flow from the intermediate pressure steam turbine. The steam was mixed with the gas turbine compressor outlet air stream in the ASPEN mixer unit operation block GT-EVAP. GT-EVAP was changed from a mixer unit operation block to an ASPEN flash unit operation block for including water injection as described in the following section. The pressure of the steam injection stream to the GT-EVAP block was assumed to be higher than the pressure of the compressor outlet air to account for pressure drops across valves and piping. Furthermore, the pressure drop of the steam at the inlet nozzle to the GT-EVAP block was not accounted for. Modifications were made to the METC model to raise the pressure of the steam injection stream to the pressure of the compressed air for the case where the steam pressure was lower than the compressed air pressure and account for the pressure drop in the steam flow at the inlet nozzle.

In the modified model the steam mass flow from the intermediate pressure steam turbine to the gas turbine is specified in the FORTRAN block STCTAIL. An ASPEN compressor unit operation block is used to raise the pressure of the steam to the sum of the pressure of the gas turbine compressor discharge air and a user specified pressure drop across the inlet nozzle to the GT-EVAP block. An ASPEN mixer unit operation block was added to account for the pressure drop across the steam injection inlet nozzle. If the steam pressure is lower than the sum of the compressor discharge air pressure and the steam injection inlet nozzle pressure drop, then the steam pressure at the compressor outlet is set to be equal to the sum of the compressor outlet air pressure and the steam injection inlet nozzle pressure drop.

Since the steam cycle is a closed cycle, the mass flow of steam injected to the gas turbine should be compensated by increasing the makeup water in the steam cycle to compensate for the steam extracted from the steam cycle. In the original METC model no provisions were made to increase the steam cycle makeup water corresponding to the mass flow of steam injected. A new stream called SCINJMU was added to the set of streams representing losses in the steam cycle. The mass flow of this stream was initialized to a value of 250,000 lb/hr at ambient conditions. The mass flow rate of this stream is reset to the mass flow rate of the steam injection rate in the new FORTRAN block STCTAIL.

Figures 3.3a and 3.3b shows the gas turbine process flow diagram modifications for steam injection and Table 3.5 lists the associated unit operation block descriptions.

3.5.10 Water Injection to Gas Turbine

Injecting water to the CerHx inlet air stream is expected to raise the net plant power output and plant efficiency (Parsons and Bechtel, 1991). The METC model did not incorporate water injection to the gas turbine. Modifications were made to include the option of water injection design in the EFCC system.

Modifications were made to the EFCC performance model to incorporate water injection at ambient conditions into the gas turbine compressor discharge air. The water injection rate was initialized to a small number at ambient conditions with the stream WTINJMU. The water injection rate can be specified and reset in the new FORTRAN block STCTAIL. An ASPEN pump unit operation block was added to represent the pump required to boost the pressure of the water injected and an ASPEN mixer block was added to account for the pressure drop across the water injection nozzle. Depending on the gas turbine compressor outlet air pressure, the discharge pressure of the water injection pump is reset in the FORTRAN block STCTAIL. If the specified discharge pressure of the water injection pump is less than the gas turbine compressor outlet pressure plus the pressure drop across the water injection inlet nozzle, then the water injection pump discharge pressure is set equal to the sum of the compressor outlet pressure and the pressure drop across the water injection inlet nozzle. Since the water used for water injection is treated similar to the boiler feed water, the mas flow of water injection is added together with the makeup water of the steam cycle in the cost subroutine. The water injection stream is mixed with the gas turbine compressor discharge air in the GT-EVAP block. The GT-EVAP block was changed from a mixer unit operation block to an ASPEN flash unit operation block to allow for injected water to flash to water vapor. The flash operation was specified as adiabatic.

Figures 3.3a and 3.3b show the process flow diagram modifications for including water injection, and Table 3.5 lists the associated unit operations block description.

3.5.11 Estimation of NO_X Emissions

In the base case METC model generation of NO and NO_2 (NO_X) has not been modeled in the coal combustor unit operation block (RSTOIC). The current NSPS NO_x emission standard for large coal fired power plants is 0.5 lb/10^6 Btu coal input. The development target for NO_X emissions in the EFCC system is 0.1/10^6 Btu coal input (Orozco and Seger, 1993). For a conservative estimate of NO_X emissions in EFCC systems, the NO_x emissions were assumed to be 0.5 lb/10^6 Btu coal input. Therefore, the amount of NO_X generated is estimated by the following equation:

$$m_{NO_x,C,o} = 0.5\ Q_{coal} \tag{3.6}$$

If the fraction of NO in the total NO_X is f_{NO} and the fraction of NO_2 is f_{NO2} on a mass basis, then the emission rate of NO and NO_2 can then be estimated by the following equations:

$$m_{NO,C,o} = f_{NO}\, m_{NO_x,C,o} \tag{3.7}$$

$$m_{NO_2,C,o} = f_{NO_2}\, m_{NO_x,C,o} \tag{3.8}$$

3.5.12 Estimation of Particulate Matter Emissions

In the METC model the ash generated by coal combustion is assumed to be completely removed as slag. While most of the coal ash is expected to be removed as slag, a portion of the ash will be carried as flyash by the flue gas exiting the combustor. The flyash would have a continous distribution of size ranges, and would be removed with various size-dependent removal efficiencies in the slag screen, CerHx, and the fabric filter.

For the purpose of particulate matter emissions estimation, the flyash size distribution has been discretized in three size ranges: low, medium, and high. The percent of ash removed as slag, the ash size distribution, and the particulate removal efficiencies of the slag screen, CerHx, and the fabric filter, have all been parameterized. The values for these parameters can be set by the user depending on the particular process equipment used in the EFCC system. The particulate matter emissions are reported on dry, 15 percent O_2 basis.

3.5.13 Estimation of SO_2 Emissions

A wet limestone FGD process with forced oxidation was incorporated in the METC model for removal of sulfur dioxide from flue gas. In the METC model 95 percent removal efficiency of SO_2 was assumed by specifying 95 percent conversion of sulfur dioxide in the ASPEN FGD reactor unit operation block FGDRXR. Since the removal efficiency of the FGD process may be different for different FGD designs, the removal efficiency has been parameterized. The SO_2 conversion in the FGD reactor can be specified in the FORTRAN block STCTAIL. Therefore, the FGD sulfur removal efficiency can be set corresponding to the technology employed. A 90 percent removal efficiency was assumed in the base case simulation based on an EPRI FGD design of a wet limestone forced oxidation system (EPRI, 1991). The sulfur dioxide concentration in the stack gas is accessed and reported as the EFCC plant SO_2 emissions on a pound per million BTU basis.

3.5.14 Auxiliary Power Estimation

In the METC performance model the auxiliary power requirement was assumed to be three percent of the net electrical power output, for which no basis was cited. The auxiliary power requirement is expected to change with a change in the process performance variables of the EFCC system. To estimate a more realistic auxiliary power consumption, the power consumed by individual process sections has been estimated. The details of new models for estimating the auxiliary power consumption are given in Chapter 4.

3.5.15 Flue Gas Pressure Drops

The pressure drops along the air flow path in the gas turbine and flue gas flow path in other process sections have been set in the FORTRAN block STCTAIL. In addition to the pressure drop parameters included in the METC model, pressure drops along the flue gas path in the slag screen and the fabric filter have also been parameterized and can be specified in the FORTRAN block STCTAIL. ASPEN mixer unit operation blocks are used to represent these pressure drops in the performance model. The outlet flue gas stream pressure from the low pressure evaporator had been set to 14.12 psia in the METC model. This specification was changed to a pressure drop of 0.028 psi. All pressure drop specifications are shown in Table 3.14.

The flowsheet modification to include the slag screen and fabric filter pressure drops are shown in Figures 3.1 and 3.4 respectively, and the associated listing of unit operation blocks are shown in Tables 3.3 and 3.6 respectively.

3.5.16 FGD Recirculation Water and Sludge Water

The METC performance model does not account for recirculation water to the slurry feed stream and water in the sludge. Modifications have been made to include the water recirculated to the limestone slurry feed from the FGD reactor exit stream and the water in the sludge using two ASPEN separator unit operation blocks. The first separator block, FGDWAT, sets the mass flow of water in the recirculation stream RECWAT to a negligibly small number. The second separator block, FGDSEP, sets the mass flow of water in the sludge to a negligibly small number, the fraction of solids in the sludge stream to one, and the fraction of all gaseous components in the sludge to be zero. A FORTRAN block, SLWATER, was added to the model to reset the water in the sludge and the water recirculated to the limestone slurry feed. In this FORTRAN block the liquid water exiting the FGD reactor is accessed and the water in the sludge is specified as a fraction of dry sludge flow rate. The portion of water which does not exit with the sludge is then set as the mass flow rate of the recirculation water stream. Thus the flue gas exiting the FGD system contains no liquid water.

Figure 3.4 shows the process flow diagram modifications for the recirculation water and sludge water, and Table 3.6 lists the associated unit operation block descriptions.

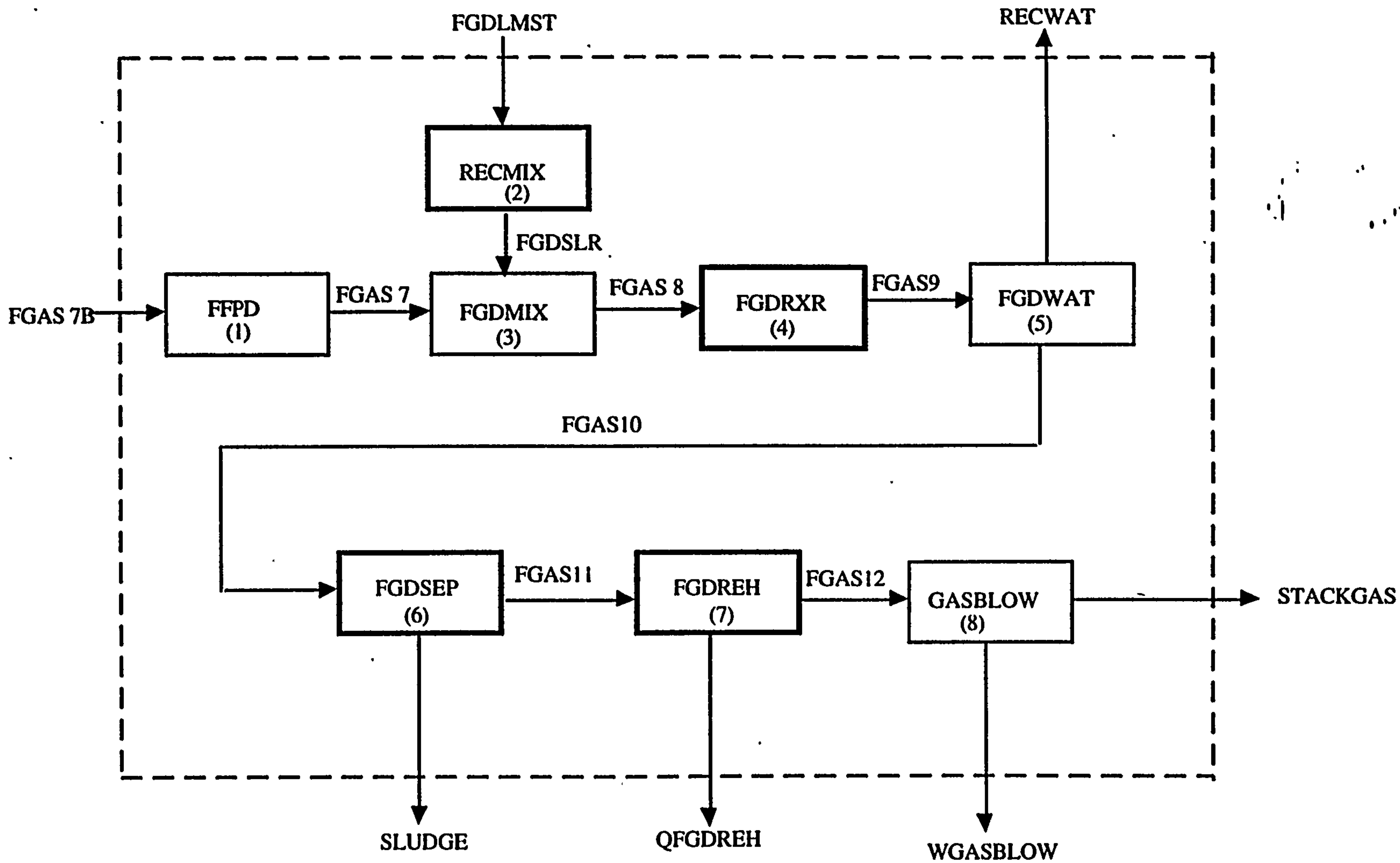
3.5.17 FGD Flue Gas Reheat

The flue gas exits the FGD unit in the METC model at 118 ^oF. This temperature is not high enough to give the stack gas sufficient buoyancy for dispersion. Therefore, the temperature of the flue gas needs to be raised. This is done by heating it using low quality steam from the steam cycle. Steam usage for FGD reheat imposes an energy penalty on the system. FGD reheat was not included in the original METC model.

The temperature of the flue gas exiting the stack is higher than the temperature of the gas exiting the FGD reheater due to energy input from the plant induced draft fan. Therefore, design specification block, SGTEMP, was added where the stack gas temperature can be specified by the user. The temperature of the flue gas exiting the FGD reheater is then varied to achieve the specified stack gas temperature taking into account the temperature increase across the ID fan. An ASPEN heater unit operation block was added in the performance model to estimate the heat duty of the FGD reheater. Steam is extracted from the input stream to the second low pressure steam turbine stage for providing the heat input for the FGD flue gas. The enthalpy required for the FGD flue gas reheat is subtracted from the enthalpy of the stream entering the second low pressure steam turbine stage. The work output of this turbine stage is then scaled down in proportion to the reduction in the mass flow rate of steam input, which represents the efficiency penalty of the steam cycle. The gross power output of the steam cycle and the new plant efficiency is then estimated.

3.5.18 Process Parameter Assumptions

Some of the process parameters assumed for the modified METC model were different from those assumed for the original METC model, and some additional parameter assumptions were introduced, as previously described and documented in Table 3.3 through 3.6. These changes and additions to the original process parameter assumptions are listed in Table 3.7



Note: Boxes outlined in bold indicate newly added unit operation blocks.

Figure 3.4. Modified Flue Gas Desulfurization Flowsheet

Table 3.6. **Modified Flue Gas Desulfurization Section Unit Operation Block Description**[a]

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
1	FFPD (MIXER)	Pressure = -0.27 psi NPK = 1 KPH = 1	Simulates the pressure drop across the fabric filter which is 0.27 psi/
2	RECMIX (MIXER)	PARAM PRES=-0.0	This block is used to mix the inlet limestone slurry feed and water recirculated from the FGD reactor outlet stream.
3	FGDMIX (MIXER)	Pressure = -0.2 psi	Mixes the limestone slurry feed with the HRSG exhaust flue gas. The pressure drop is specified to be 0.2 psi.
4	FGDRXR (RSTOIC)	Pressure = -0.02 psi DUTY = 0	Represents a stoichiometric reactor for the flue gas desulfurization reaction. The pressure drop and the heat duty is specified to be 0.02 psi and 0 respectively. The conversion of SO_2 in the flue gas is specified to be 95%.
5	FGDWAT (SEP)	MASS-FLOW SUBS=MIXED STRM=RECWAT COMP=H2O FLOW=1.0D-9	This block of used to set the mass flow of water recirculated from the FGD reactor outlet stream to the limestone feed stream. The actual mass flow is calculated in the FORTRAN block SLWATER. The fraction of all solids components and all gaseous components other than water are specified to be zero for the recirculaton water stream.
6	FGDSEP (SEP)	MASS-FLOW SUBS=MIXED STRM=SLUDGE COMP=H2O FLOW=1.0D-9	This block is used to set the mas flow of water in the sludge exiting the FGD unit. The actual mass flow is esimated based on user specified water content in the sludge in the FORTRAN block SLWATER The fracton of all solids components in the FGD reactor outlet stream is set to one for the sludge stream

(continued)

Table 3.6. (continued)

NO[b]	BLOCK ID[c] (ASPEN-BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
7	FGDREH (HEATER)	Temperature = 181.0 °F	The temperature of the flue gas exiting the FGD reheater is set in this block and its value of adjusted for a specified stack gas temperature in the DESIGN-SPEC SGTEMP. Theheat duty for the reheat.is calculated by this block.
8	GASBLOW (COMPR)	Pressure = 14.7 psia TYPE = 3 Isentropic Efficiency = 0.85	Represents an Induced Draft (ID) fan, which is modeled as an isentropic compressor with a specified efficiency of 85%. The pressure of the inlet flue gas stream is raised to 14.7 psia.

[a] Block number 1 has been added to the original METC performance model to represent the pressure drop across the fabric filter; block number 2 has been added to represent the mixing of recirculation water with the inlet limestone slurry feed, block number 5 has been added to represent separation water for circulation to the limestone slurry feed, block number 6 has been added to represent the separation of water from the FGD reactor exit stream which is carried with the sludge, and block number 7 has been added to represent FGD reheater for flue gas reheat.

[b] Numbers for each unit operation block correspond to those in Figure 3.4.

[c] 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 B.1 in Appendix B.

[d] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

Table 3.7. **Summary of Base Case Parameters Values for New EFCC Performance Model**

Process Unit Parameter	**Units**	**Value**
Gas Turbine		
Pressure drop across the compressor inlet ducting	psi	0.145
Compressor pressure ratio		13.5
Compressor isentropic efficiency	%	88
Expander isentropic efficiency	%	89
Generator efficiency	%	98.6
Turbine Inlet Temperature	°F	2,300
Total Cooling requirement	% compressor inlet air flow	18
Cooling flow which expands in the turbine and contributes to power generation	% of total cooling requirement	83.33
Cooling flow which bypasses the turbine and does not contribute to power generation	% of total cooling requirement	16.67
Coal Combustor		
Higher Heating Value of coal, dry basis	Btu/lb	12,774
Auxiliary Air flow rate	lb/lb coal	0.79
Radiative Heat Loss	% of heat duty	0.5
Heat directed to combustor water wall	% of heat duty	0.0
Carbon Conversion	%	99.0
Pressure drop	psi	0.11

(continued)

Table 3.7. (continued)

Process Unit-Parameter	**Units**	**Value**
Percent of ash in coal		10.0
Percent of coal ash removed as slag		65.0
Percent of flyash in small particle size range		25.0
Percent of flyash in medium particle size range		50.0
Percent of flyash in high particle size range		25.0
Ceramic Heat Exchanger (including slag screen)		
Pressure Drop across the slag screen	psi	0.356
Particle removal efficiency of slag screen for particles in the low size range	% removal	0.1
Particle removal efficiency of slag screen for particles in the medium size range	% removal	99.0
Particle removal efficiency of slag screen for particles in the high size range	% removal	0.1
Effectiveness		0.85
Heat loss	%	2.0
Air leakage from tube side to shell side	% of inlet air mass flow	0.5
Flue Gas side pressure drop	psi	1.116
Turbine Air side pressure drop	psi	3.50
Particle removal efficiency of CerHx for particles in the low size range	% removal	0.1
Particle removal efficiency of CerHx for particles in the medium size range	% removal	0.1

(continued)

Table 3.7. (continued)

Process Unit Parameter	**Units**	**Value**
Particle removal efficiency of CerHx for particles in the high size range	% removal	99.0
Steam Cycle (1785 psia/1050 °F/1050 °F)		
Steam Turbine Generator efficiency	%	98.6
Deaerator Vent loss	%	1.0
Deaerator pressure	psia	17.5
Bottoming Cycle heat loss	%	1.0
Superheater steam side pressure drop	psi	30
Reheater steam side pressure drop	psi	30
Blowdown, HP evaporator	%	2.0
Blowdown, LP evaporator	%	1.0
High pressure turbine isentropic efficiency		0.92
Intermediate pressure turbine isentropic efficiency		0.92
Low pressure turbine isentropic efficiency		0.89
Steam injection compressor efficiency	%	88
Steam injection electric motor efficiency	%	95
Make up water temperature	°F	59
Condenser pressure	in. Hg	2.0
HRSG		
Minimum HRSG pinch point	°F	20
HRSG flue gas pressure drop	psi	0.14

(Continued)

Table 3.7. (continued)

Process Unit-Parameter	Units	Value
Flue gas temperature leaving HRSG	°F	181
FGD		
Limestone Ca/S molar ratio		1.05:1
Limestone slurry solids concentration	wt.%	10
Flue gas pressure drop	psi	0.22
Solids percent in sludge	weight %	45
Sulfur removal by wet scrubber FGD	%	90
Fabric Filter		
Flue gas pressure drop	psi	0.27
Particle removal efficiency for particles in the low size range	% removal	99.0
Particle removal efficiency for particles in the medium size range		99.0
Particle removal efficiency for particles in the high size range		99.0
ID Fan		
Flue gas outlet pressure (at stack)	psia	14.7

4.0 DOCUMENTATION OF ENERGY MODEL

Some process areas of of the EFCC plant consume significant amounts of electrical power for the operation of certain components (e.g., conveyor belt, pumps, fans, etc.). The net saleable power output of the plant is reduced due to these auxiliary power requirements. Previously developed ASPEN performance models do not adequately estimate these power requirements. New auxiliary power requirement models have been developed and integrated with the ASPEN performance model. These models are based on process variables such as flow rates.

In the METC performance model of the EFCC system the auxiliary power requirement has been assumed to be three percent of the net electrical power output. No basis has been cited for this assumption. The auxiliary power requirement need not necessarily be three percent of the plant total and is expected to change with a change in the process variables of the system. A change in auxiliary power affects the net plant power output and plant efficiency. For example, steam injection to the gas turbine may significantly change gross power output without significantly changing auxiliary power loads. For these reasons a better estimate of the plant auxiliary power requirement is needed.

4.1 Coal Handling

The EFCC system uses dry, pulverized coal as feed to the coal combustor. Therefore, a coal handling system similar to that used in a Pulverized Coal Fired Steam (PCFS) plant is required for coal handling in an EFCC plant. For a 600 MW PCFS power plant with a mass flow rate of coal of 6350 tons/day the auxiliary power requirement is reported to be 2480 kW (Pietruszkiewicz, 1988). Based on this data, the auxiliary power requirement for the coal handling section can be estimated by the following equation:

$$W_{e,CH} = 0.391\ m_{cf,CH,i} \tag{4.1}$$

A typical mass flow rate of coal for a 264 MW EFCC plant is 2,180 tons/day, and the auxiliary power requirement using the above equation is 850 kW.

4.2 Boiler Feedwater Treatment

The boiler feedwater system supplies the water for steam generation in the Heat Recovery Steam Generator (HRSG). The boiler feedwater consists of the raw water (makeup water) and the steam turbine condensate. Raw water is treated and mixed with steam condensate. The combined stream is then chemically polished. Figure 4.1 shows a simple schematic of the boiler feedwater treatment system.

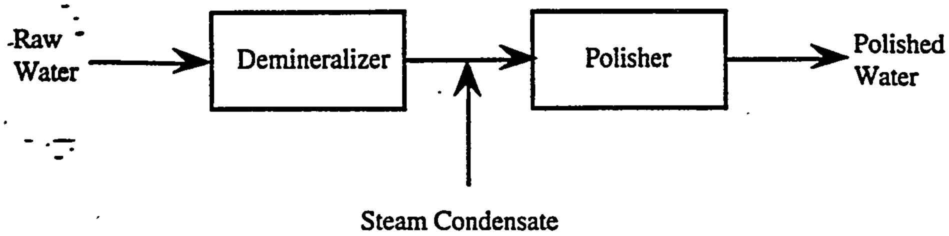


Figure 4.1 Diagram of Boiler Feedwater Treatment System

The equipment in the boiler feedwater system consists of a water demineralization unit for raw water, a demineralized water storage tank, a condensate surge tank for storage of both demineralized raw water and steam turbine condensate water, a polishing unit, and a blowdown flash drum. The major streams in this process section are the raw water inlet and the polished water outlet.

The boiler feedwater section is generic to the steam cycle. Therefore, the following equation which estimates the auxiliary power requirement for the boiler feedwater treatment section in the steam cycle of an IGCC system (Frey and Rubin, 1990) can be applied to the EFCC system:

$$W_{e,BF} = 20.8 + 2.13 \times 10^{-4} m_{pw} \qquad (4.2)$$

where,

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

For this model $R^2 = 0.975$, $n = 14$, and the standard error of the estimate is 38 kW.

The coefficient of determination, R^2, is a measure of the portion of the variance in the predicated variable, $W_{e,BF}$, that is explained by variance in the independent variable, m_{pw}. The number of data points, n, is one indicator of the statistical significance of the regression model. The standard error of the estimate is an indicator of the residual uncertainty in predictions of power consumption that cannot be explained by only the polished water flow rate. This uncertainty may be reduced by developing a more detailed model that accounts for additional performance or design variables. However, in this case the standard error is small and is not expected to contribute substantially to overall uncertainty in the energy balance model.

A typical mass flow rate of polished water for a 264 MW EFCC plant is 778,250 lb/hr, which is within the valid range of values for this model.

4.3 Induced Draft Fan

An Induced Draft (ID) fan is required downstream of the fabric filter to overcome pressure drops in the slag screen, CerHx, HRSG, FGD unit, and the fabric filter. Therefore, the power requirement of the ID fan depends upon the pressure drops across these units. In the METC performance model the ID fan was modeled as a compressor required to boost the pressure of the flue gas to atmospheric pressure.

An ID fan is more realistically modeled as a fan since the pressure differential is less than 15 kPa (2.176 psi). Compressors are more complicated equipment which are used for applications where the pressure differential is 3.5 kPa (0.51 psi) to several hundred kPa (psi). Therefore, it is more practical and economical to use a fan to raise the pressure of the flue gas to one atmosphere.

The power requirement of an ID fan is estimated by the following equation (Ulrich, 1984):

$$W_{e,ID} = 0.0542\left(\frac{m_{fg,ID,i} \times \Delta p_{ID}}{D_{fg,ID,i} \times \eta_{fan}}\right) \tag{4.3}$$

The magnitude of the pressure difference across the ID fan depends on the pressure drop of the flue gas across the coal combustor, slag screen, ceramic heat exchanger, HRSG, FGD unit, and the fabric filter. Representative values of these pressure drops are shown in Table 4.1. The pressure difference between the fabric filter compartment and the outside atmosphere is 1.426 psi, which is equal to 40 inches of water. Although fabric filter designs with a pressure difference of 20 inches of water are common, designs are commercially available from vendors for 40 inches of water pressure difference (Mars, 1995).

For an EFCC plant, a typical ID fan pressure rise of approximately one psi is required. For a typical 264 MW_{net} EFCC power plant, this corresponds to a power requirement of 3.1 MW.

4.4 Primary Air Fan

A primary fan is required to add auxiliary air to the coal combustor. Auxiliary air is required to deliver the pulverized coal from the coal handling system to the coal combustor. Therefore, the auxiliary air requirement is proportional to the coal feed rate to the combustor. In a 264 MW EFCC system, 144,000 lb/hr of auxiliary air is assumed to be required for pneumatic transport of the pulverized coal feed (METC EFCC performance model, 1993). In the METC performance model the primary air mover has been modeled as a compressor. The pressure rise required for the auxiliary air is 1.7 psi. Therefore, for reasons outlined in Section 4.3, the primary air mover is more accurately modeled as a fan.

The equation for estimating the power requirement is similar to the one used for the ID fan.

Table 4.1. Representative Values of Pressure Drops of the Flue Gas

Process Equipment	Inlet Pressure (psia)	Pressure Drop (psi)	Outlet Pressure (psia)
Coal Combustor	15.130	0.11	15.020
Slag Screen	15.020	0.356	14.664
Ceramic Heat Exchanger	14.664	0.76	13.904
Heat Recovery Steam Generator			
Reheater	13.904	0.028	13.876
Superheater	13.876	0.028	13.848
High Pressure Boiler	13.848	0.028	13.820
Economizer	13.820	0.028	13.792
Low Pressure Boiler	13.792	0.028	13.764
FGD			
Mixer	13.764	0.2	13.564
Reactor	13.564	0.02	13.544
Fabric Filter	13.544	0.27	13.274
ID Fan	13.274	1.426 (pressure rise)	14.7

$$W_{e,PA} = 0.0542\left(\frac{m_{a,PA,i} \times \Delta p_{PA}}{D_{a,PA,i} \times \eta_{fan}}\right) \tag{4.4}$$

The pressure rise across the primary air fan in the base case 264 MW EFCC system is 1.7 psi and the power consumption is 3.02 MW.

4.5 Steam Cycle Power Requirement

In the steam cycle auxiliary power is consumed by the condenser pump and the boiler feed pump. The power consumption is estimated directly from the ASPEN simulation.

The condenser pump and the boiler feedwater pump have been modeled in ASPEN as centrifugal pumps using the unit operation block "PUMP". The outlet pressure has been set at 17.5 psia for the condenser pump, and at 2000 psia for the boiler feedwater pump. Using the inlet stream information and outlet pressure requirement, "PUMP" calculates the electrical power requirement.

4.6 Flue Gas Desulfurization (FGD) Power Requirement

In the FGD unit, limestone slurry is injected into absorber towers to remove sulfur dioxide from the flue gas. Electrical power is consumed in the FGD by the different FGD process areas. Table 4.2 shows the various FGD process areas which consume electrical power and the process variable related to the amount of power consumed.

The power consumed by the different process areas of the FGD unit in an EFCC plant can be estimated based on the reported values of the power consumed by similar FGD process areas in a base case utility plant. This is done by scaling the power consumption to the magnitude of the process variable related to the process area. The power consumed in the flue gas system of the FGD unit in the EFCC plant is estimated as part of the power consumption of the ID fan. The power consumed by the different process areas of a 300 MW base case utility plant FGD unit is shown in Table 4.2. A process area description of a FGD unit is given in Table 4.3

The power consumption of process area 73 is already accounted for in the ID fan power requirement. Based on the data in Table 4.2 the power consumed by the various process sections of an FGD unit in kW, can be estimated by the following equations:

$$W_{e,FGD,71} = 0.00234\ m_{L,FGD,i} \tag{4.5}$$

$$W_{e,FGD,72} = 0.865\ Y_{SO_2} \tag{4.6}$$

$$W_{e,FGD,74} = 0.0023\ m_{slag,FGD,o} \tag{4.7}$$

$$W_{e,FGD,75} = 0.006 \sum_{j} W_{e,FGD,j} \tag{4.8}$$

where,

j = process areas 71,72,74, 75

Table 4.2. **Electrical Power Consumption of Base Case Utility Plant FGD Unit for FGD Process Areas**[a]

Area No.	Area Description	Process Variable	Process Variable Value	Power Consumed (kW)
71	Reagent Feed System	Reagent feed (gpm)	82	230
72	SO_2 Removal System	SO_2 concentration (ppm)	3,571	3,090
73	Flue Gas System	Flue gas volumetric flow rate (ft^3/min)	971,000	1,550
74	Solids Handling System	Sludge flow rate (lb/hr)	30,800	70
75	General Support Equipment	Power consumed for other FGD Areas (kW)	4,940	30

[a] EPRI (1990)

The total power consumption of the FGD unit, excluding the power consumption of the flue gas system, is given by:

$$W_{e,FGD} = W_{e,FGD,71} + W_{e,FGD,72} + W_{e,FGD,74} + W_{e,FGD,75} \tag{4.9}$$

4.7 General Facilities

The general facilities include power requirement for cooling water systems; plant and instrument air; potable and utility water; fuel system; effluent water treating; flare system; fire water system; interconnecting piping; buildings; railroad facilities, roads, and lighting; computer control system; and electrical system. The in-plant power consumptions for general facilities for a 600 MW PCFS plant is reported to be 3,400 kW. The total auxiliary power consumption for this plant is 40,500 kW, which is based upon the power consumption by coal handling and preparation section; an FD fan; an ID fan; a PA fan; an electrostatic precipitator; ash handling system; a cooling water system; a seam, boiler feedwater, and condensate system, an FGD unit; and transformer losses . The general facilities power consumption is therefore estimated to be 9.3 percent of all other auxiliary power consumption. Since the components of general facilities of an EFCC plant

Table 4.3. Process Area Definitions for the FGD System

Process Area Number	Process Area Description	Equipment
71	Reagent Feed System	All equipment required for storage, handling, and preparation of raw materials, reagents, and additives, used in each process.
72	SO_2 Removal System	Equipment required for SO_2 scrubbing, such as the absorption tower, recirculation pumps, and other associated equipment.
73	Flue Gas System	Duct work and fans required for flue gas distribution to the SO_2 scrubbing system, plus gas reheat as required.
74	Solids Handling System	Equipment required for fixation, treatment, and transportation of all sludge/dry solids materials produced by each scrubbing process.
75	General Support System	Additional equipment required to support FGD system operation such as makeup water and instrument air.
76	Miscellaneous Equipment	This area will include plant modification necessitated by the addition of the FGD system. Also included are costs for electrical equipment tie-ins and other associated systems.

are similar to that of the PCFS plant, it is assumed that the general facilities power consumption for an EFCC plant will be on a similar scale to that for the PCFS plant. The general facilities power consumption for the EFCC plant can therefore be estimate by the following equation:

$$W_{e,GF} = f_{e,GF} (W_{e,CH} + W_{e,BF} + W_{e,ID} + W_{e,PA} + W_{e,ST} + W_{e,FGD}) \qquad (4.10)$$

where,

$f_{e,GF} = 0.093$

The total auxiliary power demand is the sum of the auxiliary power consumed by each of the process sections, and is given by:

$$W_{aux} = W_{e,CH} + W_{e,BF} + W_{e,ID} + W_{e,PA} + W_{e,ST} + W_{e,FGD} + W_{e,GF} \qquad (4.11)$$

4.8 Net Power Output and Plant Efficiency

The net power output of the EFCC plant is the total power generated from the gas turbine and steam turbines less the total auxiliary power demand.

The gas and steam turbines have been modeled in ASPEN as a series of compressors and turbines using the unit operation block "COMPR". The outlet pressure and isentropic efficiencies are parameters of this unit operation. The ASPEN performance model calculates the power consumed by the compressors and the power generated by the turbines.

The gas turbine energy balance accounts for multiple stages of the compressor and turbine, the flow of cooling air from the compressor to the turbine, and pressure drops at the compressor inlet (due to an intake air filter), ceramic heat exchanger, and turbine outlet. To represent multiple stages of the compressors and turbines, a series of "COMPR" unit operation blocks are employed. The cooling air flow to the gas turbine is extracted from the compressor outlet flow. For a gas turbine with a pressure ratio of 13.5, the pressure rise across the compressor block is 184.3 psi, the pressure drop across the ceramic heat exchanger is 3.5 psi, and the pressure drop across the turbine block is 179.8 psi.

The steam turbine consists of four stages: one high pressure stage; one intermediate pressure stage; and two low pressure stages. Steam enters the high pressure steam turbine at 1050 ^{o}F and 1178.5 psia, and exits at 785.7 ^{o}F and 720 psia. Steam enters the intermediate pressure stage at 1050 ^{o}F and 690 psia, and exits at 712.3 ^{o}F and 200 psia. The inlet steam conditions to the first low pressure steam turbine stage are 712.3 ^{o}F and 200 psia and the exit conditions are 255.5 ^{o}F and 17.5 psia. Steam enters the second low pressure steam turbine at 255.5 ^{o}F and 17.5 psia, and exits at 101.1 ^{o}F and 0.98 psia.

The power output of the gas turbine and the steam turbine are calculated within the ASPEN performance model. The newly developed auxiliary power models are implemented as part of a FORTRAN subroutine called by the ASPEN input file. The net power output is given by:

$$MW_{net} = MW_{GT} + MW_{ST} - 0.001\ W_{aux} \tag{4.12}$$

where,

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

$$\eta_{plant} = \frac{3.414 \times 10^6\ MW_{net}}{m_{cf,CH,i} \times HHV} \tag{4.13}$$

The coal feed rate and the higher heating value are estimated directly from the ASPEN simulation.

5.0 DIRECT CAPITAL COST OF AN EFCC PLANT

New direct capital cost (DCC) models for major EFCC plant sections are presented in this chapter. The EFCC plant is divided into ten major process areas for the purpose of estimating the direct capital costs, which are listed in Table 5.1. For several process areas, such as the combined cycle, costs are further subdivided. In most cases, parameter values for each direct cost model are presented here. In some cases, due to the proprietary nature of some process areas, cost data are not available to enable the assignment of values to parameters of selected cost models. The cost data which are not available for selected process areas are listed at the end of the corresponding cost model description. In these cases, the mathematical form of the cost model is presented.

Using the appropriate *Chemical Engineering* Plant Cost Index (PCI) the direct cost of a process section can be adjusted for other years. Table 5.2 shows the PCI values for 1983 to 1994.

5.1 Coal Handling Section

The coal handling section for an EFCC system delivers dry pulverized coal to a high temperature slagging combustor. The type of coal may range from bituminous to low rank coal. All equipment in the coal handling section is commercially available. The equipment includes bottom dump railroad car unloading hoppers, vibrating feeders, conveyors, belt scale, magnetic separator, sampling system, double boom stacker, bucket wheel reclaimer, surge bins, hammer mill, vibrating fluid bed dryer, circulating gas blower, baghouse, and dust suppression system.

The coal throughput determines the sizing of most of the process area equipment, and is used as the primary parameter for estimating the direct cost. The equipment in the coal handling section of an EFCC plant is similar to the equipments of the coal handling section of a PCFS plant. Therefore, the direct cost of the EFCC coal handling section is expected to be equal to the direct cost of the coal handling section of a PCFS plant which delivers the same mass flow of coal. The direct capital cost of the coal handling section of a 600 MW subcritical PCFS power plant at 100 percent load condition and coal utilization of 6,350 tons/day, after deducting the 6 percent sales tax included, is reported to be $21 million in 1984 dollars (Bechtel, 1988). In previous studies, a linear dependence of coal handling section cost on coal flow rate was found to yield statistically better results than a nonlinear exponential scaling model (Frey and Rubin, 1990). Therefore, based on the cost data for the PCFS plant, a linear model for the direct cost of the coal handling process area as a function of the coal mass flow to the coal combustor was developed, which is given by:

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

Area Number	Cost Section
10	Coal Handling
20	High Temperature Coal Combustor
30	Slag Screen
40	Ceramic Heat Exchanger
50	Boiler Feedwater System
60	Combined Cycle
61	Gas Turbine
62	Heat Recovery Steam Generator
63	Steam Turbine
70	Flue Gas Desulfurization (FGD) Unit
71	Reagent Feed System
72	SO_2 Removal System
73	Flue Gas System[a]
74	Solids Handling System
75	General Supports System
76	Miscellaneous Equipment
80	Fabric Filter (FF)
81	Particulate Collectors
82	Ductwork
83	Fly Ash Handling System
84	Differential Cost[b]
90	ID Fan
100	General Facilities

[a] An ID Fan is required to overcome the flue gas pressure drop across process areas 71 and 72. If an ID fan is included for the entire plant, then a separate ID fan for the FGD process area is not required.

[b] An ID fan is required to overcome the pressure drop across the process areas 81 and 82. If an ID fan is included for the entire plant, then a separate ID fan for the FF process area is not required

Table 5.2. Plant Cost Index Values[a]

Year	Plant Cost Index[b]
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	361.3
1992	358.2
1993	359.2
1994	368.1

[a] Source: *Chemical Engineering*

[b] All values are annual index

$$DC_{CH} = 3.6\text{x}10^{-7}\, m_{cf,C,i}\left(\frac{PCI}{320.3}\right) \tag{5.1}$$

5.2 High Temperature Coal Combustor

This section presents the direct cost model for the high temperature coal combustor of an EFCC plant. The direct cost of a complete coal combustor includes the following items:

- Combustor vessel
- Refractory lining on the inside of the combustor vessel
- Structural supports
- Miscellaneous features such as flue gas ductwork

In the EFCC system pulverized coal is burned in an atmospheric slagging combustor. As part of Phase II of the EFCC Development Program a 25 million BTU/hr test facility has been constructed by HI in Kennebunk, Maine. The direct cost model of a commercial scale combustor is developed based upon the combustor design of this facility. Data are not currently available for the parameters of the model. The values of the parameters can be assigned in the future as data become available.

The Kennebunk Test Facility (KTF) utilizes a down-fired coal combustor. The cylindrical combustor is divided into two chambers, with a primary stage directly on top of the secondary stage. The total height of the combustor, including the burner, is approximately 40 feet, and the outer diameter of the cylindrical casing is nominally 13 feet. The primary stage occupies approximately two-thirds of the total volume, with the remaining part devoted to the second stage. A collar divides the primary stage from the secondary stage. The collar provides convenient access ports for introduction of the secondary air.

A slag tap is located at the base of the combustor. Slag is expected to be released in both the primary and secondary chambers. A portion of the slag formed in the primary chamber will flow down the walls and exit the collar. The primary region was sized to provide for several seconds of residence time for maximum carbon burnout. The molten slag will fall into the secondary region where the remaining combustion air is mixed and combustion is completed. Through proper control of the secondary air, the second stage temperature can be controlled to facilitate slag capture. The coal feed rate to the combustor is approximately 2,000 lb/hr (Orozco, 1993).

The direct capital cost of the coal combustor includes the cost of the combustor vessel, refractory lining for the inside of the combustor vessel, structural supports, and other miscellaneous features. The coal combustor is a cylindrical vessel made of refractory-lined carbon steel. The internal diameter of the combustor vessel can be determined based on the superficial flue gas velocity requirement. The material cost of the combustor vessel is proportional to the surface area.

To estimate the cost of a coal combustor vessel based on the cost of a combustor of known dimensions, a sizing parameter is required which can be appropriately scaled. Vessel surface area is a good choice of sizing parameter since material costs are proportional to the surface area. A significant portion of the combustor vessel is cylindrical in shape, therefore approximating the vessel as a cylinder will give a reasonable estimate of the surface area.

The internal radius of the combustor vessel is given by:

$$r_C = \sqrt{\frac{G_{fg,C,o}}{3600\,\pi\,N_{O,C}\,V_{fg,C,o}}} \tag{5.2}$$

The radius of the combustor vessel must be larger than this internal radius to accommodate the thickness of the refractory lining. If the thickness of the refractory material is t_r, the internal radius of the vessel is:

$$r_{C,v} = r_C + t_r \tag{5.3}$$

The required height of the combustor is estimated by using an approximate flue gas residence time together with an estimate of the flue gas superficial velocity in the combustor:

$$h_{C,v} = V_s \tau_C \tag{5.4}$$

The surface area of the combustor vessel can then be estimated by the following equation for the surface area of a cylinder of radius $r_{C,v}$:

$$SA_{C,v} = 2\pi(r_{C,v})^2 + 2\pi(r_{C,v})\, h_{C,v} \tag{5.5}$$

The coal combustor used in the EFCC test facility at Kennebunk has been used as the design basis for scaling the size and the cost of the coal combustor in a commercial scale EFCC plant. The refractory thickness of the base case combustor vessel is not reported in the literature. The internal diameter of the vessel is twice the internal vessel radius estimated from equation 5.2. The coal combustion stoichiometric equation can be used to estimate the stoichiometric requirement of air for complete combustion of coal. The excess air ratio can then be used to perform a mass balance of the combustion equation from which the product flue gas molar flow rate and its composition can be calculated. Based on a coal feed rate of 2,000 lb/hr the volumetric flow rate of the flue gas can be calculated at a typical gas temperature of 2,750 °F. The average molecular weight and average density of the product gases can be estimated by simulation using the ASPEN performance model of the EFCC for combustion of similar coal. From these data the volume of product gases per unit coal energy released can be calculated.

The superficial flue gas velocity can be calculated once the cross sectional area and the flue gas volumetric flow rate is known. The overall residence time of the flue gas can be calculated from the superficial flue gas velocity and height of the combustor vessel.

The direct capital cost of a coal combustor can be estimated based on the ratio of the surface areas referenced to the base case design. If the superficial gas velocity and the total gas residence time for the coal combustor are not specified, the base case values may be assumed as a default. The internal radius and surface area of the combustor vessel can then be calculated from Equations (5.2) and (5.5). If the cost of the base case coal combustor is $DC_{C,v,base}$, then the direct capital cost of the coal combustor can be estimated by the following equation:

$$DC_{C,v} = DC_{C,v,base}\, N_{T,C} \left(\frac{SA_{C,v}}{SA_{C,v,base}}\right) \left(\frac{PCI}{PCI_{base}}\right) \tag{5.6}$$

The cost of the refractory material is given by the refractory surface area required to cover the internal side of the combustor vessel and the cost of the refractory material per square foot:

$$DC_{C,r} = 2\pi\, r_{C,v}\, h_{C,v}\, UC_r \tag{5.7}$$

Each combustor requires structural supports. The structural support is assumed to have some economy of scale with respect to size. The surface area of the combustor is used as a surrogate for the size and weight of the combustor and, hence, the proportional size of the structural supports. In the method of scaling direct cost of process equipment based on a sizing parameter and the cost of a reference equipment, 0.6 is the most commonly used exponent. Hence, this method is known as six-tenths scaling rule (Ulrich, 1984). As a default assumption, a six-tenths scaling rule is assumed (Frey, 1994):

$$DC_{C,s} = DC_{C,s,base}\, N_{T,C} \left(\frac{SA_{C,v}}{SA_{C,v,base}}\right)^{0.6} \left(\frac{PCI}{PCI_{base}}\right) \tag{5.8}$$

Other miscellaneous features of the combustor include flue gas ductwork, the cost of which is assumed to be proportional to the volumetric flow rate of the flue gas. For the supports, economies of scale are assumed. Assuming a six-tenths scaling rule, the direct cost for the miscellaneous features of the combustor can be estimated by the following equation:

$$DC_{C,m} = DC_{C,m,base}\, N_{T,C} \left(\frac{G_{fg,C,o}}{G_{fg,C,o,base}}\right)^{0.6} \left(\frac{PCI}{PCI_{base}}\right) \tag{5.9}$$

The total direct cost of the coal combustor can then be estimated by the following equation:

$$DC_C = DC_{C,v} + DC_{C,r} + DC_{C,s} + DC_{C,m} \tag{5.10}$$

Combustor cost model completion is pending the receipt of cost data from METC. The parameters for which data are needed are: the base case combustor direct cost for the vessel, structural supports, and miscellaneous features; the base case excess air ratio for coal combustion; and the unit cost of refractory.

5.3 Slag Screen

The combustion gases exiting the coal combustor enter a slag screen, which is an impact separator. Larger ash particles are collected here to prevent fouling of the ceramic heat exchanger located downstream of the slag screen. The slag screen is constructed in the form of a staggered array of refractory tubes, which prevent the ash particles in a certain size range entrained in the flue gas from passing by. The cost of a slag screen is expected to be proportional to the volumetric flue gas flow rate. Other factors which may affect the size and, hence, the cost, are the ash particle size distribution and ash particle density in the flue gas. Ash particle density depends primarily on the chemical composition of coal. Chemical composition influences the ash fusion temperature, which determines the "stickiness" of the coal ash particles (Orozco and Seger, 1993).

The direct cost of a slag screen can be estimated by scaling it to the cost of a similar slag screen used for a known flue gas volumetric flow rate. The direct cost is given by:

$$DC_{SS} = DC_{SS,base}\left(\frac{G_{fg,SS,i}}{G_{fg,SS,i,base}}\right)^{a}\left(\frac{PCI}{PCI_{base}}\right) \tag{5.11}$$

Slag screen cost model completion is pending the receipt of data from METC. The parameters for which data are needed are: the direct cost of base case slag screen, the volumetric flue gas flow rate for the base case slag screen, and the appropriate scaling parameter value.

5.4 Ceramic Heat Exchanger

The ceramic heat exchanger is a multi-pass shell and tube heat exchanger with an overall heat exchanger effectiveness of 0.85. The compressor discharge air from the gas turbine flows through the tubes, while the ash-laden combustion flue gas flows across the tubes in the shell. Most of the heat exchanger components are made of ceramics, which can withstand temperatures well in excess of 1,900 °F and are resistant to corrosion from the combustion gas. Metal alloys can be used in the low temperature end of the heat exchanger which operates in the range of 600 °F to 1,400 °F (Orozco and Seger, 1993).

The direct capital cost of the heat exchanger can be estimated separately for the ceramic part and for the metal part, based on the cost of a ceramic and a metal alloy heat exchanger of some standard size.

$$DC_{HX} = N_{T,HX}\left[DC_{HX,cer,base}\left(\frac{A_{HX,cer}}{A_{HX,cer,base}}\right)^{a} + DC_{HX,met,base}\left(\frac{A_{HX,met}}{A_{HX,met,base}}\right)^{b}\right]\left(\frac{PCI}{PCI_{base}}\right) \tag{5.12}$$

The surface area of the heat exchanger tubes can be estimated by the following equations (Ulrich, 1984):

$$Q_{HX} = U\, A_{HX}\, \Delta T_{lm} \tag{5.13}$$

$$\Delta T_{lm} = \frac{\Delta T_h - \Delta T_c}{\ln\left(\frac{\Delta T_h}{\Delta T_c}\right)} \tag{5.14}$$

$$\Delta T_h = \Delta T_{h,in} - \Delta T_{c,out} \tag{5.15}$$

$$\Delta T_c = \Delta T_{h,out} - \Delta T_{c,in} \tag{5.16}$$

The overall heat transfer coefficient for a ceramic heat exchanger being used in the KTF is reported to be 10 Btu/hr-ft^2-°F. The heat duty and the temperature of the streams are estimated directly from the ASPEN simulation of the performance model.

Ceramic heat exchanger cost model completion is pending the receipt of data from METC. The parameters for which cost data are needed are: the direct cost of the metallic and the ceramic part of the base case heat exchanger; the surface area of the metallic and the ceramic part of the base case heat exchanger; and the appropriate cost scaling factors.

5.5 Gas Turbine

In this research, the modeling of EFCC systems is intended to include performance representative of typical high-firing temperature (2,350 °F) heavy duty gas turbine technology. However, the intent is not to model exactly the performance of any one proprietary gas turbine model. Instead, the goal is to achieve reasonable accuracy in reproducing the key performance characteristics of this class of gas turbines. Two of the most common 2,350 °F turbine inlet temperature heavy-duty gas turbine models which are offered commercially are the General Electric MS7001F, also referred to as the "Frame 7F", and the Westinghouse/Mitshubishi 501F (Frey and Rubin, 1990).

The gas turbine assumed in the EFCC system is based upon the GE Frame 7F. This gas turbine is designed for a turbine inlet temperature of 2,350°F and has a rated power output, when firing natural gas, of 159 MW. A cost estimate for a single gas turbine has been developed based on cost data available for the GE Frame 7F used in IGCC systems. In four recent site specific IGCC studies performed for EPRI, the cost of the Frame 7F in the first phases of a phased IGCC construction schedule ranged from $30.8 to $33.6 million, with an average of $32.0 million. In two other studies, the cost of the Frame 7F for application in natural gas fired combined cycle plants was estimated at $28.3 and $26.8 million, respectively (Frey and Rubin, 1990). Since hot air is used as the working fluid in the gas turbine in EFCC systems, problems associated with firing the gas turbine with coal gas (e.g., corrosion and erosion) are not expected. However, modifications will have to be made to the off-the-shelf available gas turbine for EFCC applications. The combustor and the fuel valves would have to be removed and replaced with the ceramic heat exchanger. This would eliminate the cost of the combustor and the fuel valves from the total cost of the gas turbine. High temperature insulated piping would be required to carry the hot air from the heat exchanger to the gas turbine. This would add to the cost of the turbine since the cost of the piping is about $1,000/ft (Smith, 1994). Because the modifications to the gas turbine are likely to be custom work that is labor intensive, we choose to use the higher end of the cost range for the more conventional natural gas-fired turbine as a preliminary estimate of the cost of a gas turbine modified for EFCC applications. In addition, we will assign the gas turbine cost greater uncertainty than for conventional turbines in later probabilistic analysis.

The cost of a GE Frame 7F gas turbine in an EFCC system is estimated by:

$$DC_{GT} = 28{,}300\, N_{T,GT} \left(\frac{PCI}{351.5}\right) \tag{5.17}$$

5.6 Heat Recovery Steam Generator

The heat recovery steam generator (HRSG) is a set of heat exchangers in which heat is removed from the exhaust flue gas exiting the ceramic heat exchanger to generate steam. Typically, steam is generated at two or three different pressures. There is one steam drum for each steam pressure level. High pressure superheated steam is generated for use in the steam turbine, and the exhaust from the steam turbine first stage is reheated. The input streams to the HRSG section include the flue gas from the ceramic heat exchanger and boiler feedwater to the deaerator. The major output stream is the high pressure steam to the steam turbine. Several parts of the HRSG must be sized to accommodate the high pressure steam flow.

The HRSG in the EFCC system consists of a 1050 °F superheater, a 1050 °F reheater, an 1785 psia high pressure boiler (evaporator), a high pressure economizer, and a low pressure boiler (evaporator). The low pressure boiler is used to produce steam for the deaerator from the flue gas leaving the economizer.

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 flue gas volumetric flow rate, the number of steam drums, and, to a lesser extent, the boiler feed water and saturated steam flowrates in each of the heat exchanger in the HRSG.

The cost of the HRSG can be calculated based on the cost of a similar HRSG of a specified sizing variable such as the flue gas flow rate. The reheater, superheater, and high pressure economizer can be modeled as a shell and tube heat exchanger, and the high and low pressure boiler can be modeled as an evaporator.

Because of the unique nature of the HRSG in EFCC applications, a vendor quote was obtained in response to a specific design basis. Since the temperature of the inlet flue gas to the HRSG is high (greater than 1500 °F), special material of construction is required for the superheater and the reheater. According to an HRSG vendor's recommendation, the hot end of the reheater requires T-91 material (nine percent chrome ferritic steel without Ni), and the balance of the superheater and reheater requires T-22 material (2.25 percent chrome, one percent molybdenum steel). The other HRSG components can be constructed with carbon steel. The tubing in the HRSG typically has an outside diameter of two inches, and is seamless and welded. All tube sections use spiral-wound high frequency welded fins. The fin size varies from 3 x 0.5 inches to 6 x 0.75 inches (fins/inch x fin height) (Pasha, 1994).

The direct cost of the HRSG can be directly scaled to the volumetric flow rate of the flue gas (Pasha, 1994). The cost of HRSG can be estimated using the following equation:

$$DC_{HR} = DC_{HR,base}\left(\frac{G_{fg,HR,i}}{G_{fg,i,base}}\right) N_{T,HR}\left(\frac{PCI}{PCI_{base}}\right) \tag{5.18}$$

The installed cost of a HRSG for a 264 MW_{net} EFCC plant with a flue gas flow rate of 3,842,000 lbs/hr has been estimated to be 11.05 million dollars (Pasha, 1994). In this cost estimate the following simplifications were taken into consideration:

1. There was no provision for blowdown. In reality all drum type boilers need one to two percent blowdown.
2. The HP turbine flow and the reheater flow were considered to be equal. However, there will always be a loss of up to two percent due to leakage, etc. Therefore, the reheater flow would be 2 percent less than the HP flow.
3. There will be radiation and other energy losses from the gas flow which were not considered. These can be anywhere from 0.4 to 1.0 percent.
4. The gas enthalpies vary with the constituents, particularly water vapor content. Such variations were not considered in the cost estimate analysis provided by the vendor.

5.7 Boiler Feedwater System

The boiler feedwater system consists of equipment for handling raw water and polished water in the steam cycle. This equipment includes a water demineralization unit for raw water, a demineralized water storage tank, a condensate surge tank for storage of both demineralized raw water and steam turbine condensate water, a condensate polishing unit, and a blowdown flash drum. The major streams in this process section are the raw water inlet and the polished water outlet. The boiler feedwater section is generic to the steam cycle.

The cost of the boiler feedwater section depends on both the raw water flow rate through the demineralization unit and the polished water flow rate through the polishing unit. The polished water flow rate includes primarily both the raw water and the steam turbine condensate. The steam cycle condensate is typically larger than the raw water flow rate. A two-variable regression model of the boiler feed water system cost as a function of the raw water and polished water flow rates yields the following equation for estimating the direct cost of this section (Frey and Rubin, 1990):

$$DC_{BF} = 0.145\ m_{rw}^{0.307}\ m_{pw}^{0.435} \left(\frac{PCI}{351.5}\right) \tag{5.19}$$

$R^2 = 0.991$, $n = 14$

where,

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

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

Typically only one train of equipment is used in this section, and all equipment is commercially available.

5.8 Steam Turbine

The steam turbine in the EFCC system consists of a high pressure, an intermediate pressure, and two low pressure turbine stages, and an exhaust steam condenser. The high pressure stage receives high pressure superheated steam from the HRSG. The outlet steam from the high pressure stage returns to the HRSG for reheat, after which it enters the intermediate pressure stage. The outlet steam from the intermediate pressure stage goes to the two low pressure stages.

The cost of the steam turbine is expected to depend primarily on the mass flow rate of steam through the system, the pressure in each stage, and the generator output. A single-variate regression based on the steam turbine generator output yields the following equation for estimating the cost of a single steam turbine (Frey and Rubin, 1990):

$$DC_{ST} = 158.7\ MW_{ST} \left(\frac{PCI}{351.5}\right) \tag{5.20}$$

$R^2 = 0.958$; $n=9$

where,

$200 \leq MW_{ST} \leq 550$

The standard error of the estimate is $5.5 million.

5.9 Flue Gas Desulfurization

A cost model of the wet limestone with forced oxidation flue gas desulfurization (FGD) system developed by Kalagnanam and Rubin (1993) is adapted for use here. The FGD model is based upon a model developed by EPRI called FGDCOST. FGDCOST is a spreadsheet-based system which contains cost models for fifteen FGD processes. However, it does not contain any FGD performance models. Therefore, key performance assumptions must be supplied by the user. A cost model for FGD systems has been developed based on extensive sensitivity analysis of FGDCOST. The results of the sensitivity analysis have been used to derive "response surface" models, or regression models, in which FGD costs are related to selected performance parameters (Kalagnanam and Rubin, 1993).

The direct capital cost of the FGD system is calculated as the sum of direct costs of each major component of the system. The cost areas, along with a brief description, of the FGD system are listed in Table 4.3. The equipment size is based on mass balance performed by FGDCOST. From the required size, direct costs are determined. The sum of the capital cost for all process areas provides the total direct capital cost.

The main variables which effect the process capital costs are flue gas flow rate, SO_2 concentration in flue gas, L/G ratio, and stoichiometric ratio of limestone to sulfur. The capital costs are normalized by plant size and expressed in \$/kW units for all process areas. Regression models for the direct capital cost of all the FGD process areas are presented in Table 5.3 (Kalagnanam and Rubin, 1993):

The total direct cost for the FGD is given by:

$$DC_{FGD} = [C_{71} + C_{72} + C_{74} + C_{75} + C_{76}]\left(\frac{PCI}{PCI_{base}}\right) \tag{5.21}$$

The differential direct cost of the ID fan, which is included in section 73, has not been included in the FGD cost model because it is estimated separately as the plant ID fan.

5.10 Fabric Filter

A reverse gas fabric filter design has been assumed for the EFCC system as discussed in Section 2.3.8.3. A direct capital cost model developed by Kalagnanam and Rubin (1994) is employed. This model includes direct costs for separate components of the fabric filter. The component costs are summed to give the total direct cost of the fabric filter. These components are shown in Table 5.4. The major design parameters which can significantly impact the total system

Table 5.3. Equations for Direct Cost of FGD Process Areas[a]

FGD Process Area	Process Area Description	Direct Cost Equation	R^2	n
71	Reagent Feed System	$\log(C_{71}) = 3.05 - 0.79 \times \log(G_{fg,FGD,i}) + 0.22 \times \log(Y_{SO2}) + 0.137 \times \log(R_{L/G}) + 1.664 \times \log(R_{Ca/S})$	0.995	93
72	SO_2 Removal System	$\log(C_{72}) = 2.67 - 0.44 \times \log(G_{fg,FGD,i}) + 0.019 \times \log(Y_{SO2}) + 0.137 \times \log(R_{L/G}) + 1.372 \times \log(R_{Ca/S})$	0.997	93
73	Flue Gas System	$\log(C_{73}) = 2.422 - 0.398 \times \log(G_{fg,FGD,i}) + 1.372 \times \log(R_{CaCO3})$	0.989	93
74	Solids Handling System	$\log(C_{74}) = 1.32 - 0.619 \times \log(G_{fg,FGD,i}) + 0.379 \times \log(Y_{SO2}) + 1.655 \times \log(R_{Ca/S})$	0.982	93
75	General Support Area	$\log(C_{75}) = 2.717 - 0.899 \times \log(G_{fg,FGD,i}) + 1.433 \times \log(R_{CaCO3}) + 0.022 \times \log(Y_{SO2})$	0.997	93
76	Miscellaneous Equipment	$\log(C_{76}) = 2.523 - 0.697 \times \log(G_{fg,FGD,i}) + 0.016 \times \log(Y_{SO2}) - 0.012 \times \log(R_{L/G}) + 0.958 \times \log(R_{Ca/S})$	0.998	93

[a] Source: Kalagnanam and Rubin (1993)

Table 5.4. Cost Areas for Direct Cost Breakdown of the Fabric Filter

Process Area Number	Process Area Description
81	Particulate Collectors
82	Ductwork
83	Fly Ash Handling System
84	Differential Cost (e.g., ID Fan)

Table 5.5. Model Scaling Factors for Fabric Filter Direct Cost Estimation[a]

Process Area Number	f	a	b	c
81	0.86	0.84	0.15	0
82	0.75	0	0	0
83	0.55	0.29	0.275	0.083
84	0	0	0	0

[a] Source: Kalagnanam and Rubin (1994)

Table 5.6. Process Area Direct Cost for a Base Case Fabric Filter[a]

Process Area	Direct Cost (1990, million $)[b]
81	7.86
82	0.25
83	1.59
84	0.15
Total	9.85

[a] Source: Kalagnanam and Rubin (1994)

[b] Process facility cost for the fabric filter process areas have been reported. Theses costs have been converted to equivalent direct cost by subtracting the indirect construction cost and the sales tax.

cost of the fabric filter are gas flow volume (which depends on the generating unit size), air-to-cloth ratio, the flange-to-flange pressure drop in the baghouse, bag life, and the bag fabric that is used for filtering.

All of the fabric filter component costs are calculated using a common mathematical function. The direct cost of each process area is normalized against the cost for a base case. The model was based on price quotes from equipment vendors obtained for EPRI, and is given by (Kalagnanam and Rubin, 1994):

$$\frac{DC_{FF,i}}{DC_{FF,i,base}} = \left\{\left(\frac{G_{fg,FF,i}}{G_{fg,FF,i,base}}\right)^f\left(\frac{R_{A/C,base}}{R_{A/C}}\right)^a\left(\frac{N_{B,base}}{N_B}\right)^b\left(\frac{m_{a,FF,o}}{m_{a,FF,o,base}}\right)^c\right\}\left(\frac{PCI}{PCI_{base}}\right) \tag{5.22}$$

The total process facility cost of the fabric filter unit is the sum of the process facility cost of the different process areas, and is given by:

$$DCC_{FF} = DCC_{FF,81} + DCC_{FF,82} + DCC_{FF,83} \tag{5.23}$$

The differential direct cost for the ID fan section 84 has not been included in the fabric filter cost model because the direct cost of the ID fan is estimated separately.

The exponents for the model in this equation are given in Table 5.5. These exponents have been verified against cost data collected for unit sizes ranging from 125-500 MW. The process area direct cost for the base case power plant is shown in Table 5.6. The base case power plant is a 250 MW PCFS plant with a flue gas flow rate of 0.95 million acfm, air-to-cloth ratio of 2.0 acfm/ft^2, 360 bags per compartment, 14 bag compartments, and ash flow rate to silo of 10.44 tons/hr.

5.11 Induced Draft Fan

A centrifugal Induced Draft (ID) fan is required downstream of the fabric filter to overcome the pressure drop along the flue gas path. The cost of a centrifugal fan depends on the flue gas flow rate it handles. The volumetric flow rate of a 265 MW_{net} EFCC plant is 64 million ft^3/hr compared to the volumetric flow rate of 103 million ft^3/hr for a 606 $MW_{net\ plant}$. The cost of the ID fan used in EFCC plant can be estimated by scaling the cost to the required flue gas flow rate, and is given by:

$$DC_{ID} = DC_{ID,base}\left(\frac{G_{fg,ID,i}}{G_{fg,ID,i,base}}\right)^a\left(\frac{PCI}{PCI_{base}}\right) \tag{5.24}$$

where,

a = scaling exponent (= 1.17) (Peters and Timmerhaus, 1991)

ID Fan cost model completion is pending the receipt of data from METC. The parameters for which cost data are needed are: the direct cost of the base case ID fan and the flue gas flow rate handled by the base case ID fan.

5.12 General Facilities

The general facilities include cooling water systems; plant and instrument air; potable and utility water; fuel system; flare system; fire water system; interconnecting piping; buildings; railroad facilities, roads, and lighting; computer control system; and electrical system. These equipment items are typically represented in preliminary, study-grade cost estimates as a percentage of direct costs. Most studies assume that general facilities are approximately 15 percent of direct costs (Frey and Rubin, 1990). The costs of general facilities can be estimated as:

$$GF = f_{GF} \sum DC_i \tag{5.25}$$

where,

f_{GF} = general facilities factor (= 0.15)

6.0 TOTAL CAPITAL COST MODEL

A framework for estimating the total capital requirement (TCR) for an EFCC power plant is presented in this section. The method is based on the EPRI Technical Assessment Guide (TAG), 1986). The TCR can be divided into several components as shown in Table 6.1. The direct cost estimation methodology for all the EFCC process areas was presented in the previous chapter. The items included in the direct capital cost (DCC) of each process area are shown in Table 6.2. The method for estimating the other components of the TCR is outlined in the following sections.

6.1 Total Plant Cost

The total plant cost (TPC) of an EFCC power plant includes the process facilities capital (PFC), general facilities capital, engineering and home office fees, process contingency cost, project contingency cost, and cost of environmental permits. The PFC includes the total direct cost, indirect construction cost (ICC), and sales tax.

The total direct cost (TDC) is the sum of the direct capital cost of each process area of the EFCC power plant and the general facilities cost.

$$TDC = (\sum DC_i + GF) \tag{6.1}$$

Indirect construction cost (ICC) includes worker benefits, supervision and administrative labor, purchased and rented construction equipment, and construction facilities, which may include temporary buildings, roads, utilities, railroad, and minimal recreation facilities for workers. In a study grade cost estimate, the ICC may be estimated based upon the total direct cost.

$$C_{ICC} = f_{ICC}\, TDC \tag{6.2}$$

where, as a nominal (default) value,

$f_{ICC} = 0.25$ (Frey and Rubin, 1990)

The cost of sales tax is specific to the state where the power plant is constructed. It is estimated as the tax on material costs. Material costs comprise typically 80 percent of the TDC and 10 percent of ICC. Therefore, the sales tax can be estimated as:

$$C_{tax} = r_{tax}\,(0.8\ TDC + 0.1\ C_{ICC}) \tag{6.3}$$

where,

r_{tax} = typically 0.06

Table 6.1. Items Included in the Total Capital Requirement

Capital Cost Item
Total Plant Cost
Process Facilities Capital (PFC)
Summation of Direct Cost (DC) for all process areas (TDC)
Indirect Construction Cost
Sales Tax
General Facilities Capital
Engineering & Home Office Fees
Process Contingency
Project Contingency
Environmental Permits
Total Plant Investment (TPI)
TPC
Allowance for Funds During Construction (AFUDC)
Total Capital Requirement (TCR)
TPI
Prepaid Royalties
Spare Parts Inventory
Preproduction (or startup) Costs
Inventory Capital (e.g., fuel storage)
Initial Chemicals and Catalyst Charges
Land

The process facilities capital (PFC) cost is the sum of the total direct cost, indirect construction cost, and the sales tax, and is given by:

$$PFC = TDC + C_{ICC} + C_{tax} \tag{6.4}$$

The general facilities capital is estimated as a percentage of the process facilities capital cost, and is given by:

$$C_{GF} = f_{GF}\, PFC \tag{6.5}$$

Table 6.2. Items Included in the Direct Capital Cost of Each Process Area[a]

Item
Earthwork
Concrete
Building and Structures
Process Equipments
Piping
Electrical
Painting
Instrumentation and Controls
Insulation
Direct Field Labor

[a] Source: Kalagnanam and Rubin (1993)

The engineering and home office costs include the costs associated with: (1) engineering, design, and procurement labor; (2) office expenses; (3) licensor costs for basic process engineering; (4) office burdens, benefits, and overhead costs; and (5) fees or profit to the architect/engineer. EPRI recommends that a value of 7 to 15 percent of the process facilities cost be used (EPRI, 1986). The engineering and home office cost is given by:

$$C_{EHO} = f_{EHO}\, PFC \tag{6.6}$$

where,

$f_{EHO} = 0.07$ to 0.15.

A major cost item for advanced technology plants is the process contingency cost. Process contingency is used in deterministic cost estimates to quantify the expected increase in the capital cost of an advanced technology due to uncertainty in performance and cost for the specific design application. In the EPRI cost method, the process contingency is estimated based on separate considerations of contingencies for each process section. The contingency is expressed as a multiplier of the sum of the direct and indirect capital costs for each plant section. Table 6.3 shows the recommended ranges of process contingency factors. The process contingency decreases as the commercial experience with a process area increases. Process contingencies employed for innovative technologies are intended to represent the expected costs of a commercialized (e.g., fifth of a kind) plant (EPRI, 1986). The process contingency for each major plant section is estimated as follows:

$$C_{PC,i} = f_{PC,i}\ PFC_i \tag{6.7}$$

where,

$$PFC_i = DC_i \left[1 + \frac{C_{ICC} + C_{tax}}{TDC} \right] \tag{6.8}$$

The process contingency factor for each plant section of an EFCC plant is given in Table 6.4. Equation (6.7) includes a term which prorates the process facilities costs to each plant section based on the ratio of the plant section direct cost to the plant total direct cost. The total process contingency allowance for the plant is given by the sum of process contingencies for each plant section:

$$C_{PC} = \Sigma\ C_{PC,i} \tag{6.9}$$

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. Based on the type of information used to develop the estimate, EPRI defines four levels of cost estimate. Table 6.5 shows these estimate types along with the project contingency factor for each. The type of estimates developed in this work are best classified as "simplified". The estimates are based on the costs of major equipment, and are taken from studies which present process diagrams for major plant sections typically including 10 or 20 equipment items per section.

The project contingency is given by:

$$C_{PJ} = f_{PJ}\ PFC \tag{6.10}$$

An environmental permit is required for the construction of a new power plant. An allowance for the cost of obtaining a permit should be included in the total capital requirement. This represents the cost associated with obtaining the services of a consultant who provides the various services associated with permits, including estimates of emissions and discharges of gaseous, liquid, and solid wastes; dispersion modeling of air emissions; and preparation of permit applications. The permitting cost is assumed to be one million dollars, and is a rough, order-of-magnitude estimate only. The cost of environmental permits is thus given by:

$$C_{EP} = 1{,}000 \tag{6.11}$$

Table 6.3. Process Contingency Factors Recommended by EPRI[a]

State of Technology Development	Percent of Process Area Cost
New concept with limited data	≥ 40
Concept with bench scale data available	30 - 70
Small pilot plant data (e.g., 1 MW) available	20 - 35
A full-size module has been operated (e.g., 20-100 MW)	5 - 20
The process is used commercially	0 - 10

[a] Cost estimate using these contingency factors are intended to represent the cost of commercialized (e.g., fifth of a kind) process plant

Source: EPRI (1986)

Table 6.4. Process Contingency Factors for Each Plant Section of an EFCC Plant[a]

Area Description	Area No.	Process Contingency Factor
Coal Handling	10	0-10
Coal Combustor	20	30-70
Slag Screen	30	50-100
Ceramic Heat Exchanger	40	50-100
Boiler Feed Water	50	0
Gas Turbine	61	3-25
Heat Recovery Steam Generator	62	5-20
Steam Turbine	63	0
FGD unit	70	0-10
Fabric Filter unit	80	0-10
Induced Draft Fan	90	0-10
General Facilities	100	5

[a] Expressed as a percentage of the sum of the direct and indirect capital cost for each plant section.

Table 6.5. Project Contingency Factors Recommended by EPRI[a]

Type of Estimate	Design Information	Percent of Direct Cost
Simplified	General site, process flow diagram	30 - 50
Preliminary	Major equipment, preliminary piping and instrumentation diagrams	15 - 30
Detailed	Complete process design, site-specific, engineering design in progress, construction contract and schedule	10 -20
Finalized	Complete engineering of process plant	5 - 10

[a] Expressed as a percentage of the sum of the total direct, total indirect, and process contingency

Source: EPRI (1986)

The total plant cost (TPC) is the sum of the PFC, general facilities cost, engineering and home office cost, project contingency, process contingency, and cost of environmental permits:

$$TPC = PFC + C_{GF} + C_{EHO} + C_{PJ} + \sum C_{PC,i} + C_{EP} \tag{6.12}$$

The total plant cost is estimated as if the entire plant were constructed at a single instant, thereby disregarding the time value of money and the time required for construction.

6.2 Total Plant Investment

The total plant investment (TPI) includes the total plant cost plus an allowance for funds used during construction (AFUDC), also referred to as "interest during construction". If the expenditure for the total plant cost is spread uniformly over the construction period, measured in years, then the total plant investment is given by:

$$TPI = AF \times TPC \tag{6.13}$$

where,

$$AF = \frac{Z^N - 1}{N(Z-1)} \tag{6.14}$$

$$Z = \frac{1 + i}{1 + e_a} \tag{6.15}$$

The total plant cost should be based on the date at which construction begins. The total plant investment is expressed in the same year dollars as the total process capital. Based on the construction and start up time table for the Penelec EFCC repowering project, the typical construction time for an EFCC plant is expected to be three to four years (LaHaye and Bary, 1994).

6.3 Total Capital Requirement

The total capital requirement (TCR) includes the TPI, prepaid royalties, spare parts inventory, preproduction (or startup) costs, inventory capital, initial chemicals and catalyst charges, and land costs.

Prepaid royalties are fees paid to the owners of proprietary process technology designs, and are typically estimated as a fraction of the TPI if specific data are not available. The spare parts inventory is also estimated as a fraction of the TPI. An allowance of 0.5 percent of the TPI is assumed for prepaid royalties and spares parts inventory each. Therefore, these costs are estimated as follows:

$$C_{PR} = 0.005\ TPI \tag{6.16}$$

$$C_{SPI} = 0.005\ TPI \tag{6.17}$$

Preproduction costs include one month of fixed operating costs, one month of variable operating costs (excluding fuel and byproduct credit) based on full plant capacity, one-quarter of the full capacity fuel cost for one month of operation at full load, and an allowance of two percent of the TPI.

The total preproduction cost is given by:

$$PPC = PP_{FC} + PP_{OC} + PP_{Fuel} + 0.02\ TPI \tag{6.18}$$

The relationships for calculating preproduction costs for fixed operating cost, variable operating cost, and fuel are presented in Section 7.3.

Inventory capital includes the costs of fuels and other consumables which are inventoried prior to plant startup. For a baseload power plant, this includes 60 days of fuel and consumable inventories based on 100 percent of plant capacity. Examples of consumables include water treatment chemicals. The inventory capital is given by:

$$IC = \sum IC_i \tag{6.19}$$

The inventory capital includes costs for coal, boiler feed water demineralizer and treatment chemicals, water polishing chemicals, cooling water treatment chemicals, plant and instrument air adsorbent, liquified petroleum gas for a flare, fuel oil, and limestone. A detailed expression for the inventory capital is developed and presented in Section 7.3.

The initial catalyst and chemical charge is distinct from the inventory capital, and includes the cost of catalyst and chemicals that are contained in process equipment. Fuel oil is required as a startup and auxiliary fuel for gas turbines. For a nominal 550 MW power plant, 80,000 barrels of fuel oil are required initially. As an approximation, this amount may be scaled with the size of the plant. The estimated initial fuel oil requirement in barrels is therefore given by (Frey and Rubin, 1990):

$$FO_i = 80{,}000\ (MW/550) \tag{6.20}$$

The total capital requirement (TCR) for an EFCC system is given by:

$$TCR = C_{PR} + C_{SPI} + PPC + IC + C_{IC\&C} + A_L UC_L \tag{6.21}$$

UC_L = \$6,500 per acre (EPRI, 1986)

It is common to express the TCR on a normalized (\$/kW) basis. The ambient temperature for which the normalized cost is reported should be specified because the gas turbine power output is a function of ambient temperature.

7.0 OPERATING AND MAINTENANCE COST MODEL

EFCC plant operating and maintenance costs are estimated using the method presented in EPRI Technical Assessment Guide (EPRI, 1986). The operating and maintenance cost of an EFCC plant consists of fixed and variable costs. The various components of these costs are shown in Table 7.1. Fixed operating costs are independent of plant capacity factor or plant load. Variable costs, which include fuel, consumables, and waste disposal, are directly proportional to the amount of energy produced by the plant. Fixed operating costs are discussed in section 7.1 and variable operating costs are discussed in section 7.2.

7.1 Fixed Operating Costs

Fixed operating costs are annual costs incurred regardless of the level of plant output of the plant. These include operating labor, maintenance labor, maintenance materials, and overhead costs associated with administrative and support labor.

The operating labor cost is based on an estimate of the number of personnel hours required to operate the plant multiplied by an average labor rate. It is commonly assumed that four shifts per day are required for plant operation, allowing two hours overlap for transition between shifts. An allowance for personnel on sick leave or vacation is incorporated into a "shift factor" which is an equivalent number of shifts per day. A shift factor of 4.75 has been assumed for this study.

The total operating labor cost is estimated by summing the number of plant operators per shift for all process areas, applying the shift factor, and multiplying by the average labor rate as follows:

$$OC_L = ALR\ (2{,}080\ \text{hours/year})\ SF\ \Sigma_i\ O_i \qquad (7.1)$$

The number of operators required for each process area of the EFCC system is show in Table 7.2, based on values reported by or calculated from EPRI (1988a,b); Frey and Rubin (1990); and Vatavuk (1990). The calculations for number of operators required is shown in Appendix C.1.

The cost of maintenance labor and material for new technologies is typically estimated as a percentage of the installed capital cost for each process section. Therefore, it is estimated as:

$$OC_M = \Sigma_i\ f_{M,i}\ PFC_i \qquad (7.2)$$

The process facility cost of process section i, PFC_i, is estimated by Equation (6.8)

Table 7.1. Items Included in the Operating and Maintenance Cost Model[a]

Operating and Maintenance Cost Item
Fixed Operating Cost Items
Operating labor
Maintenance labor
Maintenance materials
Overhead costs associated with administrative and support labor
Variable Operating Cost Items
Consumables (e.g., raw water, chemicals, limestone)
Slag disposal
Fuel
Byproduct credit

[a]Source: Frey and Rubin (1990)

Table 7.2. Operating Labor Requirements for the EFCC system

Process Area	Area Number	Number of Operators per Shift[a]
Coal Handling	10	4
Coal Combustor	20	4
Slag Screen	30	1
Ceramic Heat Exchanger	40	1
Boiler Feed Water	50	1
Combined Cycle	60	$4\ N_{O,GT} + 1$
FGD unit	70	2
Fabric Filter unit	80	1
ID Fan	90	1
General Facilities	100	2

[a] Sources: EPRI (1988a,b); Frey and Rubin (1990); Vatavuk (1990)

The maintenance cost factors for each process area of the EFCC system is shown in Table 7.3, based on values reported by or calculated from EPRI (1988a,b); Frey and Rubin (1990); and Vatavuk (1990). The calculations for maintenance cost factors is shown in Appendix C.2.

Table 7.3. **Annual Maintenance Cost Factors for EFCC System**

Process Area	Area Number	Annual Maintenance Cost Factors (%)[a,b]
Coal Handling	10	3
Coal Combustor	20	2-4
Slag Screen	30	1-3
Ceramic Heat Exchanger	40	1-5
Boiler Feed Water	50	1.5
Combined Cycle	60	1.5
FGD unit	70	5
Fabric Filter unit	80	1.06
ID Fan	90	1-3
General Facilities	100	1-3

[a] Annual maintenance cost as a percent of sum of plant section direct, indirect, and contingency costs, as presented in Equation (7.2).

[b] Sources: Bechtel (1988); Frey and Rubin (1990); Vatavuk (1990).

The maintenance cost can be divided into materials and labor components by assuming that 60 percent of the maintenance cost is associated with maintenance materials and the remainder is associated with maintenance labor (EPRI, 1986). Therefore, these costs are estimated as follows:

$$OC_{MM} = 0.6\ OC_M \tag{7.3}$$

$$OC_{ML} = 0.4\ OC_M \tag{7.4}$$

The administrative and support labor cost is assumed to be 30 percent of the operating and maintenance labor cost. Therefore, it is estimated as follows:

$$OC_{AS} = 0.3\ (\ OC_L + OC_{ML}) \tag{7.5}$$

The total fixed operating cost is the sum of the operating labor, maintenance, and administrative support and labor costs:

$$FOC = OC_L + OC_M + OC_{AS} \tag{7.6}$$

7.2 Variable Operating Costs

Variable operating cost includes fuels, consumables, slag and ash disposal, and byproduct credits. No significant byproduct credits are expected from an EFCC plant that employs conventional wet limestone forced oxidation FGD.

7.2.1 Fuel Consumption

Fuel consumption for the EFCC power plant is estimated by the ASPEN simulation models on a mass flow rate basis. The total annual fuel consumption, in units of million Btu, is then determined as follows:

$$Q_{coal} = 8{,}760\ c_f\ m_{cf,C,i}\ HHV_{coal} \tag{7.7}$$

7.2.2 Consumables

The consumables in the EFCC plant consist of chemicals required for boiler feed water treatment, limestone for FGD, and replacement bags for the fabric filter. The costs of these consumables represent a significant portion of the operating cost of an EFCC plant.

7.2.2.1 Boiler Feed Water Consumables

In the boiler feed water system, raw water is treated and mixed with steam condensate from the steam turbine. The combined stream is then chemically polished. The chemicals required for raw boiler feed water demineralization are sulfuric acid and sodium hydroxide. The chemicals required for raw boiler feed water treatment include sodium phosphate, hydrazine, and morpholine. The required quantity of these chemicals is proportional to the raw water intake rate. The chemicals required for polishing the boiler feed water and steam cycle condensate include sulfuric acid and sodium hydroxide. The required quantity of the polishing chemicals is proportional to the flow rate of water through the polishing unit, which is a combination of the raw water feed rate and the flow rate of condensate from steam turbine.

The required quantity of each steam cycle water treatment chemical has been estimated as a function of the raw or polished water flow rate. For each chemical and application, the regression model is presented and summarized in Table 7.4 (Frey and Rubin, 1990).

7.2.2.2 FGD Consumables

The consumables used in a FGD unit includes the limestone slurry feed, which consists of limestone and water. A ten weight-percent Greer limestone slurry is typically used for wet limestone with forced oxidation FGD systems. The cost of consumables of a FGD unit depends on the cost of limestone reagent and water, and the amount of reagent and water used. The amount of reagent used depends on the total sulfur in the flue gas, sulfur removal efficiency, and the stoichiometric ratio. The amount of limestone and water used is estimated directly from the ASPEN performance model.

Table 7.4. Regression Model for Boiler Feed Water Treatment Requirement[a]

Description	Mass Requirement Equation	Variable Limit (lb/hr)	R^2	n	Standard Error (SE)
Sulfuric Acid for Raw Water Demineralization	$m_{sa,BF,i} = c_f\,(47.0 + 2.09\text{x}10^{-3}\, m_{rw})$	$24{,}200 \leq m_{rw} \leq 613{,}000$	0.969	14	70 tons/year
Sodium Hydroxide for Raw Water Demineralization	$m_{sh,BF,i} = c_f\,(9.5 + 4.20\text{x}10^{-4}\, m_{rw})$	$24{,}200 \leq m_{rw} \leq 613{,}000$	0.969	14	15 tons/year
Sodium Phosphate for Raw Water Treating	$m_{sp,BF,i} = c_f\,(115 + 3.61\text{x}10^{-3}\, m_{rw})$	$24{,}200 \leq m_{rw} \leq 613{,}000$	0.962	14	140 lb/year
Hydrazine for Raw Water Treating	$m_{hy,BF,i} = c_f\,(529 + 0.0174\, m_{rw})$	$24{,}200 \leq m_{rw} \leq 613{,}000$	0.898	14	1,200 lb/year
Morpholine for Raw Water Treating	$m_{mo,BF,i} = c_f\,(420 + 0.0163\, m_{rw})$	$24{,}200 \leq m_{rw} \leq 613{,}000$	0.965	14	610 lb/year
Sulfuric Acid for Condensate Polishing	$m_{sa,BFP,i} = c_f\,(15 + 5.4\text{x}10^{-5}\, m_{pw})$	$1{,}200{,}000 \leq m_{pw} \leq 2{,}200{,}000$	0.992	7	2 tons/year
Sodium Hydroxide for Condensate Polishing	$m_{sh,BFP,i} = c_f\,(30 + 1.07\text{x}10^{-4}\, m_{pw})$	$1{,}200{,}000 \leq m_{pw} \leq 2{,}200{,}000$	0.991	7	4 tons/year

[a] Source: Frey and Rubin, 1990

7.2.2.3 Fabric Filter Consumables

Fabric filter consumables include the replacement bags. U.S. utilities primarily use woven fiberglass bag fabric which can withstand temperatures of up to 500 °F and can be textured to control dustcake formation. However, this material is susceptible to abrasion wear. The choice of the bag fabric affects the cost of the bags. Bags generally fall into two categories, 30-36 feet in length and 1 foot in diameter, and 20-22 feet in length and 8 inches in diameter. Except for manufacturing defects and improper installation, bag life is generally not an issue in the design of fabric filters and is usually 3-5 years (Kalagnanam and Rubin, 1994). The average life of bags is assumed to be 4 years.

The air-to-cloth ratio is determined based on the bag cleaning method, which controls the residual quantity of the material remaining on the bags. Utility baghouses typically use air-to-cloth ratios of 1.5-4.0 acfm/ft^2 depending on the bag cleaning method. Typical values used for the air-to-cloth ratio for various baghouse types based on industry experience are shown in Table 7.5. Since a reverse gas cleaning method has been assumed for the EFCC system, a typical air-to-cloth ratio is two. One fourth of the total cloth area can be assumed to be replaced annually. Therefore, knowing the cost of bag cloth in \$/area, the cost per year for bag replacement can be estimated by the following equation:

$$OC_{bags} = \left(\frac{1}{4}\right) 8{,}760\, c_f\, UC_{cloth} \left(\frac{G_{fg,FF,i}}{R_{A/C}}\right) \tag{7.8}$$

7.2.2.4 Other Consumables

Other consumables that are required for the EFCC plant are gas turbine startup fuel, plant and instrument air adsorbent, water, and liquified petroleum gas for flares. The consumables required for all systems are estimated based on a simple scaling relationship with plant size, based on the assumption that the requirement of these items scales directly with the plant capacity. The requirements for these items are based on data reported in Flour (1985), for a nominal 550 MW plant.

The annual required gas turbine startup fuel oil, in barrels, is estimated to be:

$$m_{fo,GT,i} = 48{,}000\, c_f\, (MW/550) \tag{7.9}$$

The annual plant and instrument air adsorbent, in pounds, is estimated to be:

$$m_{ads,GF,i} = 3{,}600\, c_f\, (MW/550) \tag{7.10}$$

Table 7.5. Typical Values of Air-to-Cloth Ratio for Different Baghouse Types[a]

Baghouse Type	$R_{A/C}$ (acfm/ft^2)
Reverse Gas	1.7-2.0
Reverse Gas Sonic	1.7-2.0
Shake-Deflate	2.5-3.0
Pulse Jet	4.0-4.5

[a] Kalagnanam and Rubin (1994)

The plant water consumption is given by the raw water consumption for the steam cycle.

$$m_{water} = 8{,}760\ c_f\ m_{rw} \tag{7.11}$$

The consumption of liquified petroleum gas for maintaining the plant flare is estimated to be:

$$m_{LPG} = 7{,}200\ c_f\ (MW/550) \tag{7.12}$$

7.2.3 Slag, Sludge, and Ash Disposal

The slag from the slagging coal combustor needs to be disposed and imposes a disposal cost. The total slag to be disposed may be estimated by the following equation which is based upon the total ash content in the coal:

$$m_{slag,C,o} = f_{slag}\ f_{ash}\ m_{cf,C,i} \tag{7.13}$$

The amount of coal ash which is removed as slag is a user-adjustable parameter. A typical value for f_{slag} for a wet-bottom (slagging) boiler is 0.65, based upon emission factor data presented in the U.S. Environmental Protection Agency's emission factor handbook (EPA, 1993).

The sludge disposal cost of the FGD depends on the amount of solid waste generated and the unit cost of sludge disposal. The amount of sludge generated depends on the percentage sulfur in coal and the sulfur removal efficiency. The amount of sludge is estimated as the sum of SiO_2 in the limestone slurry feed, unreacted $CaCO_3$, and $CaSO_4$ formed by the desulfurization reaction.

The ash disposal cost of the fabric filter depends on the the amount of ash removed from the flue gas and the unit cost of ash disposal. The amount of ash removed depends on the particulate matter concentration of the flue gas and the ash removal efficiency of the fabric filter. The amounts of sludge and ash removed are estimated in the ASPEN performance model.

7.2.4 Total Variable Operating Costs

To estimate the total variable operating cost, the annual material requirements of the EFCC system must be multiplied by their respective unit costs. Default unit costs of fuel, consumables, and slag disposal for a particular year are given in Table 7.6, and are adjusted to a common year by use of the industrial chemical producers price indicator (CICPPI) index. The CICPPI index values are shown in Table 7.7. These indices are published every month in the *Chemical Engineering* journal. The base year is assumed to be 1982 with a CICPPI index of 100, and the cost of a chemical for any other year can be obtained by scaling its cost to the value of the CICPPI index for that year.

The various components of the total variable cost, expressed in dollars, are:

$$OC_{fuel} = m_{cf,C,i}\, UC_{coal} \tag{7.14}$$

$$OC_{cons} = \sum_{cons} m_i\, UC_i \tag{7.15}$$

$$OC_{slag} = m_{slag,C,o}\, UC_{slag} \tag{7.16}$$

$$OC_{sludge} = m_{sludge,FGD,o}\, UC_{sludge} \tag{7.17}$$

$$OC_{ash} = m_{ash,FF,o}\, UC_{ash} \tag{7.18}$$

The total variable cost is then:

$$VOC = OC_{fuel} + OC_{cons} + OC_{slag} + OC_{sludge} + OC_{ash} \tag{7.19}$$

7.3 Preproduction Cost and Inventory Capital

The three components of the preproduction costs are fixed operating costs, variable operating costs, and preproduction fuel cost. The operating costs are expressed in units of dollars, while the capitals costs are expressed in units of 1,000 dollars. Therefore, a conversion factor is required when estimating preproduction and inventory capital costs based on the operating cost equations developed in Section 6.3.

One month of fixed operating cost (in thousand of dollars) is given by:

$$PP_{FC} = \frac{1}{12}\left(\frac{FOC}{1,000}\right) \tag{7.20}$$

Table 7.6. Unit Costs of Fuel, Consumables, and Disposal

Item	Units	Cost	Basis	Reference
Fuel				
Illinois No.6 Coal	$/MMBTU	1.55	1/85	EPRI, 1986
Consumables				
Sulfuric acid (93%)	$/ton	89.4	1/87	BGE, 1989
Sodium Hydroxide (50%)	$/ton	175	1/87	BGE, 1989
Sodium Phosphate	$/lb	0.55	1/87	BGE, 1989
Hydrazine	$/lb	2.5	1/87	BGE, 1989
Morpholine	$/lb	1.02	1/87	BGE, 1989
Limestone	$/ton	15.0	1/90	Kalagnanam and Rubin, 1993
Dibasic Acid	$/ton	360.0	1/90	Kalagnanam and Rubin, 1993
Filter bags	$/bag	80.0	12/90	Kalagnanam and Rubin, 1994
GT startup fuel oil (NO. 6)	$/gal	0.41	1/90	Kalagnanam and Rubin, 1993
Plant & Instrument air adsorbent	$/lb	2.30	1/83	Flour, 1986
Water	$/1000 gal	0.6	1/90	Kalagnanam and Rubin, 1993
LPG	$/bbl	9.24	1/87	BGE, 1989
Slag Disposal	$/ton	8.00		same as ash
Sludge Disposal	$/ton (dry)	9.25	1/90	Kalagnanam and Rubin, 1993
Ash Disposal	$/ton	8.00	1/90	Kalagnanam and Rubin, 1993

One month of variable operating costs, excluding fuel and byproduct credits, is given by:

$$PP_{OC} = \left(\frac{0.083}{c_f}\right)\left(\frac{OC_{cons} + OC_{slag} + OC_{sludge} + OC_{ash}}{1{,}000}\right) \tag{7.21}$$

Table 7.7. **Chemical Producers Price Indicator Index (CICPPI) Values**[a]

Year	**Plant Cost Index**[b]
1983	339.9
1984	347.4
1985	337.7
1986	341.7
1987	323.9
1988	349.9
1989	411.25
1990	391.87
1991	362.5
1992	360.8
1993	358.9
1994	409.6

[a] Source: *Chemical Engineering*

[b] All values are annual index

The preproduction fuel cost is given by:

$$PP_{Fuel} = \left(\frac{0.021}{c_f}\right)\left(\frac{OC_{fuel}}{1{,}000}\right) \tag{7.22}$$

The total preproduction capital costs are then estimated by substituting Equations (7.20), (7.21), and (7.22) into Equation (6.18) in Section 6.3.

The inventory capital costs include 60 days of fuel and consumable inventories, based on full plant load. These costs can be estimated from the variable operating costs by equation (7.23):

$$IC = \left(\frac{0.164}{c_f}\right)\left(\frac{OC_{fuel} + OC_{cons}}{1{,}000}\right) \tag{7.23}$$

Equation (7.23) can be substituted for Equation (6.19) in Section 6.3

9.0 MODEL APPLICATIONS

In the previous chapters, we have documented the development of a new performance, emissions, and cost model of the EFCC. The new performance model is based upon an ASPEN flowsheet originally developed at DOE/METC. This model has been substantially modified to better characterize factors affecting net plant thermal efficiency. In this chapter, we present sample model results to illustrate the application of the new EFCC plant performance and emissions models. The case studies here include a base case analysis and a series of sensitivity analyses. The intention of these analyses is to illustrate the behavior of the new model in response to changes in the input assumptions. The base case is developed as a reference point for performing the sensitivity analysis; however, it is not intended to be a definitive performance assessment for the EFCC. Because the EFCC is in the research and development phase, there is considerable uncertainty regarding the performance, emissions, and cost of a future commercial-scale system. Thus, the sensitivity analyses in this report are also intended to provide insight into model behavior that will be helpful in developing and interpreting probabilistic analyses of the EFCC as part of future work.

The modeling results presented here are for the performance and emissions models only. Due to the proprietary nature of the EFCC process, and the lack of published or publicly available cost data for key components of this process, including the combustor, slag screen, and ceramic heat exchanger, it was not possible to obtain credible values of key cost model parameters for these process areas. Thus, cost model results are not presented here.

9.1 Modeling Assumptions

Based on a detailed review of published information regarding the EFCC, as well as conversations with process engineers familiar with the EFCC, we have identified thirteen performance variables and design parameters for sensitivity analysis. A set of base case and sensitivity assumptions for the EFCC are given in Table 9.1. All of the sensitivity analyses are based upon an Illinois No. 6 coal, as documented in Table 3.1. The plant is based upon a single heavy duty gas turbine.

A set of nominal base case assumptions have been made based upon design and performance parameter values obtained from the literature. For each of the thirteen selected design and performance parameters, a sensitivity analysis was developed. The ranges of values assigned to each parameter are intended to represent plausible, but not necessarily likely, ranges over which the parameters may vary

9.1.1 Combustor

Due to the high temperature combustion process in the EFCC plant, heat loss from the combustor is expected due to radiation. A combustor heat loss of 0.5 percent has been assumed for the Penelec EFCC repowering project (LaHaye and Bary, 1994). Engineers experienced in the design of EFCC systems have suggested values for the combustor heat loss in the range of one to three percent (Micheli, 1995, and Jarvis, 1995). Therefore, to study the EFCC design over the plausible range of values for the combustor heat loss, this value was varied from 0.5 percent to five percent.

Combustor water walls provide a means of cooling the combustor walls and minimizing the energy lost in the combustor by using some of it for steam generation. Advances in combustor design would minimize the heat loss, and water walls would not be required. Therefore, for the base case water walls were not considered in the EFCC design, and for the sensitivity analysis of the model to the heat captured by the water walls, the combustor water wall heat duty was varied from 0 to three percent.

Current designs of the EFCC combustor have assumed a one percent loss in carbon conversion in the coal combustor (LaHaye and Bary, 1994) but combustor design with no significant loss in carbon conversion is possible (Micheli, 1995). Therefore, one percent loss in carbon conversion was assumed for the base case, and sensitivity analysis was performed over the range of one and zero percent loss in carbon conversion.

Excess air for coal combustion is expected to have significant impact on the economics of the EFCC system (Smith, 1995). For the METC model, 0.8 lb air/lb coal was assumed based on the auxiliary air flow requirement for pneumatic transport of coal in addition to the secondary air flow which is obtained from the gas turbine outlet. The total mass flow of primary and secondary air corresponds to about 100 percent excess air. In a conventional coal combustor usually 20 percent excess air is used as the design value. However, because the combustor design employed here has very little heat recovery, additional air is required to maintain the flame temperature within reasonable limits. An auxiliary air ratio of 0.8 lb air/lb coal was assumed for the modified METC model, and this value was varied from 0.5 to 1.0 for sensitivity analysis study of the EFCC system to excess air.

9.1.2 Ceramic Heat Exchanger

The base case estimate for air leakage in the ceramic heat exchanger is expected to be at least 0.5 percent of the mass flow of the high pressure tube-side air (Orozco and Seger, 1993). To investigate the effect of alternative air leakage assumptions on model predictions, this value was varied from zero to three percent air leakage. It is not expected that air leakage will be as low as zero or as high as three percent. The actual range of probable air leakage rates is expected to be enclosed within this range. The main purpose of the sensitivity is to provide insight into how changes in air leakage rates affect overall process efficiency and power output.

One percent radiation heat loss was assumed for the CerHx in the Penelec project design (LaHaye and Bary, 1994). Heat loss depends on the design of the CerHx, and for a highly insulated system could be negligible with a trade off of higher capital cost. The CerHx heat loss as a percentage of the heat duty was varied from 0 to five percent to study the sensitivity of the EFCC system to alternate CerHx designs.

A 0.85 heat exchanger effectiveness was assumed for the METC model based on current CerHx technology. Active research and development is underway at Hague International for improvement in CerHx design which is expected to have higher effectiveness. Therefore, 0.85 CerHx effectiveness was considered for the modified base case and it was increased to 0.9 for sensitivity analysis case studies.

9.1.3 Gas Turbine

For the gas turbine, alternative design assumptions were made to reflect potential improvements in gas turbine technology. In Table 9.1, the assumptions regarding gas turbine inlet temperature are shown. Many performance studies of the EFCC were predicated upon a 2,300 ^{o}F firing temperature assumption, which was typical of heavy duty gas turbines several years ago. However, as gas turbine technology improves, the firing temperature tends to increase. Currently, gas turbines are commercially available with a rated 2,350 ^{o}F firing temperature. To reflect possible future improvements in gas turbine technology, we also consider a case for a 2,400 ^{o}F firing temperature. The pressure ratio also varies with firing temperature. For example, when first offered commercially, the General Electric Frame 7F was rated based upon a 2,300 ^{o}F firing temperature and a pressure ratio of 13.5. When the firing temperature was uprated to 2,350 ^{o}F, the pressure ratio was increased to 15 (Corman, 1994). Thus, for the 2,400 ^{o}F case, we consider pressure ratios of 15, 16, and 17 to reflect possible improvements in gas turbine technology and to provide insight into the optimization of plant efficiency with respect to gas turbine pressure ratio.

GE has developed a new "H" turbine technology which will enable a gas turbine firing temperature of 2,600 ^{o}F, thereby enabling substantial improvements in the thermal efficiency (Valenti, 1995). Additional research is needed to develop ceramic heat exchanger materials, though, which will be capable of providing an outlet flue gas at this temperature.

Table 9.1. Modeling Assumptions for Base Case and Sensitivity Analyses of the EFCC

Description of Sensitive Parameter	Units	Base Case Assumption	Sensitivity Range	
			Low	High
Combustor Radiative Heat Loss	% of heat released	0.5	0.0	5.0
Combustor Water Wall Heat Duty	% of heat released	0.0	0.0	3.0
Combustor Carbon Conversion	% of inlet carbon	99.0	99.0	100.0
Auxiliary Air Mass Flow Rate	lb air/lb coal	0.8	0.5	1.0
Air Leakage in CerHX	% of mass flow	0.5	0.0	3.0
CerHX Radiative Heat Loss	% of heat duty	1.0	0.0	5.0
CerHX Heat Exchanger Effectiveness		0.85	0.85	0.9
Turbine Inlet Temperature	°F	2,300	2,300	2,400
Gas Turbine Pressure Ratio[a]	Ratio	13.5	13.5	17
Steam Injection	lb/hr	0	0	500,000
Water Injection	lb/hr	0	0	500,000
Ambient Temperature	°F	59	20	95
Stack Gas Temperature	°F	181	150	200

[a] The sensitivity analysis for pressure ratio is related to the sensitivity analysis for gas turbine inlet temperature. For a 2,300 °F turbine inlet temperature, a pressure ratio of 13.5 was used (Frey and Rubin, 1990). For a 2,350 °F turbine inlet temperature, a pressure ratio of 15 was used (Corman, 1994). For a turbine inlet temperature of 2,400 °F, the pressure ratio was varied from 15 to 17.

Steam injection or water injection into the gas turbine can be incorporated in an EFCC design to increase the plant net power output or to reduce the capital cost of the plant on a per kW basis. For the base case modified model, no steam injection or water injection was considered. The Penelec project incorporates 80,000 lb/hr of steam injection, for a gas turbine of 22.0 MW base rating (LaHaye and Bary, 1994), primarily to make up for the loss of mass flow due to the lack of fuel firing in the turbine. For sensitivity analysis of the modified EFCC model, since a much larger gas turbine design was incorporated, steam injection and water injection up to 500,000 lb/hr was considered.

For the base case modified EFCC model, a nominal ambient temperature of 59 °F was considered. A sensitivity study of the system was done for an average winter temperature of 20 °F and for a summer temperature of 95 °F.

A stack gas temperature of 181 °F was considered for the base case modified model based on ranges of values reported for stack gas temperature in literature (Cooper and Alley, 1986, EPRI, 1988). This temperature was varied from 150 °F to 200 °F to study the EFCC plant performance parameters as a function of stack gas temperature. The stack gas temperature determines the need for FGD heat.

In addition we considered a separate case study in which no FGD reheat was used. This represents a so-called "wet stack" which is specifically designed to handle acid gas condensation. The stack gas temperature with no reheat would be 136 °F.

9.2 Base Case Results

The simulation results of the modified METC EFCC model based on the parameter assumptions listed in Table 9.1 are shown in Table 9.2. The simulation results of the original METC model based on the parameter assumptions listed in Table 3.2 are also shown in Table 9.2. The model runs were executed on a VAXStation 3200 using the DOE public version of ASPEN. The run time for a single simulation is 69.26 seconds.

Table 9.2. Simulation Results of the METC Model and New Model

Parameter	Unit	METC Model Result	Modified METC Model Result
Performance			
Coal Consumption	lb/hr	181,600	189,500
Coal Thermal Input	10^6 BTU/hr	2,040	2,130
Gas Turbine Inlet Air	lb/hr	3,306,000	3,500,000
Flue Gas Flow	lb/hr	3,793,000	3,928,000
Gas Turbine Output	MW	134.3	138.2
Steam Turbine Output	MW	150.3	141.3
Total Auxiliaries	MW	20.69	14.22
Net Plant Output	MW	263.9	265.3
Thermal Efficiency	(%, HHV basis)	44.10	42.49
Environmental			
Limestone Consumption	lb/hr	20,470	21,140
Water Consumption	lb/hr	210,200	217,200
Flyash	lb/10^6 BTU	Not Reported	0.008
SO_2 Emissions	lb/10^6 BTU	0.30	0.60
NO_x Emissions	lb/10^6 BTU	Not Reported	0.5
CO_2 Emissions	lb/kWh	1.58	1.623
Stack Gas Temperature	°F	137.65	181.0
Solid Discharges			
Slag	lb/hr	15,980	11,970

(continued)

Table 9.2. (continued)

Parameter	Unit	METC Model Result	Modified METC Model Result
Collected Flyash	lb/hr	Not Reported	6,426
Sludge	lb/hr	27,060	55,190
Blowdown	lb/hr	17,110	17,600

9.2.1 Performance Results

The gas turbine model for the new EFCC system model was based on choked flow conditions at the first stage expander inlet and a more detailed cooling flow circuitry, which increases the gas turbine inlet air by 200,000 lb/hr. The gas turbine power output increases from 134.32 MW for the original METC model to 138.18 MW.

The coal flow requirement increases from 181,600 lb/hr for the original METC model to 189,500 lb/hr for the modified model for several reasons: (1) the gas turbine is sized differently as noted above; (2) the carbon conversion in the original METC model was assumed to be 100 percent whereas it is 99 percent in the modified model; and (3) the heat loss from the combustor was assumed to be zero whereas in the modified model it is 0.5 percent of the heat of combustion.

In the original METC model, the FGD flue gas reheat requirement was not taken into account. Flue gas from the FGD has to be heated to increase the temperature of the flue gas coming out of the stack to increase the plume buoyancy for better dispersion. The flue gas is heated using steam from the steam cycle, which imposes a penalty on the steam cycle power output and the overall cycle efficiency. FGD reheat using steam from the steam cycle was considered in the modified model, which reduces the steam cycle power output from 150.25 MW for the original METC model to 141.3 MW for the modified model.

A more detailed account of the auxiliary electrical loads was considered in the modified model as outlined in Chapter 4. The auxiliary electrical load for the modified model was calculated to be 14.22 MW compared to 20.69 MW for the original METC model. These numbers are exclusive of the FGD reheat penalty which is calculated in the new model. Details of the various components of the auxiliary electrical loads for the original METC and the modified model are shown in Table 9.3.

Table 9.3. In-Plant Power Consumption of METC Model and Modified EFCC Model

Process Section	Power Consumed (kW)	
	METC Model	Modified Model
Auxiliary Power[a]	8,160	N/A
Primary Air Fan	230	216
ID Fan	3,020	6,210
FGD Unit[b]	7,300	3,460
Condenser Pump	14	14
Boiler Feedwater Pump	1,963	2,030
Coal Handling	-	894
Boiler Feedwater Treatment	-	193
General Facilities	-	1,210
Total In-Plant Power Consumption	20,687	14,220
Net Plant Efficiency	44.10	42.49

[a] In the METC model, auxiliary power was estimated as a percentage of gross power output. In the new model, the components of auxiliary power are disaggregated and separately estimated.

[b] Excluding reheat.

In the original METC model air leakage from the CerHx was not taken into account. An air leakage rate of 0.5 percent was considered in the modified model, which imposes an efficiency penalty on the EFCC system. The total effect of differences in the performance models and input assumptions results in a net plant efficiency of 42.49 for the modified model compared to 44.1 for the original METC model. As will be illustrated in Section 9.3 the model will yield other efficiency estimates depending on the input assumptions.

9.2.2 Environmental Results

A higher flue gas flow rate in the case of the modified versus the original model results in a slightly higher limestone and water consumption. The sulfur conversion in the original METC model was 95 percent compared to 90 percent in the modified model, which results in an increase in the SO_2 emissions for the modified model. However the user may specify other SO_2 removal efficiencies. Even though a higher coal flow is required for the modified model, the CO_2 emissions are lower than that for the original METC model because 99 percent, rather than 100 percent, carbon conversion was considered for the modified model.

In the original METC model 100 percent of the coal ash was assumed to be discharged in the form of slag. In the modified model, the percent of ash removed as slag is a user-specified parameter, which was assumed to be 65 percent based upon typical values reported by EPA. Others estimate possibly lower values. Therefore, the slag discharge for the modified model was calculated to be lower than that for the original METC model. The fraction of coal ash which is not converted to slag exits the combustor in form of fly ash with a size distribution which is characteristic of the coal combustor employed. The fly ash is subsequently removed in the slag screen, the ceramic heat exchanger, and the fabric filter, each of which have a characteristic removal efficiency. The fraction of flyash which is removed by these process areas are recovered as collected flyash and the rest is carried with the stack gas in the form of suspended particulate matter. Based on the particle removal efficiencies for the slag screen, ceramic heat exchanger, and the fabric filter given in Table 3.7, the collected flyash for the new EFCC model is 6,426 lb/hr.

Historically sludge from FGD usually consists of equal parts of water and solids on a mass basis. More recent FGD systems tend to be designed to produce sludge with low water content. The original METC model did not incorporate water in the sludge. Therefore, the original METC model underestimated the total sludge mass flow.

The thermal efficiency of the EFCC system is higher than for a conventional coal fired power plant. Therefore, the emissions for the EFCC system tend to be lower per unit of electricity produced. Since coal fired power plants are a major source of CO_2 emissions, the lower emissions of the EFCC system per unit of electricity produced are particularly significant with regard to CO_2 emissions. The CO_2 emissions from the new EFCC model is 1.6 lb/kWh compared to 1.94 lb/kWh for a conventional coal fired power plant with 99.5 percent coal conversion and 608 MW net plant power output (EPRI, 1988).

9.3 Sensitivity Analysis Results

A total of thirteen performance and design parameters were varies as part of a series of 55 sensitivity analysis runs. These analyses focused on four major process areas: (1) combustor; (2) ceramic heat exchanger; (3) gas turbine; and (4) environmental control. For each of these four major areas, the sensitivity analysis results are presented and discussed.

9.3.1 Combustor

Sensitivity analyses for the combustor focused upon variation of assumptions regarding radiative heat loss, heat duty to the combustor water wall, carbon conversion, and auxiliary air flow rate.

9.3.1.1 Combustor radiative heat Loss

The temperature of combustion in the coal combustor is approximately 2700 ^oF, which is high enough to cause heat loss due to radiation. Radiative heat loss is proportional to temperature raised to the fourth power. The combustor heat loss also depends upon the surface area to volume ratio of the combustor, and the convective and conductive heat loss increases with an increase in this ratio (Micheli, 1995). A full scale commercial combustor for the EFCC system has not yet been developed and demonstrated. Therefore, there is uncertainty regarding the exact magnitude of heat loss from the combustor.

The heat loss from the combustor was varied from zero to five percent of the heat of combustion reaction in steps of one percent. Figure 9.1 shows the variation in the gas and steam turbine output with increasing combustor heat loss. Since the mass flow rate of air to the gas turbine and the temperature of the air at the gas turbine expander inlet is constant, the gas turbine power output is independent of the combustor parameters. Therefore, the gas turbine power output does not change with an increase in the combustor heat loss. Increase in heat loss from the combustor for a constant coal flow rate would reduce the temperature of the flue gas coming out of the combustor, which would reduce the heat input to the steam cycle through the HRSG and the steam cycle power output. But the steam turbine power output is seen to increase with an increase in the combustor heat loss. In the ASPEN performance model, an increase in the combustor heat loss is compensated by an increase in the coal flow rate. The combustor outlet temperature is specified to enable sufficient heat transfer to the gas turbine in the CerHx in the gas turbine flowsheet. The auxiliary air flow to the combustor also increases in proportion to the coal flow rate. Therefore the flue gas flow rate from the combustor increases. Since the heat duty of the CerHx remains constant, the mass flow rate and the temperature of the flue gas entering the HRSG increases with an increase in combustor heat loss. This increases the heat input to the steam cycle and the power output from the steam cycle.

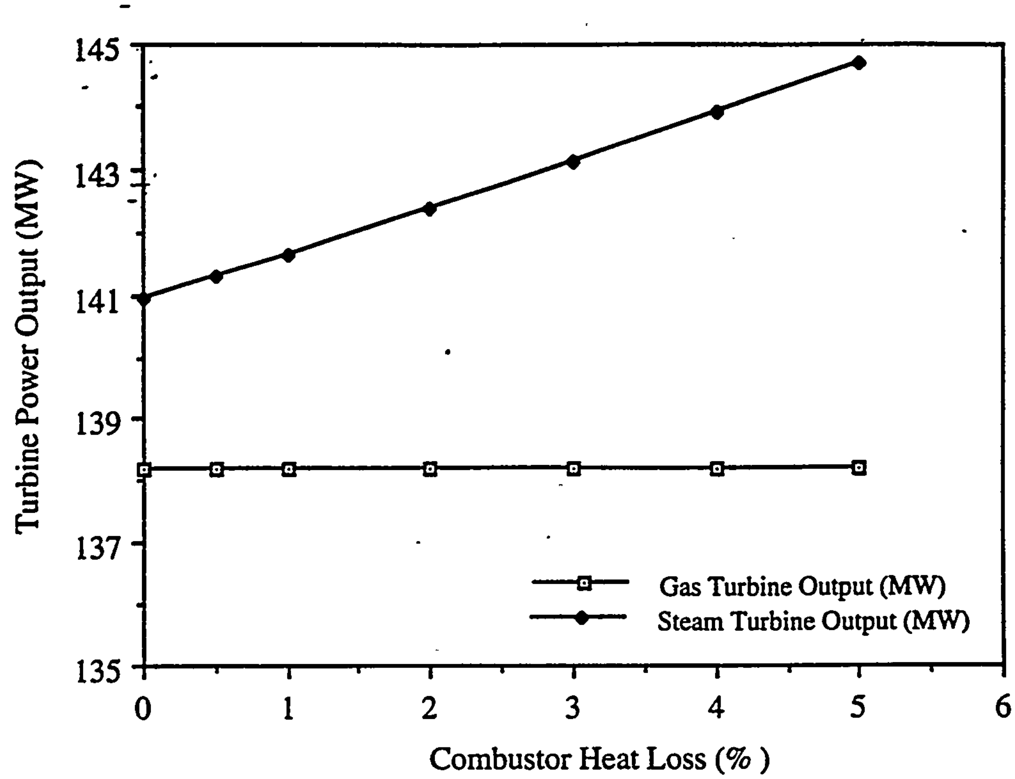


Figure 9.1. Gas and Steam Turbine Power Output versus Combustor Heat Loss.

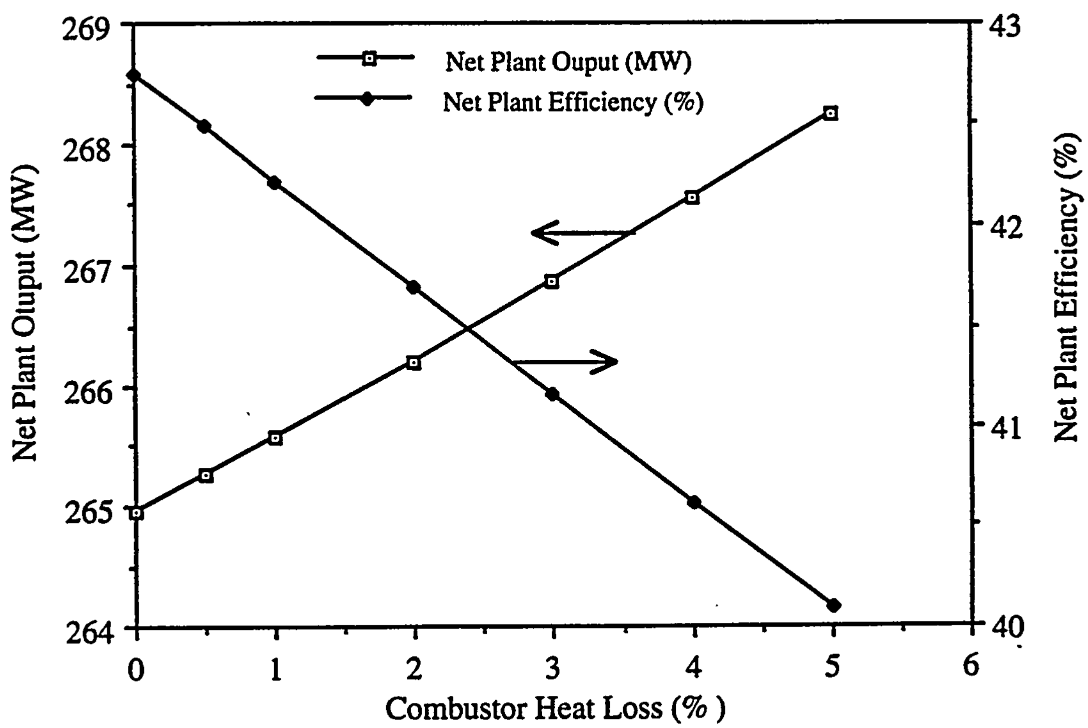


Figure 9.2. Net Plant Power Output and Net Plant Efficiency versus Combustor Heat Loss.

Figure 9.2 shows the change in the net plant power output and net plant efficiency with increase in the combustor heat loss. The net plant power output increases by 3.3 MW as the heat loss increases from zero to five percent. The steam turbine power output increases by 3.73 MW; however the net plant power output increases is less than this by 0.43 MW due to an increase in the auxiliary power consumption. As expected, the net plant efficiency decreases substantially from 42.75 for no combustor heat loss to 40.09 for five percent combustor heat loss.

9.3.1.2 Heat duty to combustor water wall

In the base case design, a dry walled combustor was assumed. However, a fraction of the water from the economizer inlet may be sent through a combustor water wall to capture some of the combustor heat losses and transfer it to the steam cycle. Such a design would increase the plant power output by capturing some of the combustor heat that would be lost to the surroundings and converting it to useful power in the steam cycle. The combustor heat directed to the water wall was varied from zero for the base case to three percent of the heat of coal combustion in steps of one percent.

Figure 9.3 shows the change in the gas and steam turbine power output with increasing heat directed to the water wall. Since the gas turbine section parameters are not affected by increasing the combustor heat to water wall, the gas turbine power output remains constant at 138 MW. With an increase in the combustor heat directed to the water wall, the mass flow of coal and auxiliary air increases to compensate for that heat in order to achieve the desired combustor outlet flue gas temperature. Therefore, the mass flow rate of flue gas at the combustor outlet increases. Since the heat duty of the CerHx is constant, the temperature of the flue gas entering the HRSG increases as increasing heat is directed to the water walls. The higher mass flow rate and temperature of the flue gas entering the HRSG increases the thermal input to the steam cycle, which increases the power output of the steam cycle. The heat captured in the combustor water wall is also transferred to the steam cycle, which further increases the steam turbine power output. The steam turbine power output increases from 141.3 MW to 155 MW for an increase of the heat directed to the water walls from zero percent to three percent of the combustor heat duty.

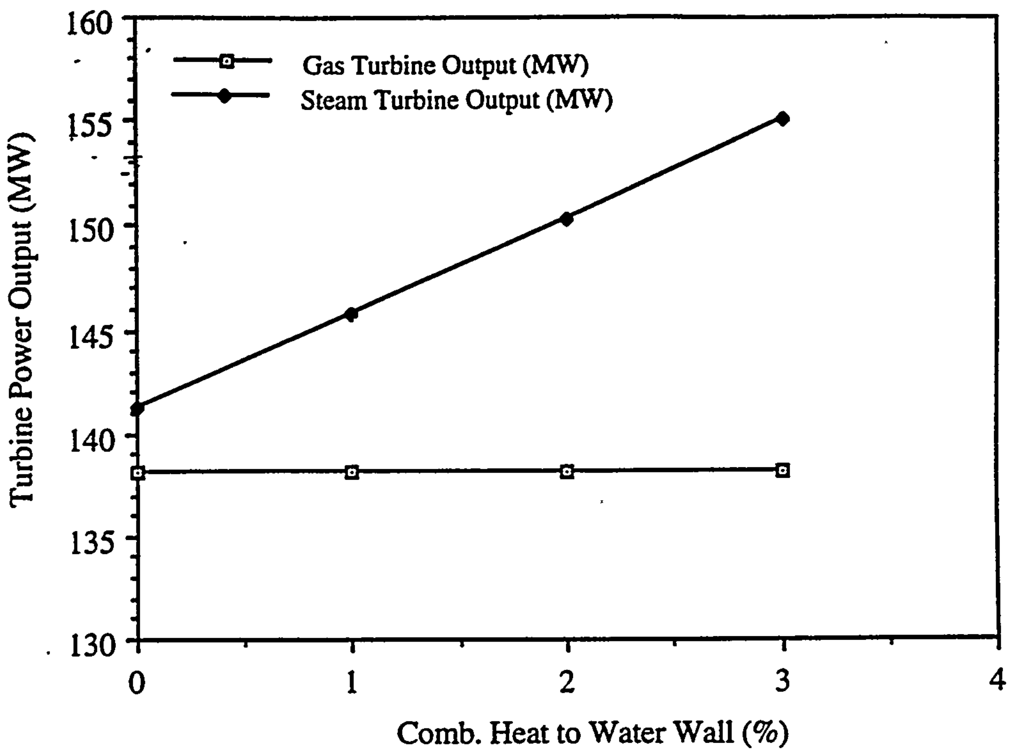


Figure 9.3 Gas and Steam Turbine Power Output versus Combustor Heat Duty to Water Wall.

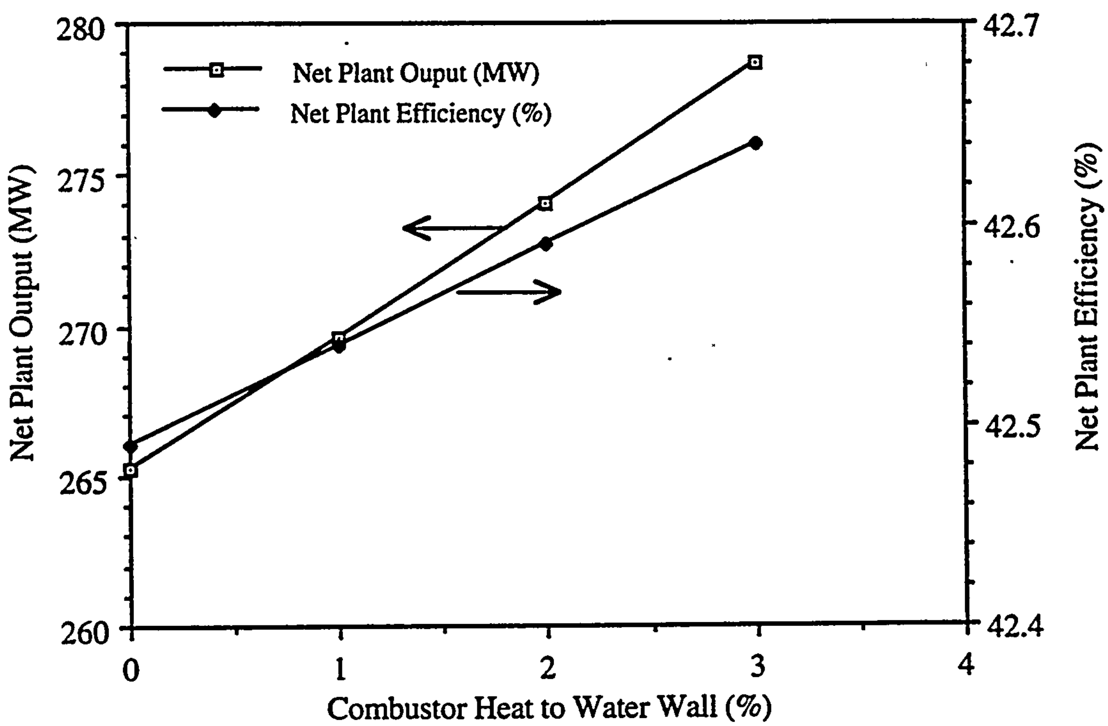


Figure 9.4 Net Plant Power Output and Net Plant Efficiency versus Combustor Heat Duty to Water Wall.

The change in the net plant power output and the net plant efficiency with increasing heat directed to the combustor water wall is shown graphically in Figure 9.4. Since the steam turbine power output increases with increasing heat directed to the water wall, the net plant power output increases from 265 MW for no heat directed to water wall to 278 MW for three percent heat directed to the water wall. The coal flow increases to compensate for the combustion heat directed to the water wall. Since the steam cycle is slightly more efficient than the gas turbine, the heat transferred to the steam cycle is used more efficiently in this case than if it had been directed to the gas turbine. Therefore, with an increase in the combustion heat directed to the water wall, the net plant efficiency increases from 42.49 for the base case to 42.64 for the case where three percent of the heat of coal combustion is directed to the water wall.

9.3.1.3 Carbon conversion

Figure 9.5 shows the variation of the gas turbine and steam turbine power output with increase in the carbon conversion. Since the parameters for the gas turbine design are specified and the gas turbine flowsheet is executed first in the ASPEN performance model, the gas turbine output remains unchanged with a change in the carbon conversion. The steam turbine power output decreases slightly, by 0.26 MW, over the range of carbon conversions considered. An increase in the coal conversion reduces the coal and auxiliary air flow requirement to achieve the desired combustor outlet temperature. Therefore, the flue gas flow rate at the combustor outlet is reduced with an increase in the carbon conversion. Since the heat duty of the CerHx remains constant, the temperature of the flue gas entering the HRSG is also reduced slightly with an increase in carbon conversion. The thermal input to the steam cycle is therefore reduced, which reduces the steam turbine power output.

Figure 9.6 shows the change in the net plant power output and the net plant efficiency with an increase in the carbon conversion. For an increase in the carbon conversion from 99 percent to 100 percent, the net plant power output decreases by 0.23 MW, and the auxiliary power loads decrease by 0.03 MW. The coal mass input to the plant in the form of coal decreases from 189,533 lb/hr to 187,117 lb/hr with an increase in the carbon conversion from 99 percent to 100 percent, whereas the net plant power output is reduced insignificantly. Therefore, with an increase in the carbon conversion of coal from 99 percent to 100 percent, the net plant efficiency increases from 42.5 to 43.0 percent.

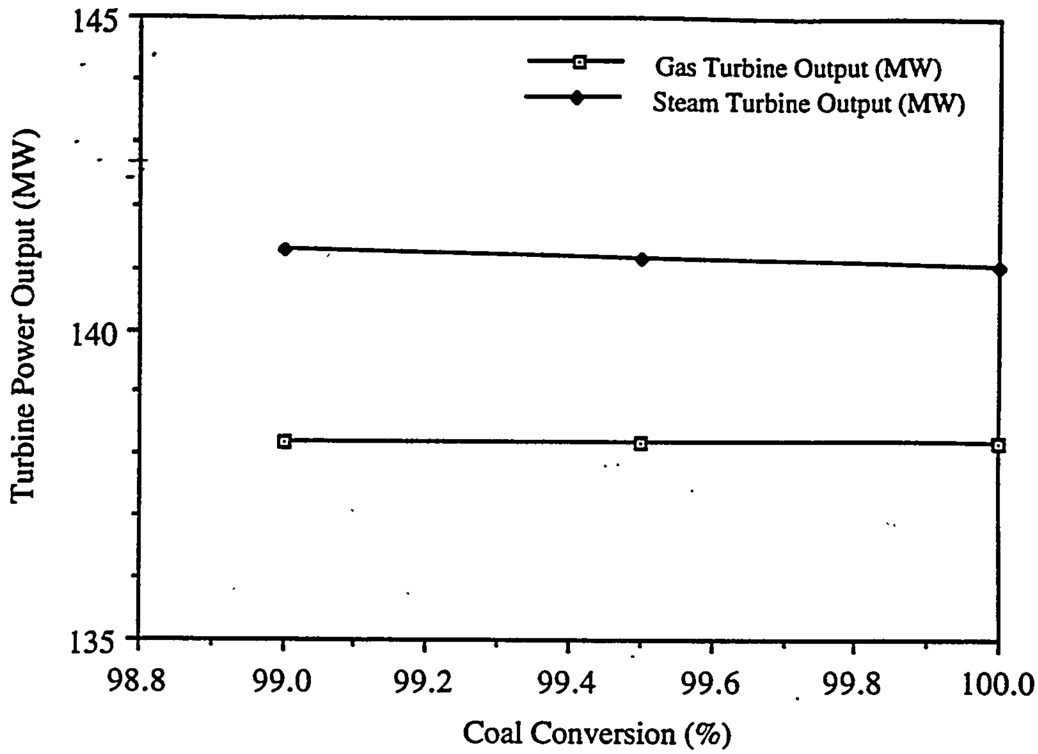


Figure 9.5. Gas and Steam Turbine Power Output versus Coal Conversion in Combustor.

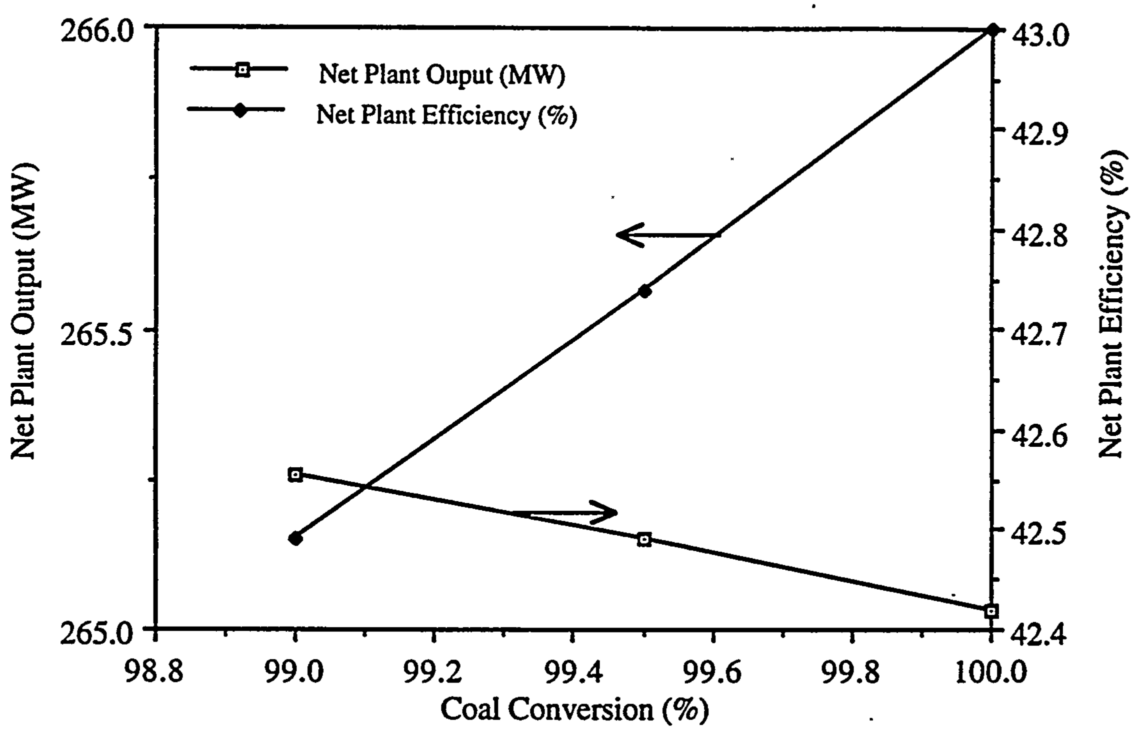


Figure 9.6. Net Plant Power Output and Net Plant Efficiency versus Coal Conversion in Combustor.

9.3.1.4 Auxiliary air flow rate

Auxiliary air is required to pneumatically transport pulverized coal to the coal combustor. The mass flow of air required per unit mass flow rate of coal depends on the design of the coal handling system. The mass flow of air per unit mass of coal transported was varied from 0.5 lb air/lb coal to 1.0 lb air/lb coal in steps of 0.1 lb air/lb coal to study the effect of auxiliary air mass flow rate on plant performance parameters.

The change in the gas and steam turbine power output with an increase in the auxiliary air flow rate is shown in Figure 9.7. The gas turbine power output remains unchanged at 138 MW with an increase in the auxiliary air flow rate. In order to achieve the desired combustor flue gas outlet temperature, as the auxiliary air flow rate increases, the coal flow also increases. Since the heat duty of the CerHx is constant, the temperature and the mass flow rate of the flue gas entering the HRSG increases. Therefore the heat input to the steam cycle increases, which increases the steam cycle power output as the auxiliary air-to-coal ratio is increased.

Figure 9.8 shows the variation in the net plant power output and the net plant efficiency with an increases in the auxiliary air flow. The total plant output increases slightly as the auxiliary air ratio is increased, inspite of a modest increase of 0.44 MW in auxiliary power consumption. The latter is due to increased flows of coal, auxiliary air, flue gas, and steam. The net plant efficiency decreases by 0.24 percentage points. Thus, the increase in power output is outweighed by the increase in coal consumption due to an increase in the thermal input and the auxiliary power consumption.

9.3.2 Ceramic Heat Exchanger

Sensitivity analyses for the CerHx focused upon variation of air leakage rate, radiative heat loss, and heat exchanger effectiveness.

9.3.2.1 Air leakage rate

The compressor discharge air flows through the CerHx tubes and is heated by the hot combustor flue gas flowing through the CerHx shell side. Leakage of high pressure air from the tube side to the shell side of the CerHx imposes an efficiency penalty on the gas turbine, and increases the flue gas flow rate through the HRSG and backend pollution control devices. To evaluate the effect of the CerHx air leakage on key plant output parameters, the high pressure air leakage was varied from zero percent to three percent of the CerHx inlet air flow.

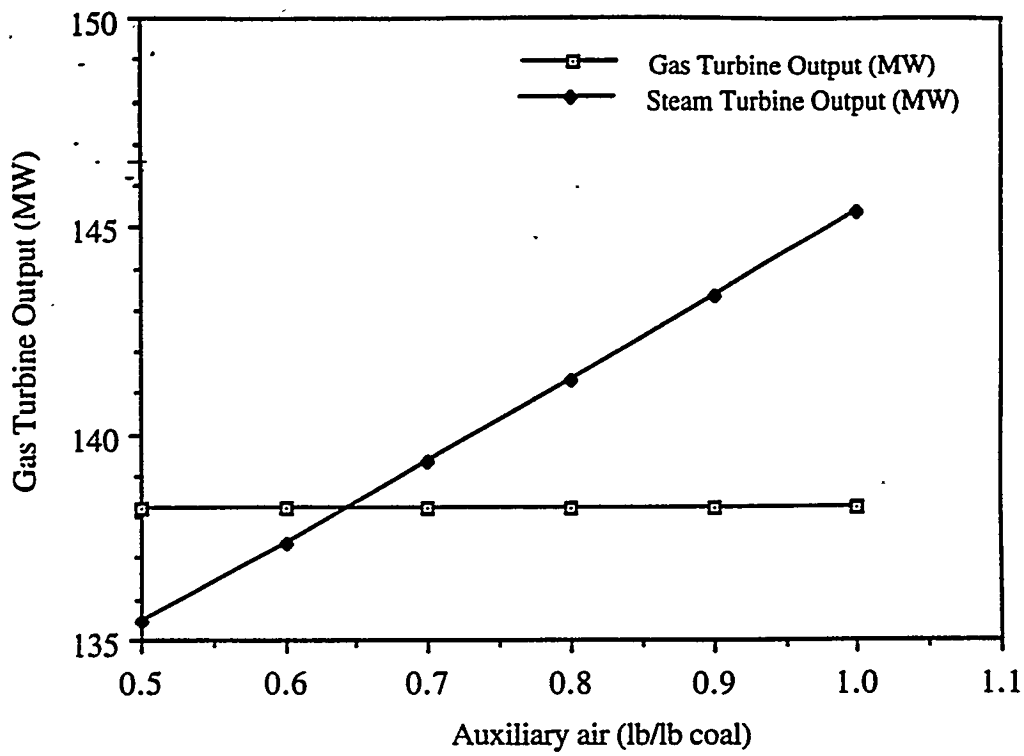


Figure 9.7. Gas and Steam Turbine Power Output versus Auxiliary Air Flow.

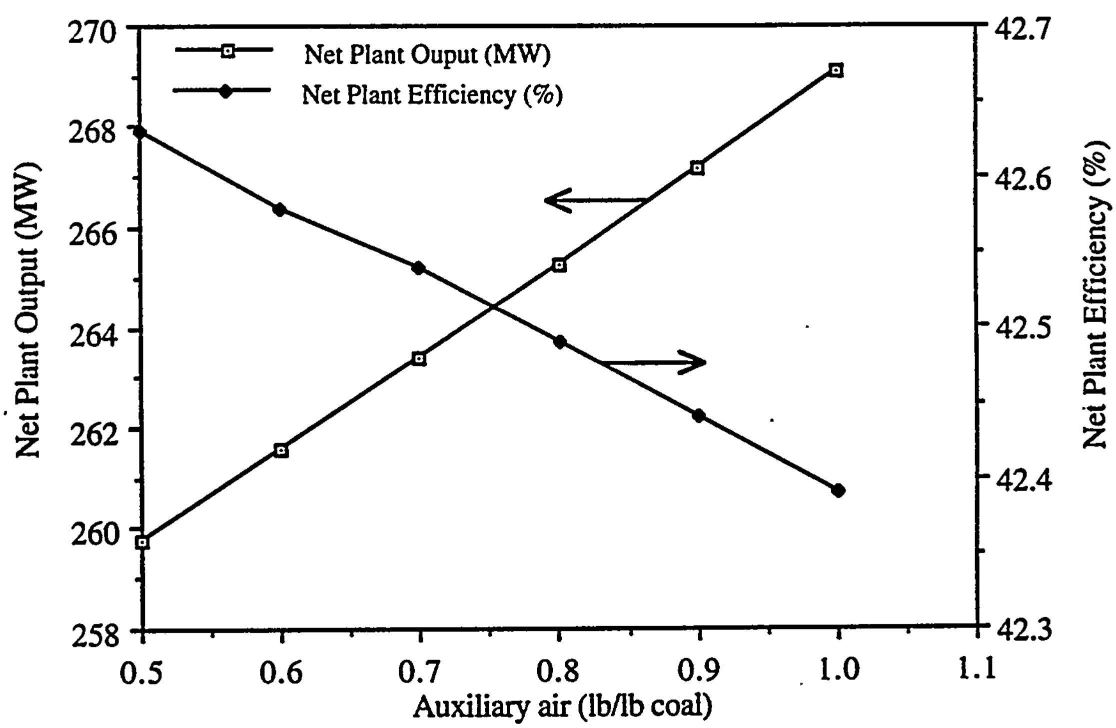


Figure 9.8. Net Plant Power Output and Net Plant Efficiency versus Auxiliary Air Flow.

Figure 9.9 shows the change in the gas turbine and the steam turbine power output with an increase in the CerHx air leakage. Since the air flow at the gas turbine first stage expander inlet is choked, with an increase in the CerHx air leakage the air flow at the turbine inlet remains constant at 3,081,000 lb/hr. To compensate for the CerHx air leakage, the gas turbine compressor air flow increases by 100,000 lb/hr for three percent air leakage. The compressor work requirement, therefore, increases which reduces the gas turbine power output by 4.4 MW.

The high pressure air leaking from the tube side to the shell side is at a lower temperature than the flue gas flowing through the shell side of the CerHx. Therefore, an increase in air leakage reduces the temperature of the CerHx flue gas but increases its mass flow rate. Thus, heat input to the steam cycle increases with an increase in the air leakage, which increases the steam turbine output by 2.25 MW for three percent air leakage versus no air leakage.

Changes in the net plant power output and the net plant efficiency with increasing air leakage are shown in Figure 9.10. Since the flue gas flow rate increases with an increase in the air leakage, the corresponding ID fan power requirement and the steam cycle auxiliary loads increase. The reduction in the gas turbine power output for an increase in air leakage is greater than the corresponding increase in the steam turbine power output. Therefore, the net plant power output decreases by 2.3 MW for three percent air leakage. The net plant efficiency also decreases by 0.7 percentage points over the range of air leakages considered.

9.3.2.2 Radiative heat loss

Heat energy is lost through the CerHx walls due to radiation, which leads to a lower flue gas exit temperature than if the heat loss had not occurred. Thus, the heat input to the steam cycle and the steam turbine power output are both decreased. Radiative heat loss from the CerHx, therefore, impose an efficiency penalty on the EFCC system.

The radiative heat loss from the CerHx was varied from zero to five percent of the heat duty required to heat the tube side high pressure air for a specified gas turbine inlet temperature. Figure 9.11 shows the variation in the gas and steam turbine power output with a change in the CerHx heat loss. Since the heat transferred from the shell side flue gas to the tube side high pressure air is constant for the specified gas turbine parameters, the gas turbine power output remains constant at 138 MW. An increase in the CerHx heat loss from zero to five percent reduces the temperature of the flue gas entering the HRSG from 1520° F to 1460° F. Therefore, the heat input to the steam cycle decreases. Thus, the steam turbine power output decreases by 10 MW over the range of CerHx heat losses considered.

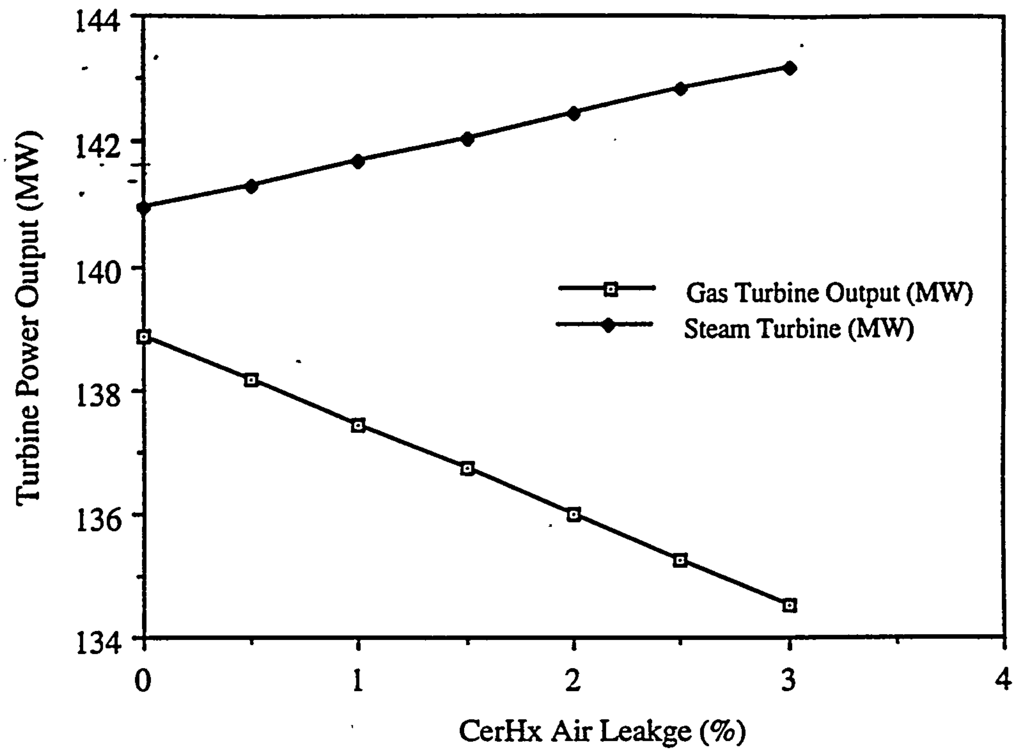


Figure 9.9. Gas and Steam Turbine Power Output versus Ceramic Heat Exchanger Air Leakage

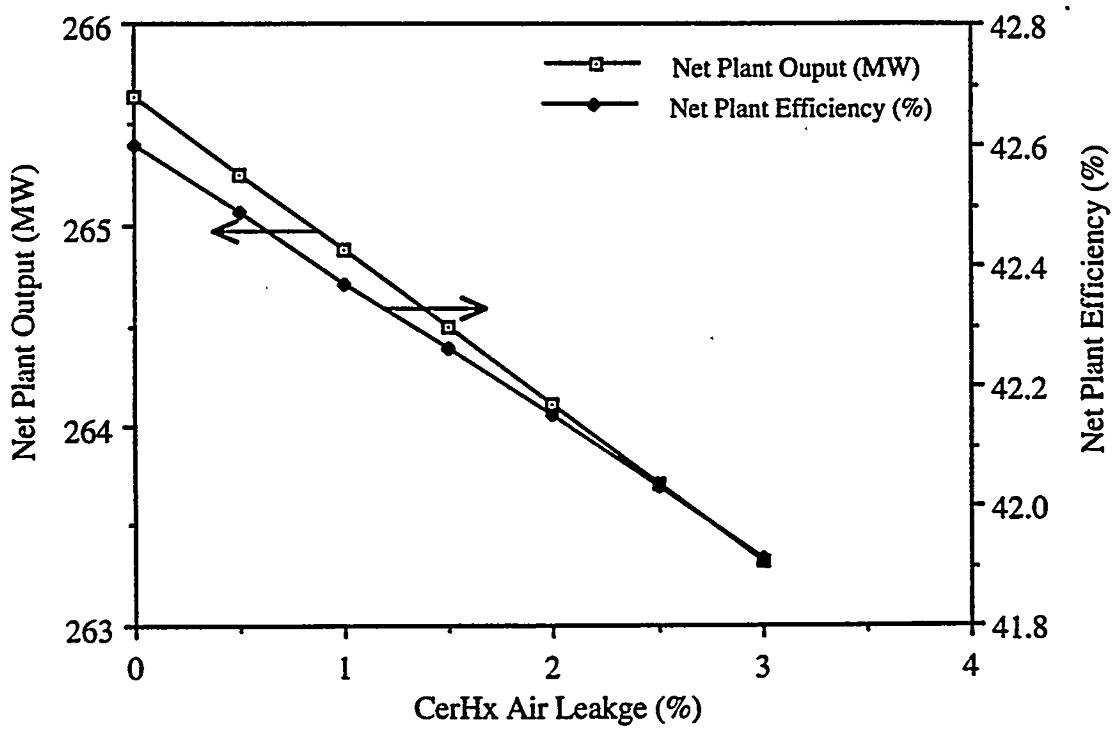


Figure 9.10. Net Plant Power Output and Net Plant Efficiency versus Ceramic Heat Exchanger Air Leakage

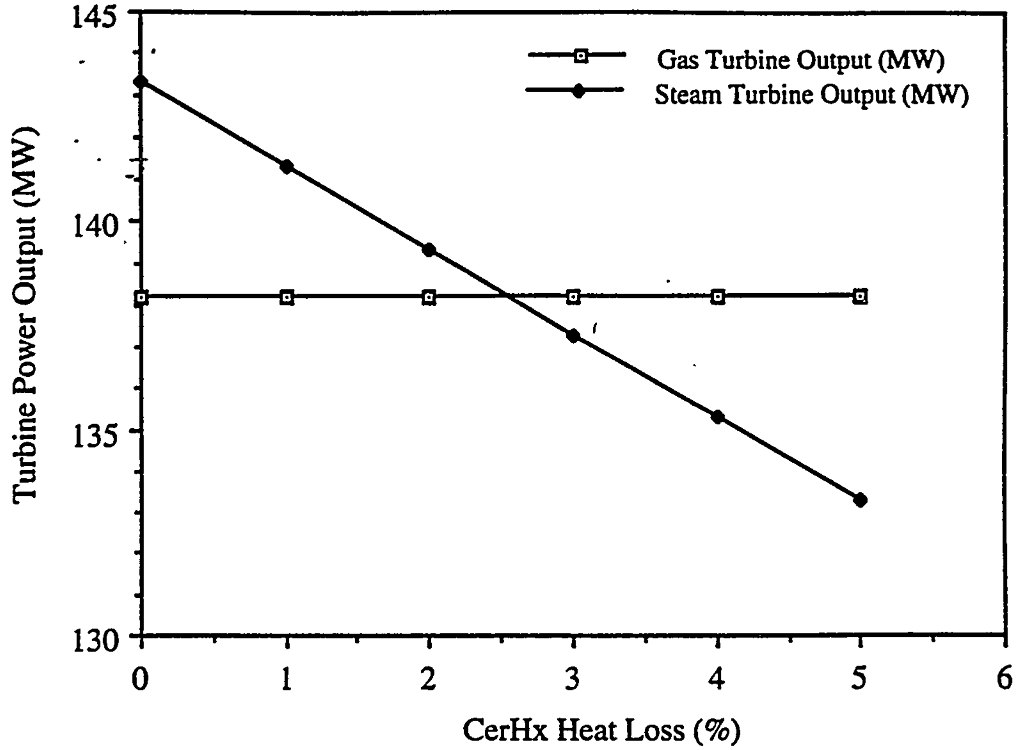


Figure 9.11. Gas and Steam Turbine Power Output versus Ceramic Heat Exchanger Heat Loss.

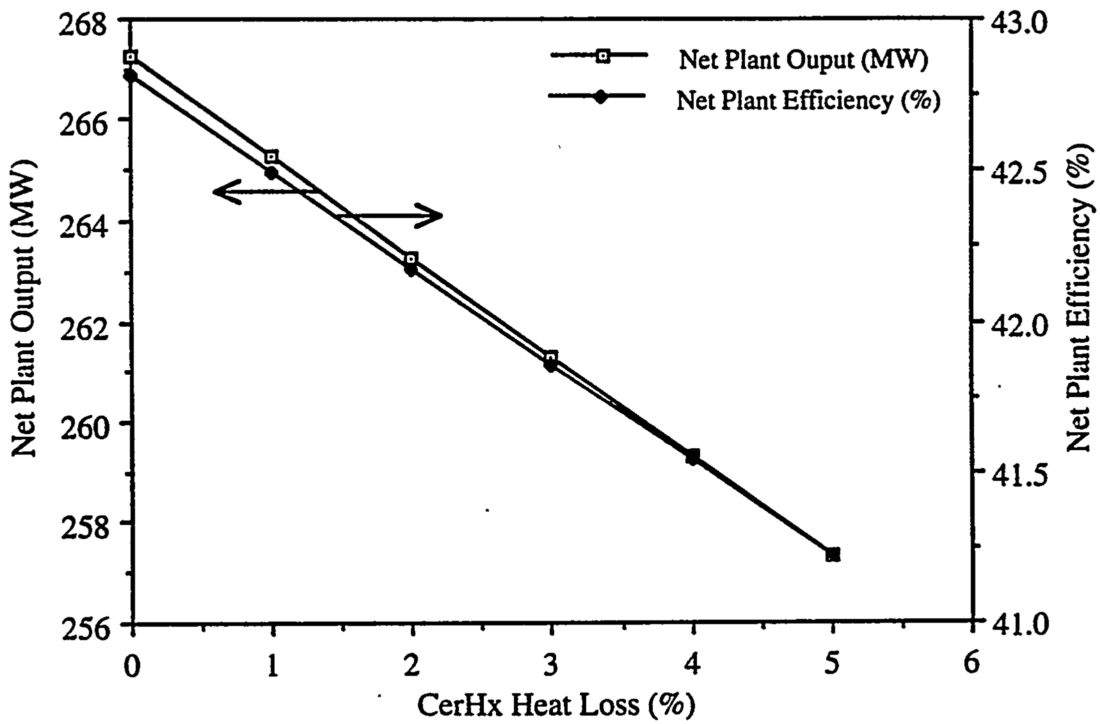


Figure 9.12. Net Plant Power Output and Net Plant Efficiency versus Ceramic Heat Exchanger Heat Loss.

The change in net plant power output and the net plant efficiency with increasing CerHx heat loss is shown graphically in Figure 9.12. Since the heat input to the steam cycle decreases with an increase in the CerHx heat loss, the steam flow rate in the steam cycle decreases proportionally, which reduces the steam cycle auxiliary load by 0.09 MW for five percent CerHx heat loss. With an increase in CerHx heat loss, the gas turbine output remains the same, but the steam turbine power output decreases substantially. Therefore, the net plant power output decreases by 9.9 MW. Since the coal input remains constant with a change in the CerHx heat loss, the net plant efficiency decrease by 1.6 percentage points from the lower to the upper limit of the CerHx heat loss.

9.3.2.3 Heat exchanger effectiveness

The CerHx effectiveness is a parameter which reflects the efficiency with which heat is transferred from the shell side flue gas to the tube side high pressure air. A higher CerHx effectiveness would require a lower CerHx flue gas inlet temperature for specified high pressure air inlet and exit temperatures. Similarly, as effectiveness increases, for a given flue gas inlet temperature, it would be possible to obtain a higher high pressure air exit temperature. The ASPEN performance model for the EFCC system has been developed such that the high pressure air inlet and exit temperatures are fixed and the flue gas inlet temperature is varied to achieve the desired CerHx effectiveness. The CerHx effectiveness was varied from 0.85 to 0.9 in steps of 0.01 and the plant performance parameters were plotted against the change in the CerHx effectiveness.

Figure 9.13 shows the change in the gas and steam turbine power output with a change in the CerHx effectiveness. Since the gas turbine section parameters are independent of the CerHx effectiveness, the gas turbine power output remains constant at 138 MW. An increase in the CerHx effectiveness enables the temperature of the flue gas exiting the coal combustor to be lowered from 2,703 ^{o}F for a CerHx effectiveness of 0.85 to 2,593 ^{o}F for a CerHx effectiveness of 0.9. Since the combustor flue gas exit temperature is decreased, the coal and auxiliary air flow is also decreased. Irrespective of the effectiveness, the heat duty of the CerHx remains constant, which leads to a decrease in the temperature of the flue gas entering the HRSG from 1,506 ^{o}F for a CerHx effectiveness of 0.85 to 1,381 ^{o}F for an effectiveness of 0.9. Thus, the flue gas flow rate and temperature decrease as CerHx effectiveness increases. This leads to a reduced heat input to the steam cycle. Therefore, the steam turbine power output decreases substantially by 23 MW from the lower to the higher value of the effectiveness.

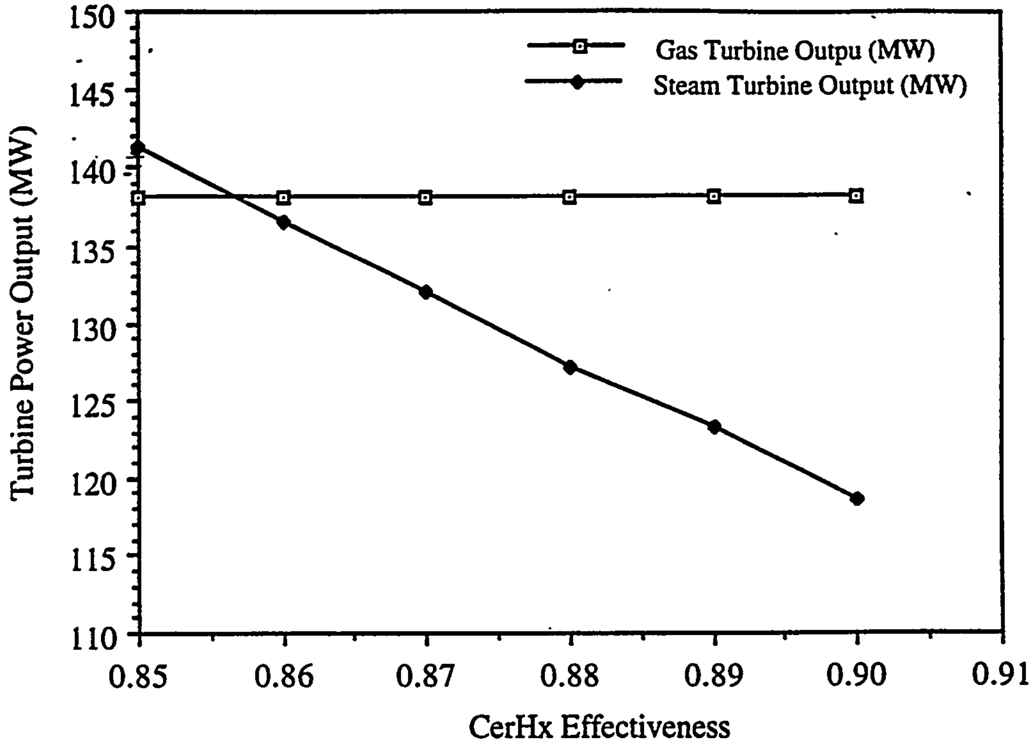


Figure 9.13. Gas and Steam Turbine Power Output versus Ceramic Heat Exchanger Effectiveness.

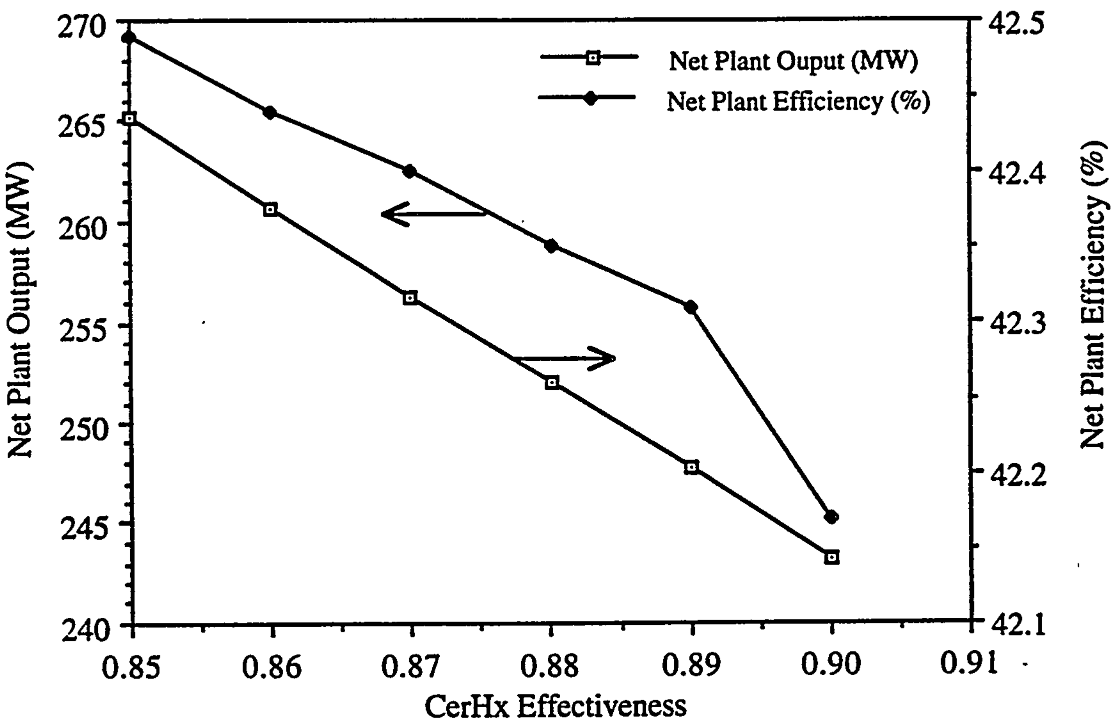


Figure 9.14. Net Plant Power Output and Net Plant Efficiency versus Ceramic Heat Exchanger Effectiveness.

Since the coal, auxiliary air, flue gas, and steam mass flows decrease with an increase in the CerHx effectiveness, the auxiliary loads decrease by 0.6 MW as the effectiveness increases. The change in net plant power output is dominated by the change in steam turbine output. It decreases by 21 MW with an increase in the CerHx effectiveness from the lower to the upper limit as shown in Figure 9.14. The figure also shows the change in the net plant efficiency with a change in the CerHx effectiveness. The steam cycle is slightly more efficient than the gas turbine for this particular system. Therefore, even though the plant thermal input decreases with an increase in the CerHx effectiveness, the reduction in only the steam cycle power output leads to decrease in plant efficiency of 0.32 percentage points. Alternatively, an increase in CerHx effectiveness can be used to increase the gas turbine inlet temperature for a given combustor flue gas temperature. Such a scenario would lead to an increase in overall plant efficiency.

9.3.3 Gas Turbine

Sensitivity analyses related to the gas turbine focused upon variation in the turbine inlet temperature, the steam injection rate, the water injection rate, and the ambient air temperature. Pressure ratio was also varied for a specified turbine inlet temperature.

9.3.3.1 Turbine inlet temperature and pressure ratio

The gas turbine efficiency and power output depends on the temperature and the pressure of the air at the turbine expander inlet. With an increase in the gas turbine inlet temperature and the compressor pressure ratio the gas turbine power output is expected to increase. The gas turbine model for the base case has been based upon a single GE Frame 7F heavy duty gas turbine with a turbine inlet temperature of 2,300 °F and a pressure ratio of 13.5. Sensitivity analysis was performed for a turbine inlet temperature of 2,350 °F and a pressure ratio of 15, and for a turbine inlet temperature of 2,400 °F with the pressure ratio ranging from 15 to 17.

From Figure 9.15 it can be seen that the gas turbine and the steam turbine power output increases substantially with an increase in the turbine inlet temperature and the pressure ratio. An increase in the gas turbine inlet temperature and pressure ratio from the base case values to a 2,350 °F turbine inlet temperature and 15 pressure ratio increases the gas turbine power output by 20 MW. At a turbine inlet temperature to 2,400 °F and a pressure ratio of 17, the power output is increased by 47 MW compared to the base case. An increase in the turbine inlet temperature also increases the efficiency of the gas turbine.

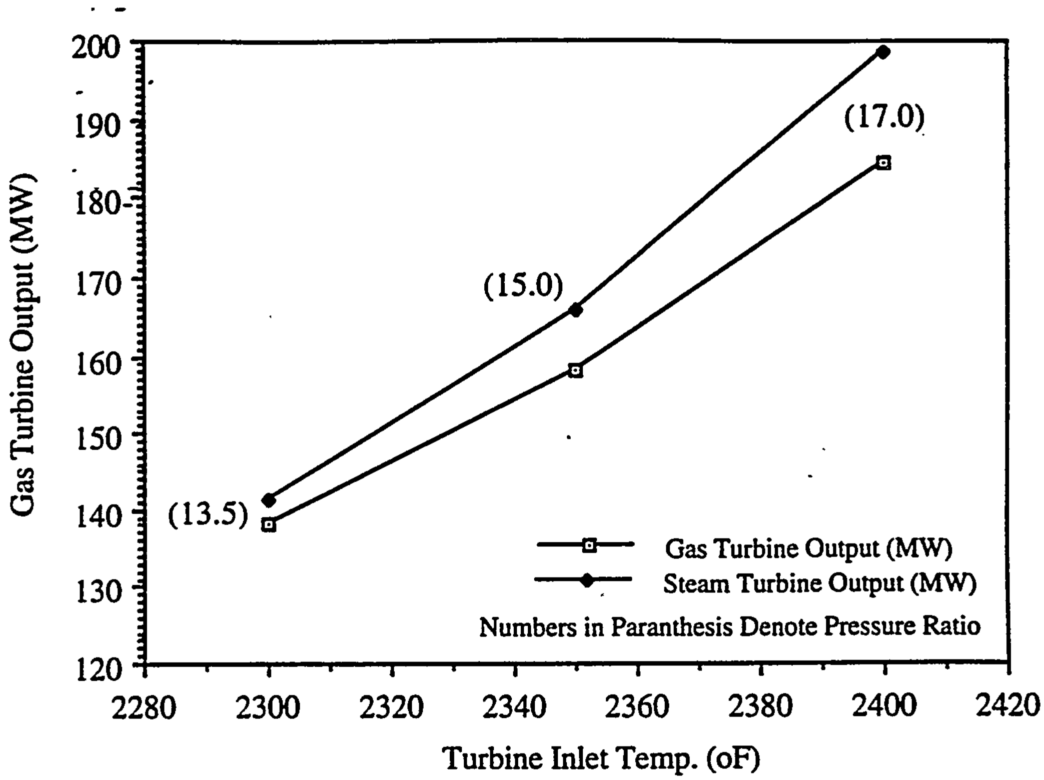


Figure 9.15. Gas and Steam Turbine Power Output versus Turbine Inlet Temperature and Pressure Ratio.

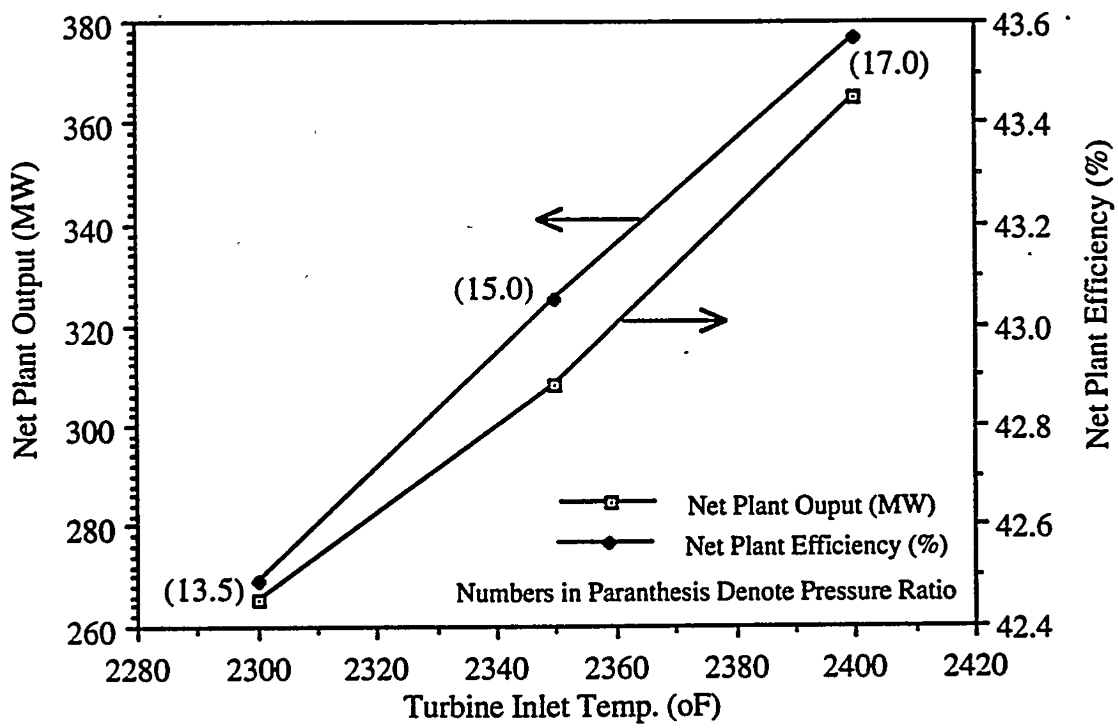


Figure 9.16. Net Plant Power Output and Net Plant Efficiency versus Turbine Inlet Temperature and Pressure Ratio.

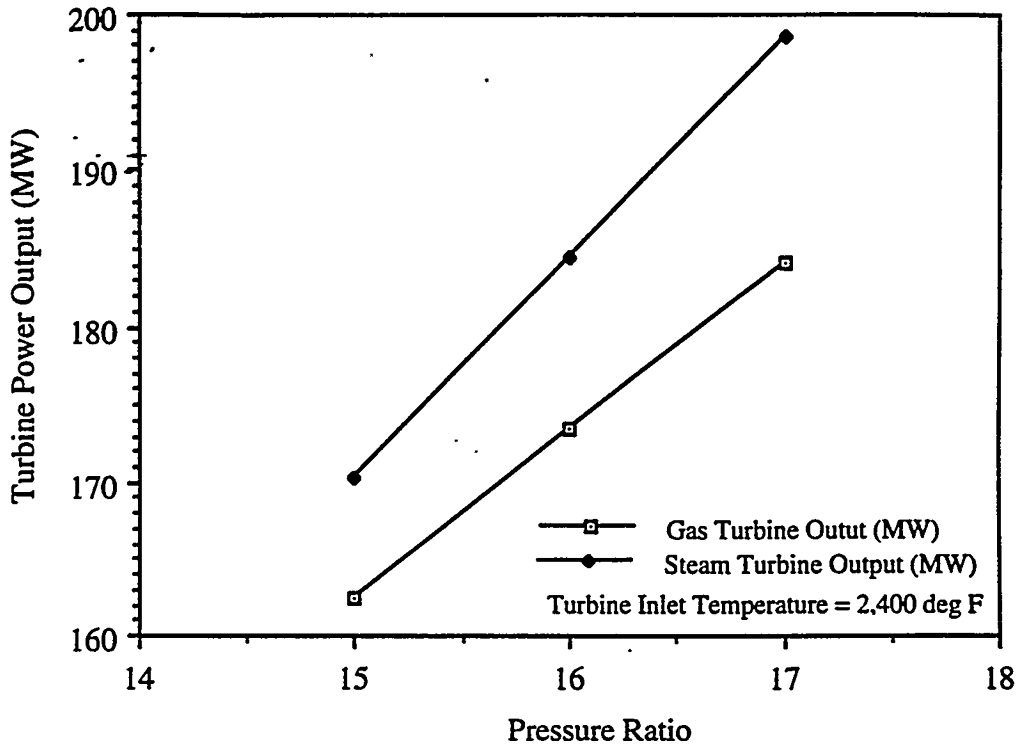


Figure 9.17. Gas and Steam Turbine Power Output versus Pressure Ratio.

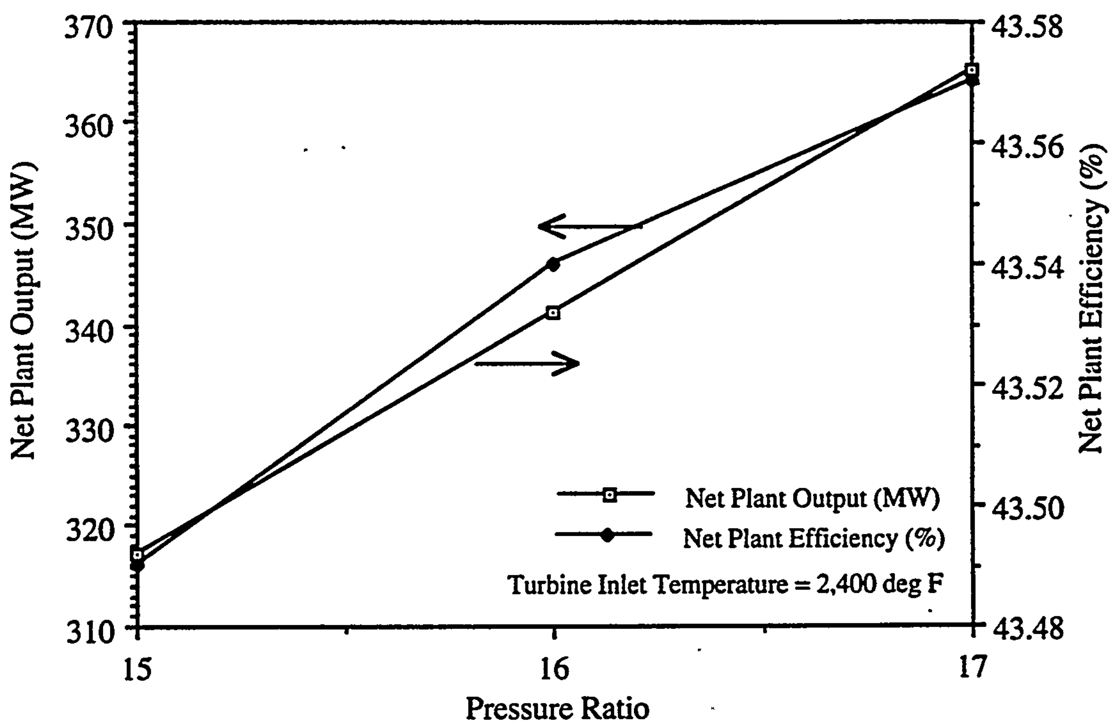


Figure 9.18. Net Plant Power Output and Net Plant Efficiency versus Pressure Ratio.

An increase in the pressure ratio increases the mass flow rate of air at the gas turbine expander inlet, thereby increasing the gas turbine power output. The increase in the gas turbine power output with an increase in the pressure ratio from 15 to 17 at a constant turbine inlet temperature of 2,400 °F is shown in Figure 9.17. At a higher turbine inlet temperature, and for a constant CerHx effectiveness, the temperature of the flue gas at the combustor exit must increase. Therefore, the temperature of the flue gas at the HRSG inlet increases, which increases the thermal input to the steam cycle and the steam turbine power output. At higher pressure ratios for a constant turbine inlet temperature, the turbine exhaust air flow rate to the combustor increases, which increases the flue gas flow rate at the HRSG inlet. Therefore, with an increase in the pressure ratio, the thermal input to the HRSG increases, which in turn increases the steam turbine power output. Figures 9.15 and 9.17 show graphically the increase in the steam turbine power output with an increase in the turbine inlet temperature and the pressure ratio.

Figures 9.16 and 9.18 show the increase in the net plant power output and the net plant efficiency with an increase in the turbine inlet temperature and pressure ratio. Improved gas turbine designs offer the potential to substantially increase both efficiency and power output of the EFCC. However, the ability to capture the benefits of improved gas turbine may be limited by temperature limitations of ceramic heat exchanger materials.

9.3.3.2 Steam injection rate

Steam injection from the steam cycle to the gas turbine increases the power output from the gas turbine by adding to the mass flow of the air at the turbine expander inlet. Steam diverted from the steam cycle leads to a reduction in steam cycle power output.

Figure 9.19 shows the change in the gas turbine and the steam turbine power output with increasing steam injection rate. The gas turbine power output increases by 53 MW and the steam turbine power output decreases by 50.6 MW for a steam injection rate of 500,00 lb/hr versus the base case.

The change in the net plant power output and the net plant efficiency with increasing rate of steam injection is shown graphically in Figure 9.20. The net plant power output increases by 2.25 MW for a steam injection rate of 400,000 lb/hr. For higher steam injection rates, there is a slight decrease in net plant output. Compared to the base case, the net plant efficiency increases by 0.12 percent points for 200,00 lb/hr steam injection. However, for higher steam injection levels, it decreases. At 500.000 lb/hr, the net plant efficiency decreases below the base case value by 0.15 percentage points.

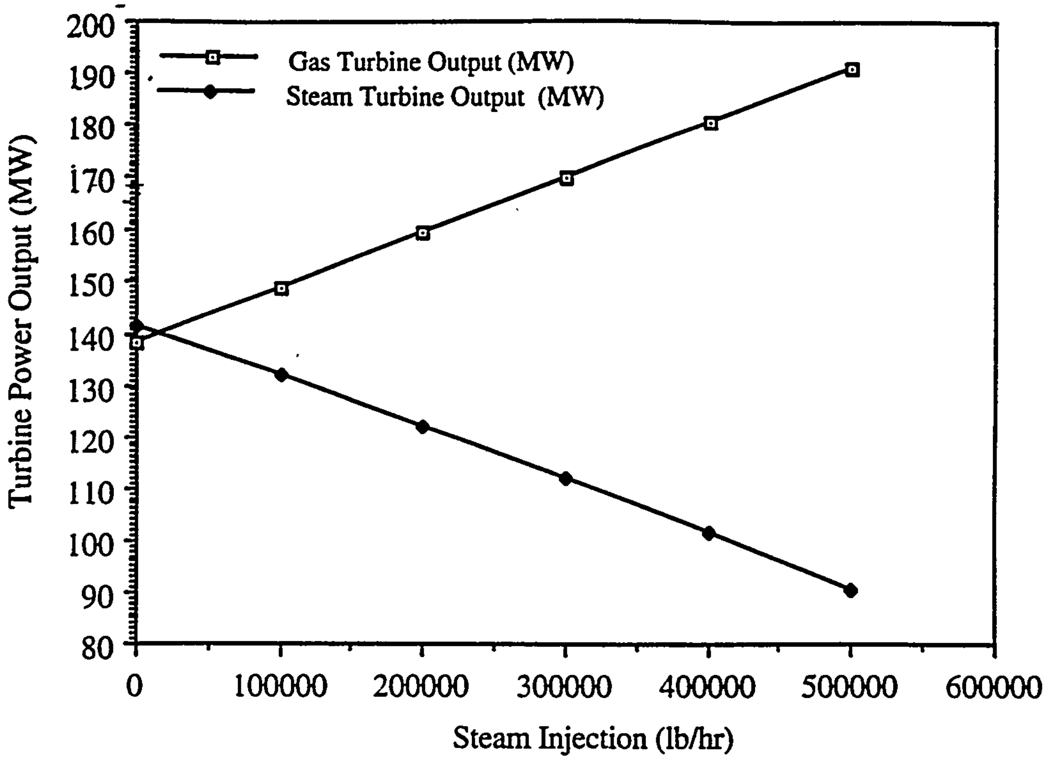


Figure 9.19. Gas and Steam Turbine Power Output versus Steam Injection Rate.

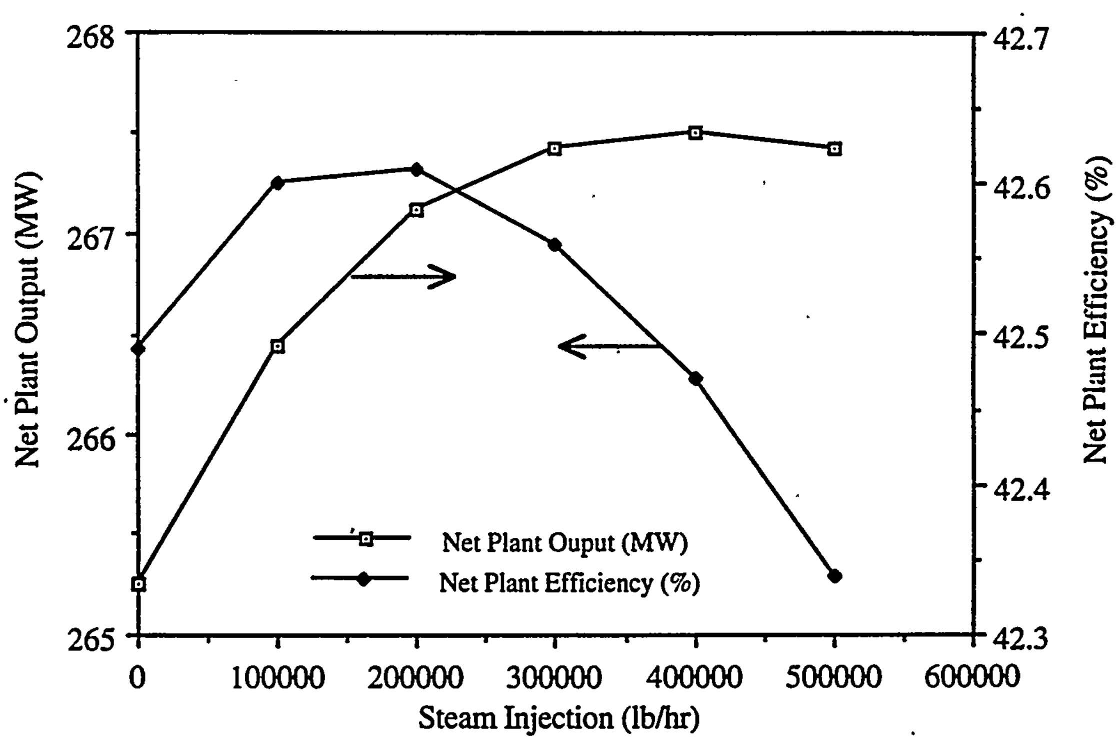


Figure 9.20. Net Plant Power Output and Net Plant Efficiency versus Steam Injection Rate.

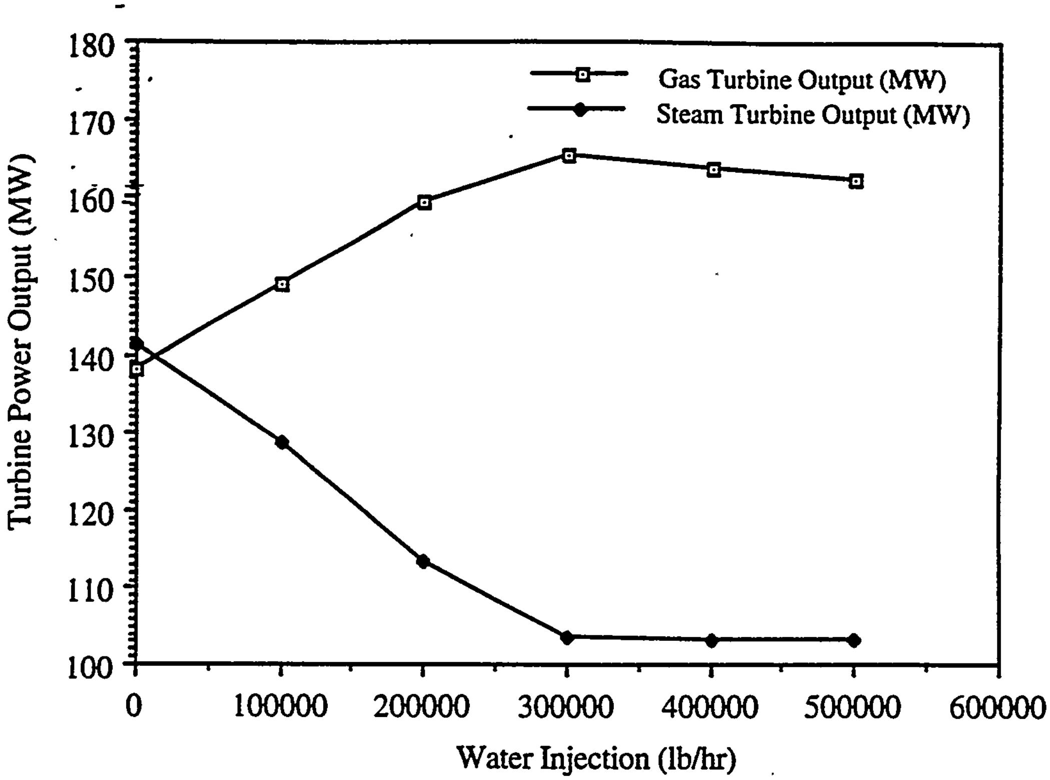


Figure 9.21. Gas and Steam Turbine Power Output versus Water Injection Rate.

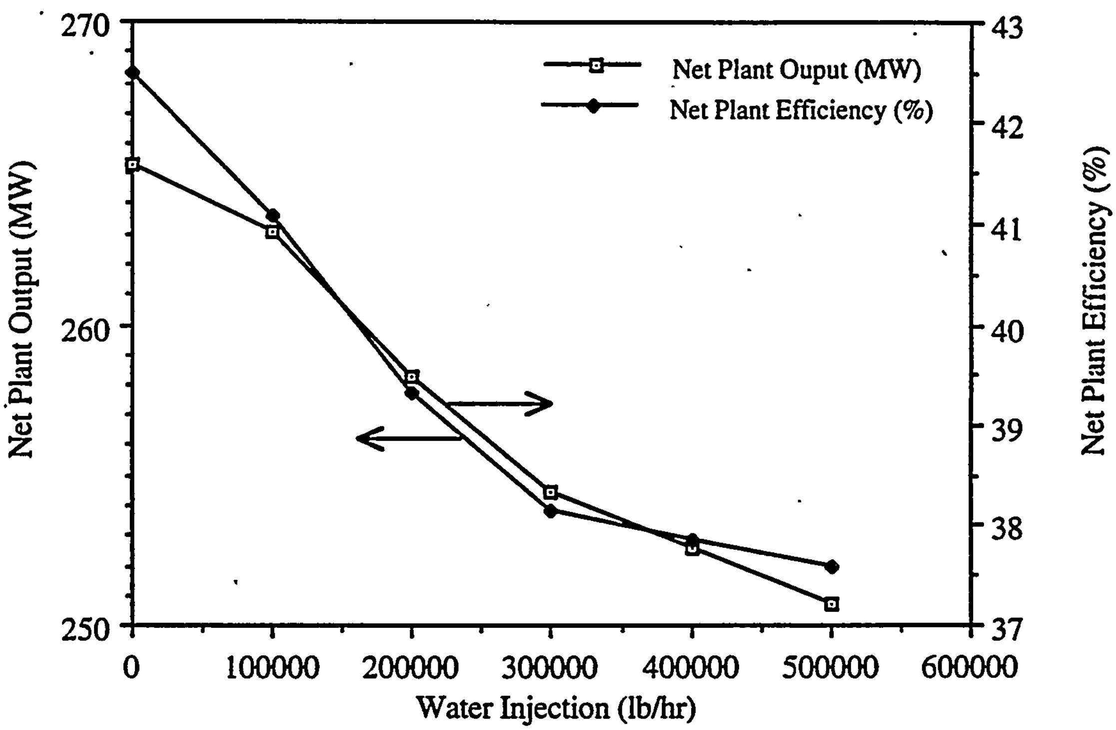


Figure 9.22. Net Plant Power Output and Net Plant Efficiency versus Water Injection Rate.

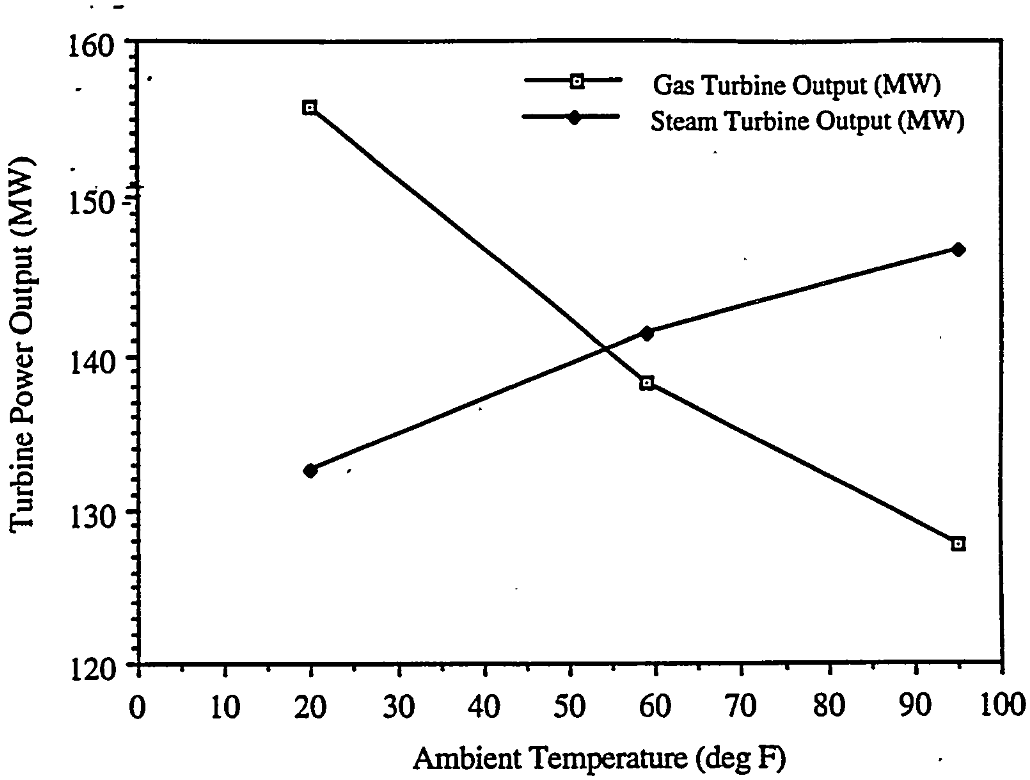


Figure 9.23.. Gas and Steam Turbine Power Output versus Ambient Temperature.

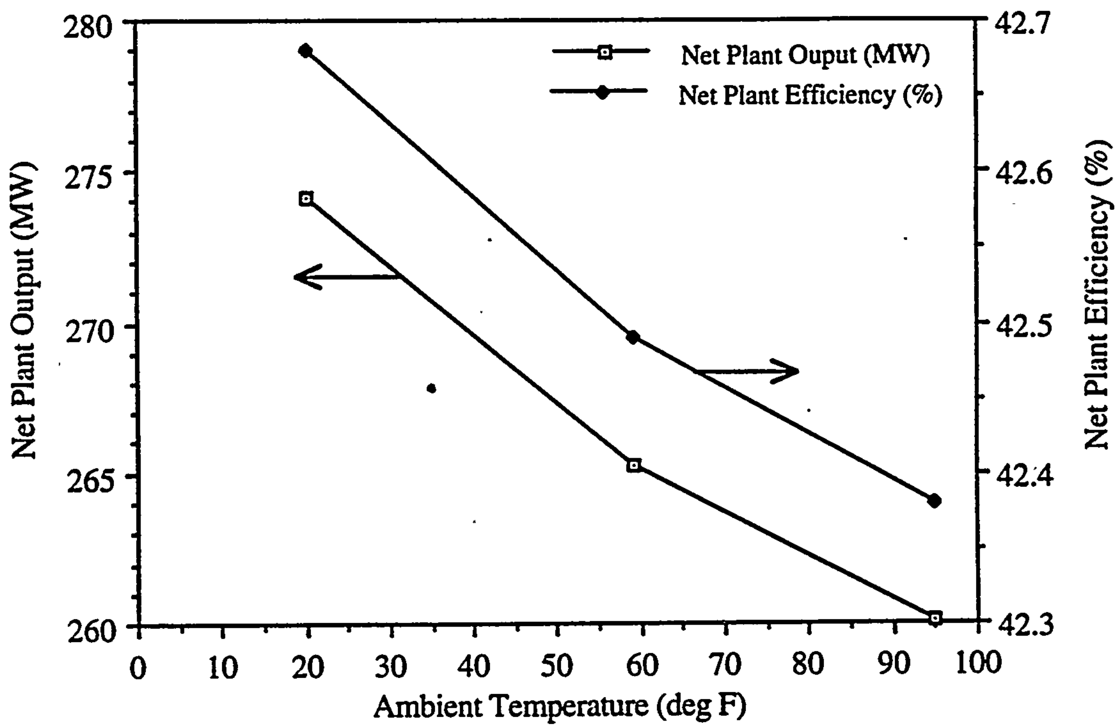


Figure 9.24. Net Plant Power Output and Net Plant Efficiency versus Ambient Temperature.

The effect of ambient temperature on the gas and steam turbine power output is shown in Figure 9.23. Since the gas turbine compressor is a constant volume machine, the mass flow of air at the compressor inlet decreases with an increase in the ambient temperature. Therefore, to maintain the mass flow of air at the turbine expander inlet at choked conditions, the volume of ambient air at the compressor inlet increases as the ambient temperature increases. With an increase in the ambient temperature, therefore, the work requirement of the compressor increases. As the ambient temperature increases from 20 °F to 95 °F, the net gas turbine power output decreases by 28 MW. At a higher ambient temperature, the temperature of the compressor discharge air is higher. Therefore, to achieve a specified turbine inlet temperature, the heat duty of the CerHx is reduced. The temperature of the flue gas entering the HRSG increases for higher ambient temperatures, which increases the thermal input to the steam cycle and the steam turbine power output. The steam turbine power output increases by 13.9 MW as ambient temperature increases for 20 °F to 95 °F.

Figure 9.24 shows the effect of ambient temperature on the net plant power output and the net plant efficiency. Since the decrease in the gas turbine power output is greater than the increase in the steam turbine power output with increasing ambient temperature, the net plant power output decreases by 14 MW. The net plant efficiency decreases by 0.3 percentage points.

9.3.4 Environmental Control

Sensitivity analyses associated with the environmental control system for the EFCC focused upon variation in stack gas temperature. A higher stack gas temperature increases the reheat requirement of the flue gas exiting the FGD unit.

The stack gas temperature was varied from 150 °F to 200 °F to study the effect of increasing the stack gas temperature on the plant performance parameters. In addition, a separate case study was considered for a "wet stack".

The gas and steam turbine power output as a function of stack gas temperature is shown in Figure 9.25. Since steam from the steam cycle is used for heating the flue gas exiting the FGD unit, the gas turbine operation is not affected. The gas turbine power output therefore remains constant at 138. Steam from the second stage low pressure turbine is extracted for heating the flue gas exiting the FGD to achieve the desired stack gas temperature. For a higher stack gas temperature requirement more steam is extracted, which imposes a higher penalty on the steam turbine output. As the stack gas temperature is increased from 150 °F to 200 °F, the penalty on the steam cycle due to reheat increases from 3.94 MW to 17.8 MW, and the steam turbine output is

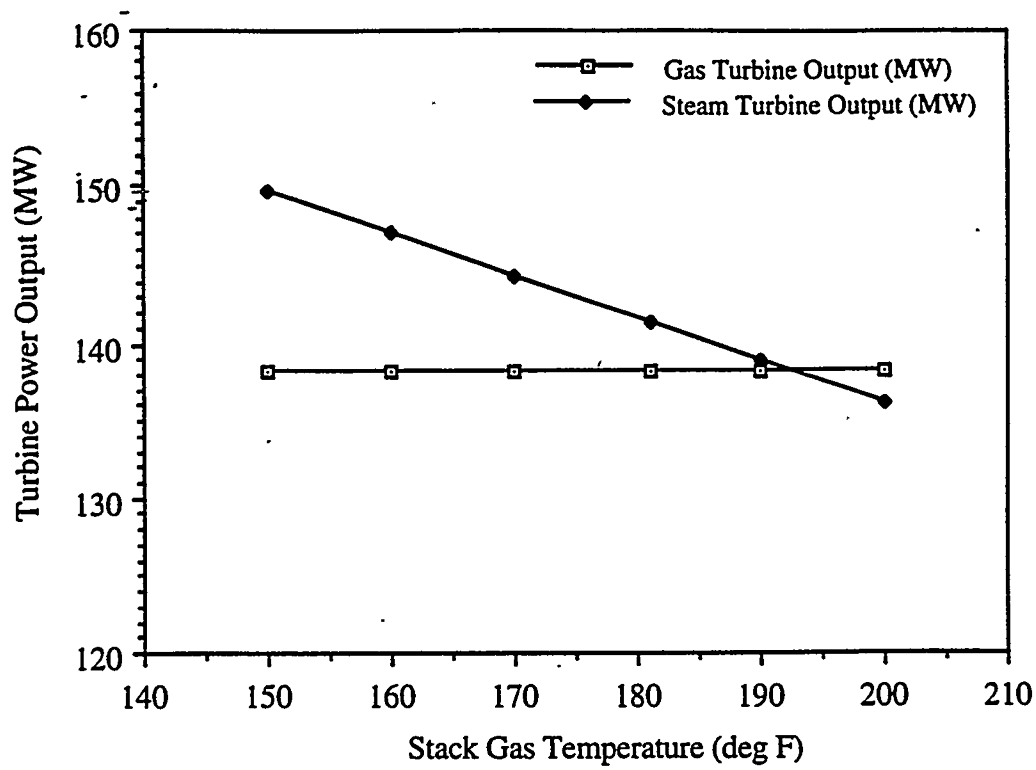


Figure 9.25. Gas and Steam Turbine Power Output versus Stack Gas Temperature.

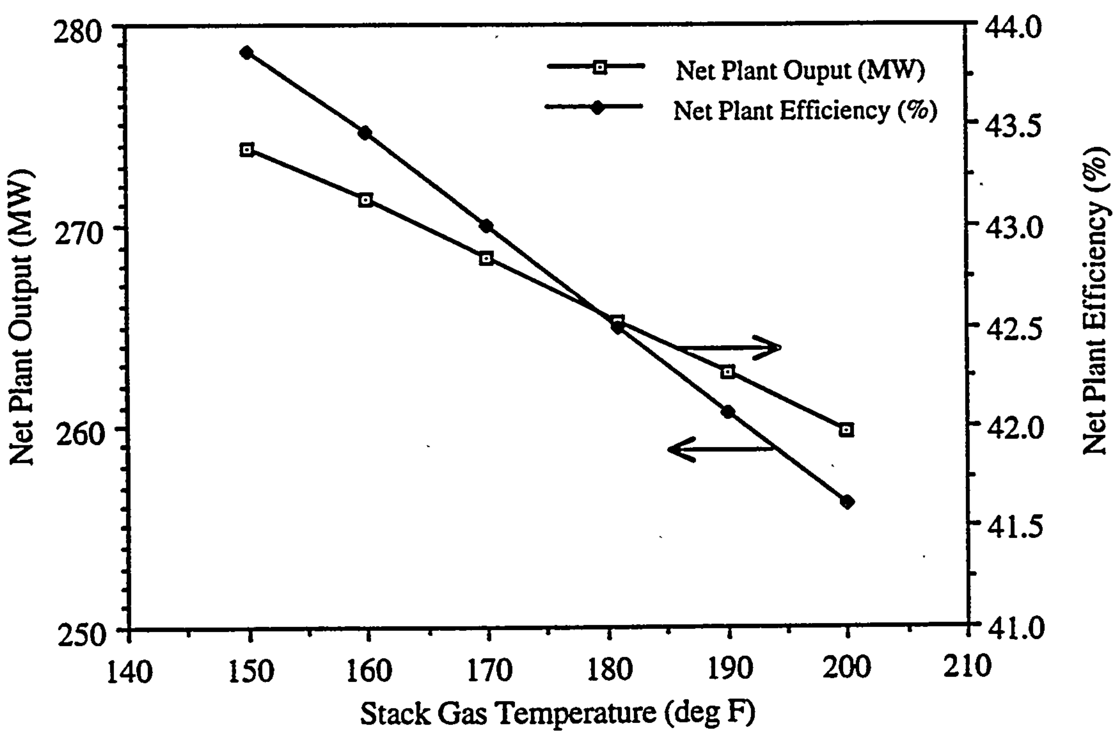


Figure 9.26. Net Plant Power Output and Net Plant Efficiency versus Stack Gas Temperature.

decrease by 14 MW. At higher stack gas temperatures, the volume of the flue gas entering the ID fan also increases, which increases the ID fan power requirement by 0.5 MW

Figure 9.26 shows the effect of increasing the stack gas temperature on the net plant power output and the net plant efficiency. Since the gas turbine power output remains constant and the steam turbine power output decreases with an increase in the stack gas temperature, the net plant power output decreases by 14.4 MW. The net plant efficiency decreases decrease by 2.3 percentage points over the range considered.

For a "wet stack", no FGD reheat would be employed. This yields a stack gas temperature of 136 ^oF. Under these conditions, the net plant output is 278.2 MW and the plant efficiency is 44.56. These improvements in plant performance illustrate the potential benefits of a wet stack. While the capital and operating costs due to reheat will be reduced, additional capital costs would be incurred for an improved stack liner.

9.4 Discussion

The sensitivity analysis of the modified EFCC model with respect to the parameters listed in Table 9.1 yield results which help in the identification of the most sensitive variables with respect to the plant performance parameters such as the net plant power output and the net plant efficiency.

The net plant power output and the net plant efficiency is seen to depend most strongly on the turbine inlet temperature and the pressure ratio. The net plant power output depends strongly also on the combustor heat loss, combustor heat directed to water wall, auxiliary air to coal flow ratio, air leakage in CerHx, CerHx heat loss, CerHx effectiveness, water injection rate, ambient temperature, and the stack gas temperature. The net plant power output depends moderately on the steam injection rate, and is insensitive to the carbon conversion.

The net plant efficiency depends strongly on the combustor heat loss, CerHx heat loss, water injection rate, and the stack gas temperature. It depends moderately on coal conversion, CerHx air leakage, CerHx effectiveness, and ambient temperature. The net plant efficiency is relatively insensitive to percentage of combustor heat directed to water wall, auxiliary air to coal flow ratio, and steam injection rate.

Design variables are those variables for which the values can be specified for a particular design of the EFCC system. Among the variables considered for sensitivity analysis, the design variables were the auxiliary air flow, CerHx effectiveness, turbine inlet temperature, gas turbine pressure ratio, steam injection rate, water injection rate, and the stack gas temperature. These variables can be varied to achieve an optimal system design.

Performance variables are those which are dependent on the design and operating conditions of the particular process equipment. There are inherent uncertainties associated with performance variables, and for unproven technologies the uncertainties tend to be the highest. Among the sensitive variables considered in this study, the uncertain performance variables are the combustor heat loss, combustor heat directed to water wall, combustor carbon conversion, air leakage in CerHx, CerHx radiative heat loss, and ambient temperature.

An optimal design of the EFCC system would have to take into account the effects of uncertainties associated with the performance variables. Thus, one may choose to optimize the system in the face of uncertainties, or to select system designs that would reduce the variance in predicted results.

10.0 CONCLUSIONS AND FUTURE WORK

This project features a review of the performance and cost of the Externally-Fired Combined Cycle, evaluation of a previously developed performance model of the EFCC, development of new performance and cost models, and extensive applications of the new models. The performance models are implemented using the ASPEN chemical process simulator. The newly developed models enable a large number of "what if" scenarios to be evaluated, which provides insights into the implications of key process uncertainties and into the potential benefits of process optimization. The models can be used to evaluate alternative EFCC designs, estimate the range of uncertainties in predictions of future EFCC performance and cost, and help make decisions regarding research, development, and demonstration priorities for the EFCC.

The details of a HI conceptual design of a 300 MW "greenfield" EFCC plant were outlined as the design basis for developing a performance model of the EFCC system. The performance model previously developed by METC, which is based on the HI conceptual EFCC design, was then documented. Several modifications were made to the original METC performance model to overcome a number of shortcomings which were identified as part of this work. These modifications enable the performance of the EFCC system to be more accurately represented. Key examples of these modifications include: (1) disaggregation of the auxiliary power requirements of the EFCC system; (2) additional detail regarding the gas turbine process area; (3) more extensive consideration of heat and pressure losses in major system components; (4) simulation of air leakage in the CerHx; (5) modeling of alternative designs based upon wet injection and combustor water wall heat recovery; (6) more extensive characterization of environmental discharges; and (7) more detailed characterization of FGD performance impacts.

A cost model of the EFCC system has been developed, based on the structure suggested by the EPRI Technical Assessment Guide (EPRI, 1986), to estimate the total capital cost and the operating and maintenance costs of the EFCC system. The analytical cost model was coded as a FORTRAN subroutine which is called by the ASPEN simulation program for the performance of the EFCC. The ASPEN performance model provides values for key system variables, such as flowrates, that are required as inputs to the cost model. The cost model developed here is complete, except for quantification of parameter values for the direct cost models for the coal combustor, slag screen, and ceramic heat exchanger. Cost data for these components, which are unique to the EFCC, were not available during the course of this work. Completion of these cost model components is a need for future work.

The efficiency and net plant output of the EFCC system depends on the values of key design parameters and performance variables. To evaluate the response of the newly developed EFCC system model to changes in these parameters, and to identify key uncertainty and design issues for the EFCC, the new performance model of the EFCC system was applied in a series of case studies. These case studies involved over 50 sensitivity analyses in which design and performance parameters were varied over likely or plausible ranges. The results of these case studies enabled evaluation of the EFCC model's behavior. While in a strict sense it is not possible to validate this model, because the system being modeled has not yet been commercially demonstrated, the sensitivity analyses provide a degree of confidence regarding the reasonableness of the model in predicting trends. Over the course of this study, expectations for EFCC system performance and model behavior were developed, and the model was then evaluated with respect to these expectations. The process of running the model for multiple cases provides insights into system performance that take into account interactions among multiple system components. For example, the effects of combustor heat losses, water wall heat recovery, ceramic heat exchanger effectiveness, steam injection rates, and other design and performance parameters lead to a variety of impacts on plant heat rate and output that can be evaluated only by a sufficiently detailed and comprehensive model.

The EFCC technology is currently undergoing proof-of-concept testing. As new test results become available, it will be necessary to update the design basis and revise assumptions regarding model input parameters values. Furthermore, related technologies, such as for gas turbines, are undergoing significant changes. For example, the impacts of improved turbine inlet temperatures and overall turbine designs should be evaluated in future work. The model developed here should be iteratively refined as new data become available.

Since the EFCC technology is in the early stages of development, there are inherent uncertainties in the performance and cost parameter estimates. Incorporating uncertainties is critical to the design and evaluation of the EFCC system. A systematic approach is required for characterizing uncertainties in the EFCC technology. Therefore, there is a need to develop a probabilistic model to account for the uncertainties in the performance and cost parameters. The results from the probabilistic model simulation would include possible ranges of values for the performance, environmental emissions, and cost parameters of the EFCC system. These results could then be used to identify key uncertainties; optimize the flowsheet configuration and parameter values; identify process areas for further research; and probabilistically compare competing advanced coal-based power generation technologies, such as the IGCC and the PFBC systems, in order to identify the risks and potential payoffs of the EFCC technology relative to other technologies. The performance and cost models developed here, which have been evaluated via extensive sensitivity analysis, will be applied in future work to probabilistic analysis of process uncertainties and to optimization under uncertainty.

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APPENDIX A: DOCUMENTATION OF METC MODEL

This appendix provides a concise documentation of an ASPEN performance model of the EFCC developed by DOE/METC, prior to modifications as part of this project. The details of the various process sections of the METC model are shown in Figures A.1 -A.7, and Tables A 1-A.7. The figures show the process flow diagram of the process section and the tables describe the function of each unit operation block associated with the flow diagram. The details of the following process sections are shown.

A.1 Combustor Section

The process flow diagram of the combustor section is shown in Figure A.1 and the unit operation block descriptions are given in Table A.1.

A.2 Gas Turbine Section

The process flow diagram of the gas turbine section is shown in Figure A.2 and the unit operation block descriptions are given in Table A.2.

A.3 Heat Recovery Steam Generator (flue gas-side) Section

The process flow diagram of the heat recovery steam generator (flue gas-side) section is shown in Figure A.3 and the unit operation block descriptions are given in Table A.3.

A.4 Heat Recovery Steam Generator (steam-side) Section

The process flow diagram of the heat recovery steam generator (steam-side) section is shown in Figure A.4 and the unit operation block descriptions are given in Table A.4.

A.5 Steam Turbine Section

The process flow diagram of the steam turbine section is shown in Figure A.5 and the unit operation block descriptions are given in Table A.5.

A.6 Flue Gas Desulfurization Section

The process flow diagram of the flue gas desulfurization section is shown in Figure A.6 and the unit operation block descriptions are given in Table A.6.

A.7 Plant Energy Balance Section

The process flow diagram of the plant energy balance section is shown in Figure A.7 and the unit operation block descriptions are given in Table A.7.

Table A.1. Combustor Section Unit Operation Block Description

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
1	AIRBLOW (COMPR)	TYPE = 3 Pressure = 16.4 psia Isentropic Efficiency = 0.72	Air blower for the auxiliary air required for pneumatic transport of coal to combustor. It increases the pressure of the air from 14.7 psia to 16.4 psia.
2	TO-COMB (CLCHNG)		Changes turbine exhaust air stream class from CONVENTIONAL to MIXNC.
3	MIXING (MIXER)	Pressure = 15.13 psia	Sets the pressure for all combustor feed streams to 15.13 psia. Mixes auxiliary air, combustion air, and coal.
4	COMBUSTR (RSTOIC)	Pressure = -0.11 psia Temperature = 2750 °F NPK = 1	The stoichiometric combustor simulates stoichiometric combustion reaction of coal. The combustion temperature is initially set at 2750 °F, and is later reset by the FORTRAN block EFFECT based on the temperature of the flue gas required to produce the specified turbine inlet temperature (TIT).
5	SLAGGER (SEP 2)	SUBS-FRAC MIXED FLUEGAS 1.0 / NC FLUEGAS 0.0 / NC ASH 1.0	Used to simulate the separation of ash from the flue gas. The fracton of gas in the combustor exit stream is specified to be 1.0 in the flue gas stream and the fractions of ash in the combustor exit stream is specified to be 1.0 in the ash stream
6	HPBOILH (CLCHNG)		Changes the flue gas stream class from MIXNC to CONVENTIONAL.
7	CERHXH (HEATER)	Pressure = -0.76 psia NPK = 1 KPH = 1	Hot side of the ceramic heat exchanger, which transfers heat to the incoming compressed cool air stream. The pressure drop is specified to be 0.76 psi. The heat duty is estimated from the cold side of the ceramic heat exchanger block (CERHXC) in the gas turbine flowsheet.

(continued)

Table A.1. (continued)

[a] Numbers for each unit operation block correspond to those in Figure A.1.

[b] 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 B.1 in Appendix B.

[c] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

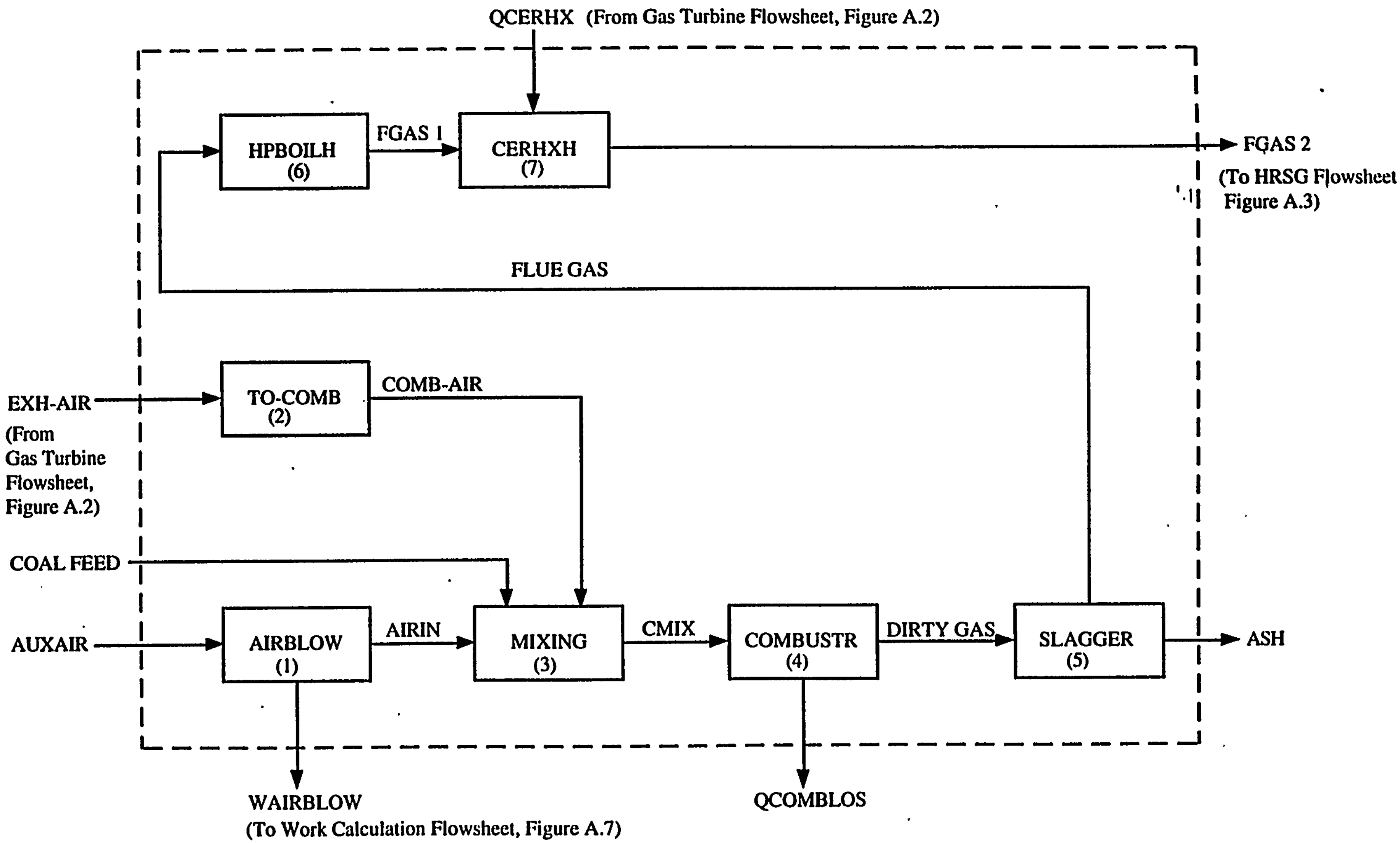

Figure A.1. Combustor Flowsheet

Table A.2. Gas Turbine Section Unit Operation Block Description

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
1	FILTER (MIXER)	Pressure = -0.5 psi NPK = 1 KPH = 1	Simulates the pressure drop across the air filter which is specified to be 0.5 psi.
2	GT-COMP 1 (COMPR)	TYPE = 3 Pressure = 52.6 psia Isentropic Efficiency = 0.92 GFLAG = 1	Simulates an isentropic compressor. The outlet pressure is specified to be 52.5 psia.
3	GT-COMP 2 (COMPR)	TYPE = 3 Pressure = 81.3 psia Isentropic Efficiency = 0.92 GFLAG = 1	The outlet pressure is specified to be 81.3 psia.
4	GT-COMP-3 (COMPR)	TYPE = 3 Pressure = 198.5 Isentropic Efficiency = 0.92 GFLAG = 1	The outlet pressure is specified to be 198.5 psia.
5	GT-SPLT 1 (FSPLIT)	FRAC HPAIR2 = 0.196	Splits the compressor discharge air for turbine blade cooling. Air bleed for 2300 °F turbine blade cooling is assumed to be 19.6% of the compressor discharge.

(continued)

Table A.2. (continued)

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
6	GT-SPLT 2 (FSPLIT)	NCHCLG = 0.4222	Splits the cooling flow into "non-chargeable" and "chargeable" cooling flows. Non-chargeable cooling flow is the fraction of cooling flow which contributes to power generation in the turbine. Non-chargeable cooling flow is assumed to be 42.44 percent of the total air cooling flow. Chargeable cooling flow is the fraction of cooling flow which does not contribute to power generation in the turbine, and bypasses the turbine inlet air flow to model turbine power output
7	GT-EVAP (MIXER)		This block is used to add steam to the turbine air stream before it enters the ceramic heat exchanger. The steam flow rate can be varied by the user to represent different levels of steam injection for power augmentation.
8	CERHXC (HEATER)	Pressure = -3.5 psia Temperature = 2435 °F	Cold side of the ceramic heat exchanger, where compressed air is heated for expansion in gas turbine. The pressure drop is specified to be 3.5 psi. The outlet air temperature is specified to be 2435 °F, which is later reset by the DESIGN-SPEC CONTIT to achieve the desired T_i.
9	CERHXHL (MULT)	Factor = 1.02	Accounts for the assumed two percent combustor and ceramic heat exchanger total heat loss based on the ceramic heat exchanger heat duty.
10	GT-MIX 1 (MIXER)	Pressure = -0 psia	Mixes the turbine inlet air with the cooling air. The pressure drop across this block is assumed to be zero.

(continued)

Table A.2. (continued)

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
11	GT-TURB 1 (COMPR)	TYPE = 3 Pressure = 83.0 psia Isentropic Efficiency = 0.913 GFLAG = 1	Simulates an isentropic turbine expander. The outlet pressure is specified to be 83.0 psia.
12	GT-TURB 2 (COMPR)	TYPE = 3 Pressure = 54.0 psia Isentropic Efficiency = 0.913 GFLAG = 1	The outlet pressure is specified to be 54.0 psia.
13	GT-TURB 3 (COMPR)	TYPE = 3 Pressure = 15.2 psia Isentropic Efficiency = 0.913 GFLAG = 1	The outlet pressure is specified to be 15.2 psia.
14	GT-MIX 2 (MIXER)	Pressure = 15.13 psia	Mixes the turbine exhaust and the chargeable cooling air flow. The combined stream forms the oxidation air stream to the coal combustor. The outlet pressure is specified to be 15.13 psia.

[a] Numbers for each unit operation block correspond to those in Figure A.2.

[b] 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 B.1 in Appendix B.

[c] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

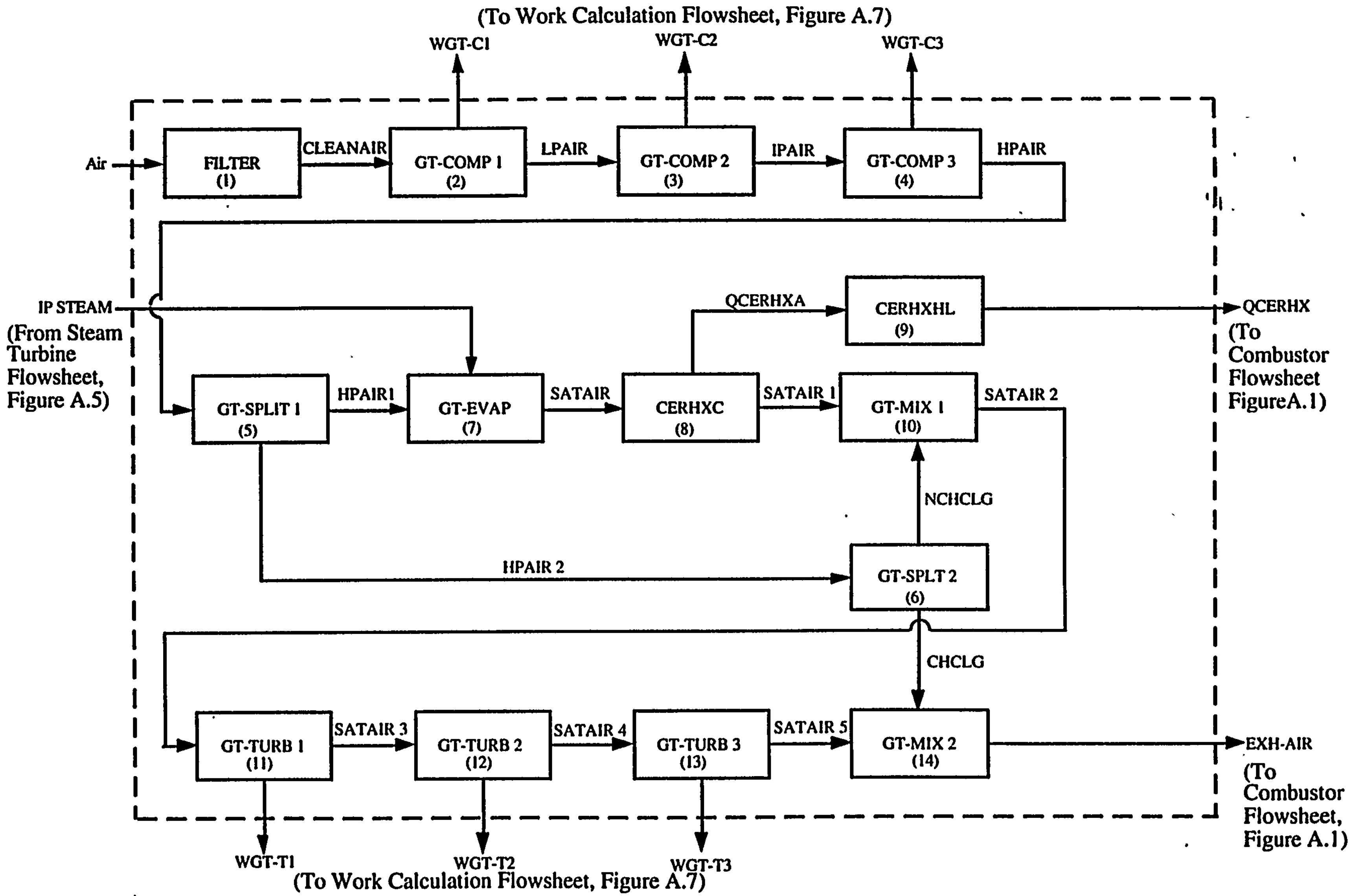

Figure A.2. Gas Turbine Flowsheet

Table A.3. **Heat Recovery Steam Generator (Flue Gas-Side) Section Unit Operation Block Description**[a]

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
1	REHEATH (HEATER)	Pressure = -0.028 psi NPK = 1 KPH = 1	The pressure drop across the hot side of the reheater is specified to be 0.028 psi.
2	SUPHTXH (HEATER)	Pressure = -0.028 psi NPK = 1 KPH = 1	The pressure drop across the hot side of the superheater is specified to be 0.028 psi.
3	HPEVAPH (HEATER)	Pressure = -0.028 psi NPK = 1 KPH = 1	The pressure drop across the hot side of the high pressure evaporator (boiler) is specified to be 0.028 psi.
4	HPECONH (HEATER)	Pressure = -0.028 psi NPK = 1 KPH = 1	The pressure drop across the hot side of the high pressure evaporator (economizer) is specified to be 0.028 psi.
5	LPEVAPH (HEATER)	Pressure = 14.12 psi NPK = 1 KPH = 1	The pressure at the outlet of the hot side of the low pressure evaporator (boiler) is specified to be 14.12 psia.
6	REHTL (MULT)	Factor = 1.01	The heat loss in the reheater is assumed to be one percent of the reheater heat duty..
7	SUPHL (MULT)	Factor = 1.01	The heat loss in the superheater is assumed to be one percent of the superheater heat duty.
8	BOILL (MULT)	Factor = 1.01	The heat loss in the high pressure boiler is assumed to be one percent of the high pressure boiler heat duty.
9	ECONL (MULT)	Factor = 1.01	The heat loss in the economizer is assumed to be one percent of the economizer heat duty.
10	LPBL (MULT)	Factor = 1.01	The heat loss in the low pressure boiler is assumed to be one percent of the low pressure boiler heat duty.

(continued)

Table A.3. (continued)

NO[b]	BLOCK ID[c] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[d]	DESCRIPTION
11	TO-WFGD (CLCHNG)		Changes the stream class of outlet flue gas stream of low pressure boiler from CONVENTIONAL to MIXCI.

[a]The outlet HRSG temperatures are calculated based on the steam cycle design. The HRSG flue side pressure loss through the LP boiler is 7.4" of water.

[b] Numbers for each unit operation block correspond to those in Figure A.3.

[c] 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 B.1 in Appendix B.

[d] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

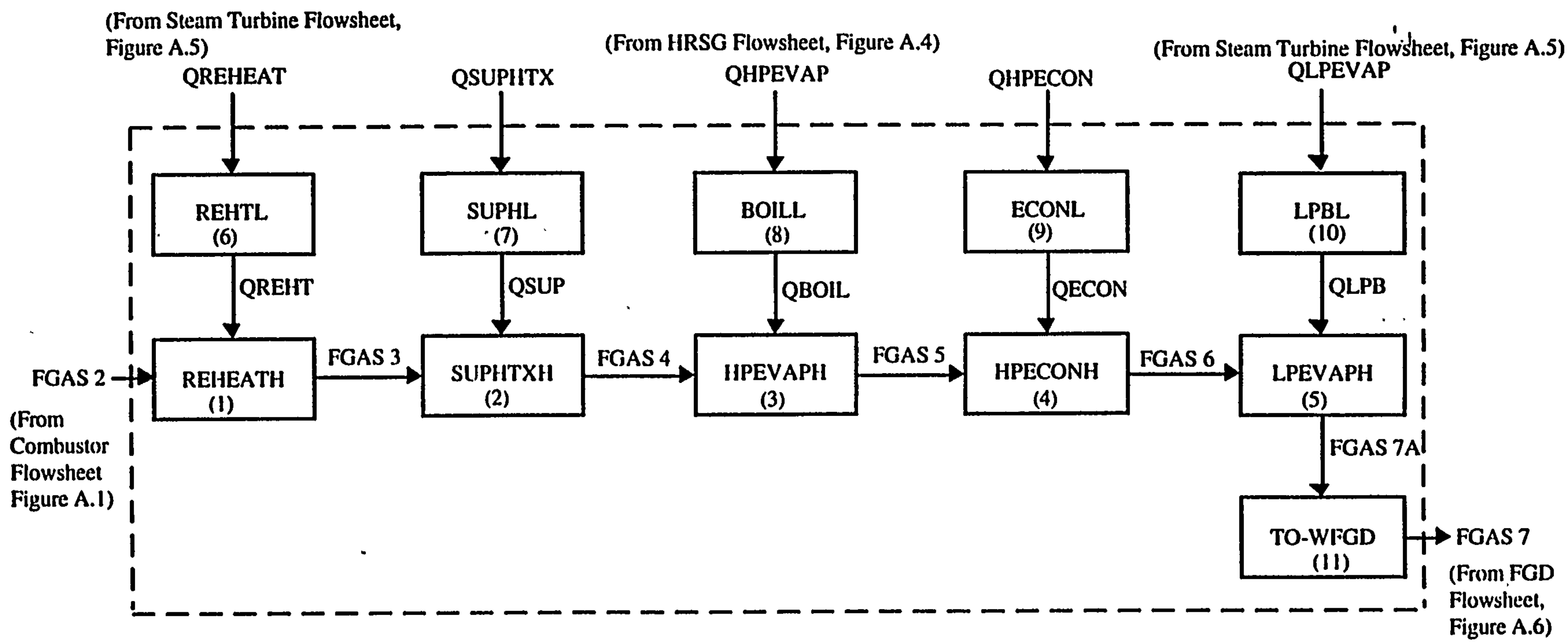

Figure A.3. HRSG (flue gas side) Flowsheet

Table A.4. Heat Recovery Steam Generator (Steam-Side) Section Unit Operation Block Description

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
1	HPECONC (HEATER)	Pressure = -85.0 psia Temperature = 563.2 °F	The pressure drop and the outlet temperature in the cold (steam) side of the high pressure economizer are specified to be 85 psi and 563.2 °F respectively.
2	ST-SPLT4 (FSPLIT)	Fraction of ST 17 = 0.1 E-10	Splits the steam from the high pressure economizer. A negligible fraction is directed to combustor water walls in this design.
3	HPBOIL (CLCHNG)		Changes the class of the inlet stream.
4	ST-MIX1 (MIXER)	Pressure = 0.0 psia	Combines the stream from the combustor water wall and the high pressure economizer.
5	HPEVAPC (FLASH 2)	Pressure = -100.0 psia V = 0.98	Steam side of the high pressure evaporator. Flashes the inlet liquid stream. 98% of the outlet stream is specified to be vapor which goes to the superheater, and the rest is blowdown. The pressure drop is specified to be 100 psi.
6	SUPHTXC (HEATER)	Pressure = -30.0 psia Temperature 1050 °F NPK = 1 KPH = 1	Steam side of the superheater. The outlet temperature and the pressure drop are specified to be 1050 °F and 30 psi respectively.

[a] Numbers for each unit operation block correspond to those in Figure A.4.

[b] 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 B.1 in Appendix B.

[c] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

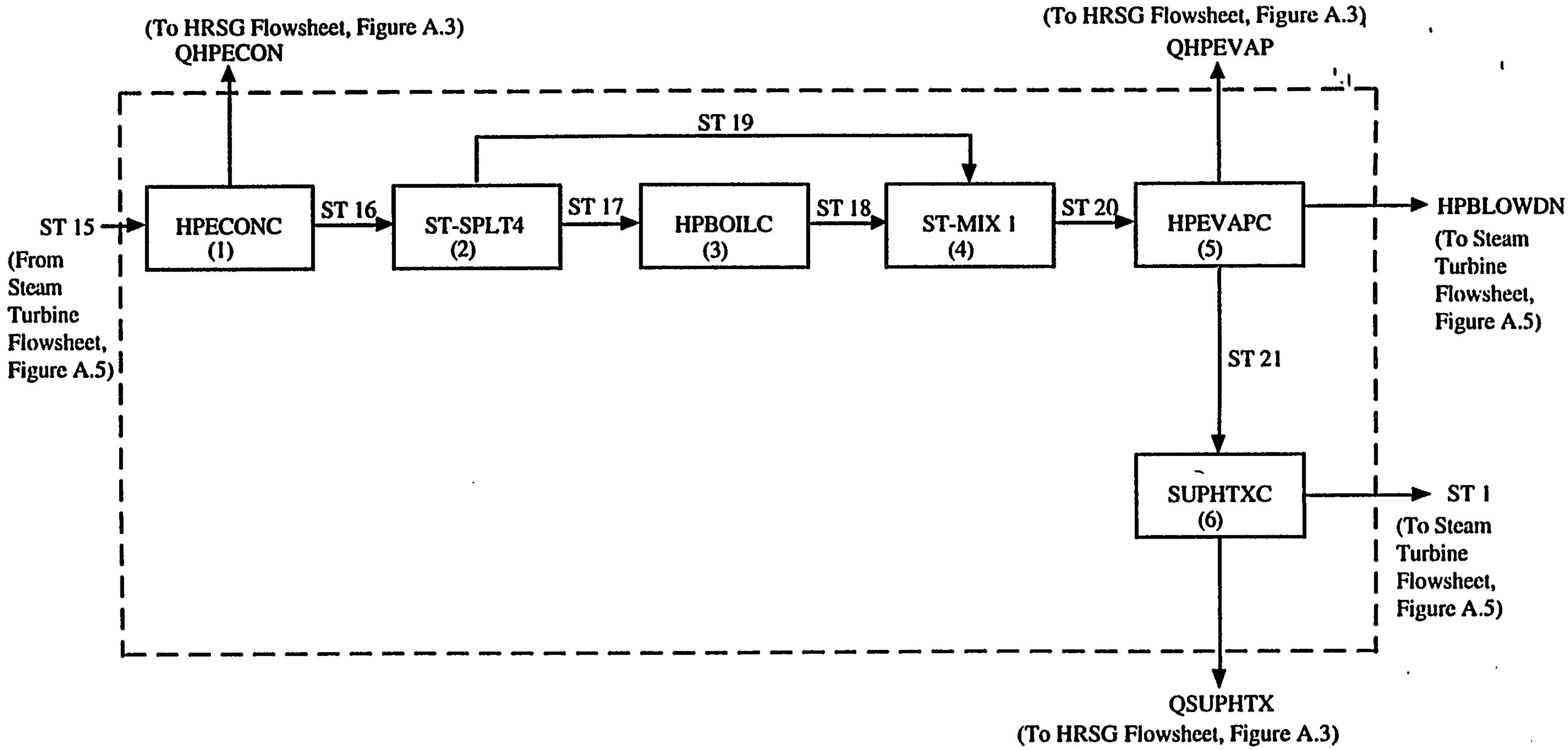

Figure A.4. HRSG (steam side) Flowsheet

Table A.5. Steam Turbine Section Unit Operation Block Description

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
1	MAT-LOSS (MIXER)	Pressure = 14.7 psia	Combines the blowdown from the high and low pressure boiler and the vapor stream from the deaerator. The outlet pressure is specified to be 14.7 psia.
2	MAKE-UP (HEATER)	Pressure = -0 psi Temperature =59 ^{o}F	Cools the vapor and liquid mixed stream from the MAT-LOSS block to liquid water, which forms the makeup water stream to the steam cycle. The temperature of the outlet stream is specified to be 59 ^{o}F and the pressure drop is assumed to be zero.
3	HPTURB (COMPR)	Pressure = 720 psia TYPE = 3 GFLAG = 1 ES = 0.92	The high pressure steam turbine isentropic efficiency and the outlet pressure is specified to be 0.92 and 720 psia respectively.
4	REHEATC (HEATER)	Pressure = -30 psi Temperature = 1050 ^{o}F NPK =1 KPH = 1	Steam side of the reheater. Steam is reheated to 1050 ^{o}F. The pressure drop across the steam side of the reheater is specified to be 30 psi.
5	IPTURB (COMPR)	Pressure = 200 psia TYPE = 3 GFLAG = 1 ES = 0.92	Intermediate pressure steam turbine. The isentropic efficiency and the outlet pressure are specified to be 0.92 and 200 psia respectively.
6	ST-SPLT1 (FSPLIT)	MASSFLOW IPSTEAM 5.0D-9	Splits the outlet steam from the IP turbine into two streams, one of which is mixed with the compressed air stream in the gas turbine section. Mass flow of IP steam can be varied by the user. The default value is negligibly small, to represent the case of no steam injection.

(continued)

Table A.5. (continued)

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
7	LPTURB1 (COMPR)	Pressure = 17.5 psia TYPE = 3 ES = 0.89	First stage low pressure steam turbine. The isentropic efficiency and the outlet pressure is specified to be 0.89 and 17.5 psia respectively.
8	ST-SPLT 2 (FSPLIT)	FRAC LPEXTRAC stream = 0.08	8% of the outlet steam from the first LP turbine stage is sent to the deaerator.
9	LPTURB 2 (COMPR)	Pressure = 0.981 psia TYPE = 3 ES = 0.89	Second stage low pressure steam turbine. The isentropic efficiency and the outlet pressure is specified to be 0.89 and 0.981 psia respectively.
10	CONDWELL (MIXER)	Pressure = -0 psi	Combines the makeup water and exhaust steam from the second stage LP turbine. The pressure drop is assumed to be zero.
11	CONDEN (HEATER)	Pressure = 0.982 psia Temperature = 90 °F	Condenses the steam and water mixed stream from the Condensor Well to water. The outlet pressure and temperature are specified at .0.982 psia and 90 °F respectively. It is designed for approximately 10 °F subcooling before being fed to the pump.
12	CONDPUMP (PUMP)	Pressure = 17.5 psia TYPE = 1	Pumps the condensate to the deaerator. The pump is specified to be centrifugal type with an outlet pressure of 17.5 psia.
13	DEAERATR (FLASH 2)	Pressure = 17.5 psia V = 0.01	Flashes the liquid stream from the condenser pump and the vapor stream from the first stage LP turbine and low pressure boiler. The vapor fraction and the pressure of the outlet stream are specified to be 1% and 17.5 psia respectively.

(continued)

Table A.5. (continued)

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
14	LPPUMP (PUMP)	Pressure = 2000 psia TYPE = 1	Represents the boiler feedwater pump which pumps the liquid stream from the deaerator. The pump is specified to be centrifugal type with an outlet pressure of 2000 psia.
15	ST-SPLT3 (FSPLIT)	FRACTION ST 13 = 0.0635	Splits the liquid stream from the boiler feedwater pump into two streams. 6.35% of the stream is directed to the economizer, and the rest to the low pressure boiler.
16	VALVE (MIXER)	Pressure = 17.5 psia Temperature = 224.8 ^oF	Decreases the pressure of the stream going to the low pressure boiler to 17.5 psia. The outlet stream temperature is specified to be 224.8 ^oF.
17	LPEVAPC (FLASH 2)	Pressure = 17.5 psia V = 0.99	Flashes the inlet stream. 1% of the stream goes to blowdown and the rest goes to the deaerator at an outlet pressure of 17.5 psia..

[a] Numbers for each unit operation block correspond to those in Figure A.5.

[b] 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 B.1 in Appendix B.

[c] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

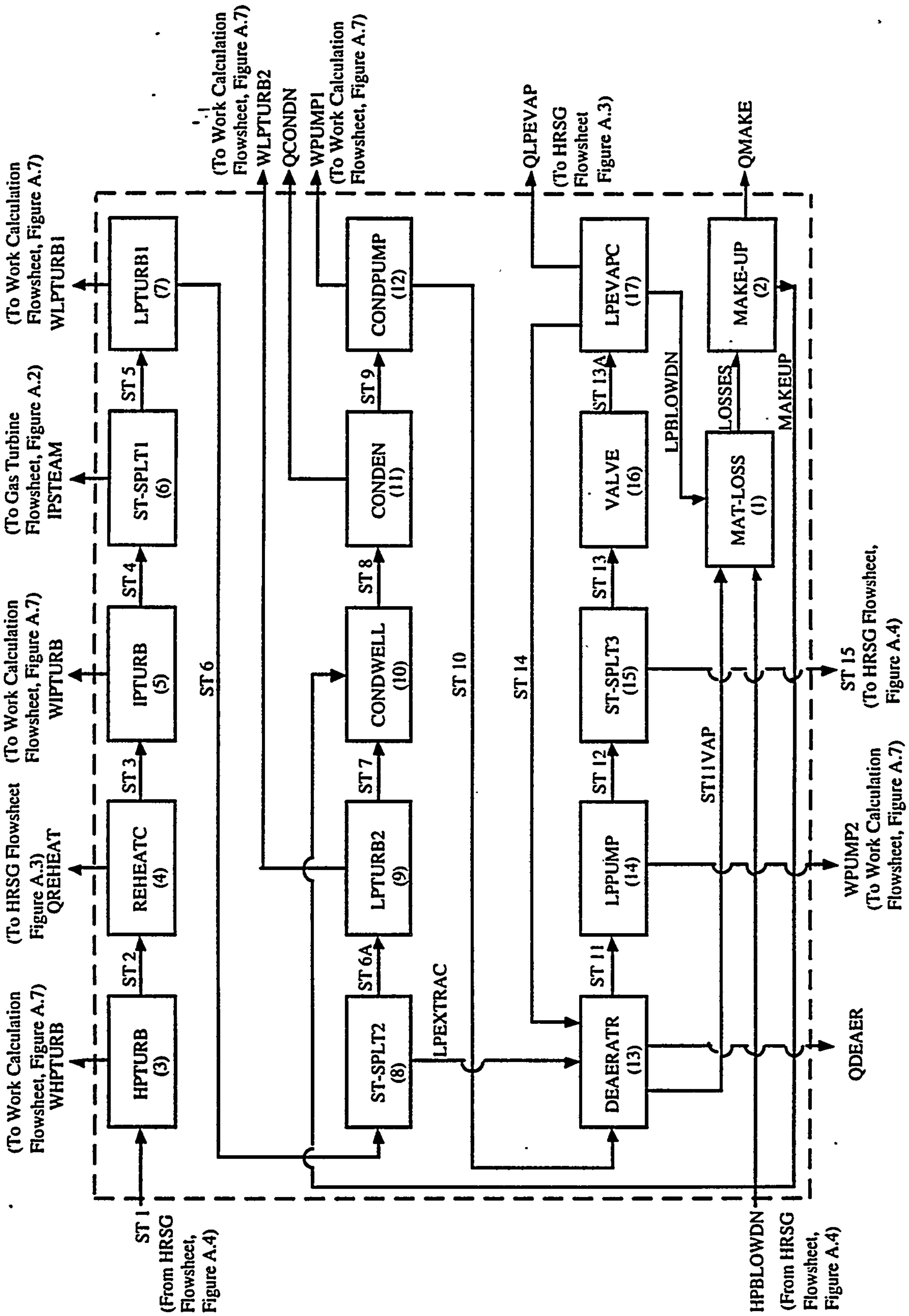

Figure A.5. Steam Turbine Flowsheet

Table A.6. Flue Gas Desulfurization Section Unit Operation Block Description

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
1	FGDMIX (MIXER)	Pressure = -0.2 psi	Mixes the limestone slurry feed with the HRSG exhaust flue gas. The pressure drop is specified to be 0.2 psi.
2	FGDRXR (RSTOIC)	Pressure = -0.02 psi DUTY = 0	Represents a stoichiometric reactor for the flue gas desulfurization reaction. The pressure drop and the heat duty is specified to be 0.02 psi and 0 respectively. The conversion of SO_2 in the flue gas is specified to be 95%.
3	FGDSEP (SSPLIT)	FRAC CISOLID SLUDGE = 1 MIXED FGAS 9 = 1	Splits the outlet stream from the FGD reactor into the sludge and flue gas streams.
4	FGDADJ (FLASH 2)	Temperature = 127 °F Pressure = -0 psi	Separates liquid and gaseous components of the flue gas stream from the FGD separator. The temperature of the outlet stream and the pressure drop are specified to be 127 °F and 0 respectively.
5	GASBLOW (COMPR)	Pressure = 14.7 psia TYPE = 3 Isentropic Efficiency = 0.85	Represents an Induced Draft (ID) fan, which is modeled as an isentropic compressor with a specified efficiency of 85%. The pressure of the inlet flue gas stream is raised to 14.7 psia.

[a] Numbers for each unit operation block correspond to those in Figure A.6.

[b] 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 B.1 in Appendix B.

[c] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

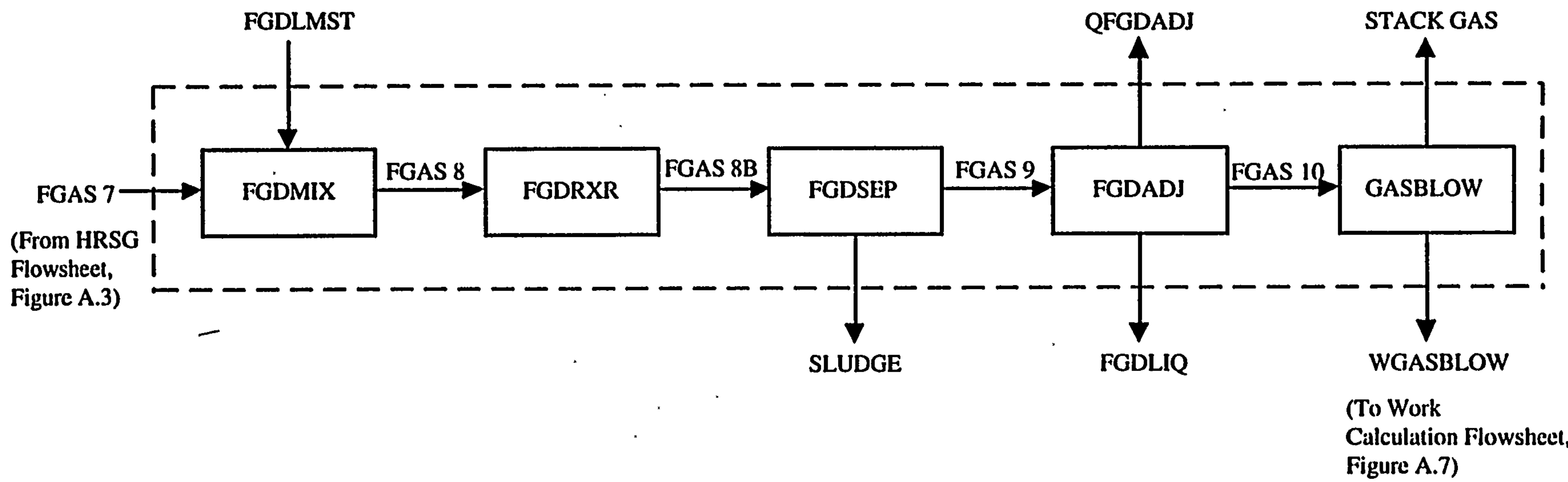

Figure A.6. Flue Gas Desulfurization Flowsheet

Table A.7. Heat and Work Section Unit Operation Block Description

NO[a]	BLOCK ID[b] (ASPEN BLOCK NAME)	BLOCK PARAMETERS[c]	DESCRIPTION
1	WGTSUM (MIXER)		Calculates the gross power output of the gas turbine section.
2	WGTSUMT (FSPLIT)	FRAC WGT = 0.986	Calculates the net power output of the gas turbine section based on 98.6% electric generator efficiency.
3	WSTSUM (MIXER)		Calculates the gross power output of the steam turbine section.
4	WSTSUMT (FSPLIT)	FRACT WST = 0.986	Calculates the net power output of the steam turbine section based on 98.6% electric generator efficiency
5	WMISCSUM (MIXER)		Calculates the total miscellaneous work, including FGD work.
6	WORK-TOT (MIXER)		Calculates the total work of the EFCC plant.
7	WORK-DUP (DUPL)		Duplicates the total work for estimations of auxiliary power and calculates the net power output of the EFCC plant by subtracting the auxiliary power from the total power.
8	WORK-AUX (MULT)	FACTOR = 0.03	Calculates the auxiliary power consumption of the EFCC plant. The auxiliary power is specified to be 3% of the total power output.

[a] Numbers for each unit operation block correspond to those in Figure A.7.

[b] 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 B.1 in Appendix B.

[c] For a glossary of ASPEN block parameters, please see Table B.2 in Appendix B.

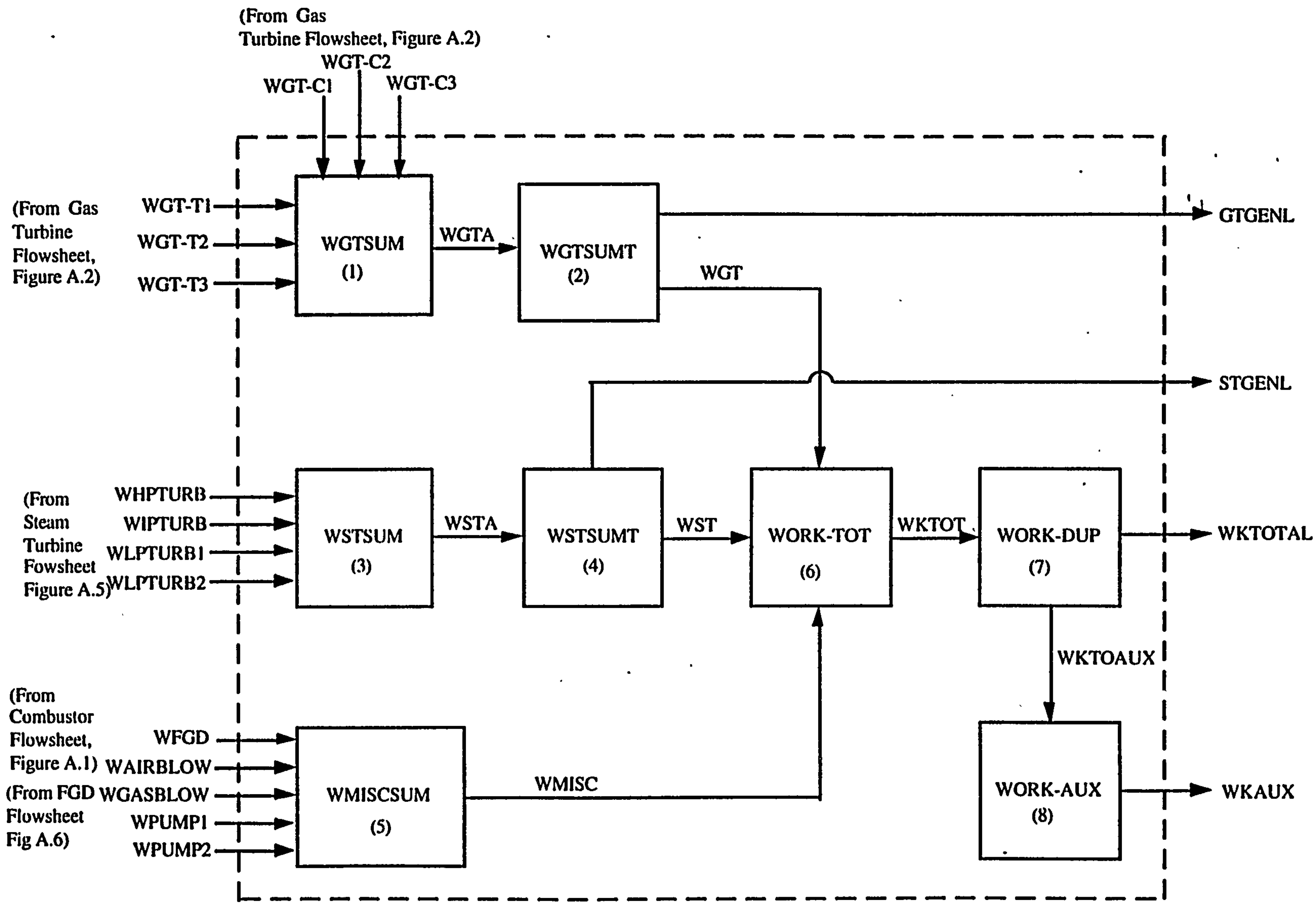

Figure A.7. Work Calculation Flowsheet

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

This appendix provides a summary of the ASPEN unit operation blocks and the associated block paramiters. 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 B.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 one 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 isentropic turbine/compressor.
DUPL	This block is used to copy 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 same 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 two or more outlet 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 inlet and 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.

(continued)

Table 1. (continued)

ASPEN MODEL NAME	DESCRIPTION
MIXER	The mixer block simulates the mixing of any number of 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 of the class "HEAT". The user can specify the outlet pressure or the pressure drop, the number of phases in the conventional substream, and the key phase.
MULT	This block is used to multiply an inlet stream by a given factor to produce an outlet stream. Therefore, the heat and material balances are not maintained in this model. The outlet stream has the same composition and intensive properties as the inlet streams. However, the extensive properties are a multiple of those of the inlet stream.
PUMP	The pump 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 the input work. This block can be used to model a cetrifugal pump, a slurry pump, or a positive displacement pump.
RSTOIC	The 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 is used to simulate 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.

(continued)

Table 1. (continued)

ASPEN MODEL NAME	DESCRIPTION
SEP	The separator block separates an inlet stream into two or more outlet streams according to the splits specified for each component. Two of the three properties, temperature, pressure, and vapor fraction, may be specified for each component.
SSPLIT	The substream splitter block splits an inlet stream into two or more outlet streams according to the split specifications given for each stream. Each outlet substream has the same composition and intensive properties as the corresponding inlet substream. However, the extensive properties are a fraction of those of the inlet.

Table 2. ASPEN Block Parameters Description

ASPEN Block Parameter	DESCRIPTION
DUTY	It is the specified heat duty for a block.
ES	It is the isentropic efficiency of the unit operation block
Factor	It is the specified multiplication factor which must be greater than zero.
FRAC	It refers to the fraction of an inlet stream which
GFLAG	It is the gas phase flag. A 0 flag refers to vapor and liquid in the stream, and 1 refers to only gas phase.
Isentropic Efficiency	It refers to the isentropic efficiency of a pump or compressor.
KPH	It is the single phase code. Options 1, 2, and 3 stand for vapor, liquid, and solid phases respectively.
MASSFLOW	It refers to the mass flow of an outlet stream.
NPK	It is the phase equilibrium code. Options 1, 2, and 3 refer to single phase, V-L, and V-L_1-L_2 equilibria respectively.
Pressure	It refers to the pressure of the outlet stream when positive and the pressure drop when 0 or negative.
SUBS-FRAC	It refers to the fraction of a substream in an outlet stream.
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 type 3 refers to an isentropic turbine/compressor.
V	It refers to the vapor fraction in the outlet stream.

APPENDIX C: ESTIMATION OF NUMBER OF OPERATORS PER SHIFT AND MAINTENANCE COST FACTORS

This appendix provides a summary of the calculations for the number of operators per shift for the EFCC process areas listed in Table 7.2 and the calculations for the maintenance cost factors for the EFCC process areas listed in Table 7.3.

C.1 Number of Operators per Shift

The number of operators for the coal handling section of a conventional coal fired power plant have been reported to be four, and for the same plant the number of operators for the coal combustor and the FGD unit are four and two respectively (EPRI, 1988b). Since these units in the EFCC plant are similar to the ones used in a conventional coal fired power plant, the number of operators for these process units of the EFCC plant can be considered to be the same. The number of operators for the boiler feed water system, the combined cycle system, and the general facilities of an IGCC plant have been reported to be one, four times the number of gas turbine plus one, and three respectively (Frey and Rubin, 1990). Since these process units for the EFCC plant are similar to the ones used in an IGCC system, the same number of operators have been considered for the boiler feed water system, the combined cycle, and the general facilities. Data were not available for the number of operators required for the slag screen, CerHx, and the ID fan. Therefore, for the base case EFCC plant the number of operators for each of these process units were assumed to be one. Based on the general consumption figures for estimating the operating and maintenance cost of a fabric filter, the operating labor requirement has been reported to be two to four hours per shift (Vatavuk, 1990). Since there are eight hours in a shift with two hours overlap, the minimum number of operating labor required for a fabric filter would be one.

C.2 Maintenance Cost Factors

The maintenance cost factors for the coal handling section, boiler feed water system, and the combined cycle system of an IGCC plant have been reported to be three, 1.5, and 1.5 respectively. Since these process sections are similar to the ones used in an EFCC system, the maintenance cost factors for these process areas have been assumed to be the same for an EFCC plant. Data for maintenance cost factors for the slag screen, ceramic heat exchanger, ID fan, and general facilities of an EFCC system were not available. Therefore, the maintenance cost factors for these process areas were given a most likely range, as shown in Table 7.3. The maintenance cost factors for the FGD used in a conventional coal fired power plant has been reported to be five (EPRI, 1988a). Since the FGD unit of the EFC system is similar to the one used in a conventional coal fired power plant, the maintenance cost factor for the FGD unit of the EFCC system has been assumed to be five.

APPENDIX D: COST MODEL AND SAMPLE OUTPUT

The cost model of the Externally Fired Combined Cycle is implemented in Fortran as a subroutine. This subroutine estimates direct capital costs based on the equations developed in Chapter 5 of the report, and estimates the total capital cost and annual costs based on the approach documented in Chapters 6 and 7, respectively. The values of the performance variables required for the cost model are taken from the ASPEN flowsheet of the EFCC, as documented in Chapters 3 and 4. The values of the performance variables are accessed using an ASPEN Fortran block. The Fortran block calls the cost model subroutine, and passes the values of the performance variables through Fortran common blocks.

This appendix contains listings of the newly developed computer code for the cost model of the EFCC. The code includes the ASPEN Fortran block EFCOST for inclusion in the ASPEN simulation model of the EFCC system and a subroutine USREFCC for estimating process costs. In addition, this appendix contains a sample output from a simulation of the integrated ASPEN performance and newly developed cost model.

D.1 ASPEN Fortran Block EFCOST

A Fortran block is required to access the values of the performance variables in the ASPEN simulation model that are required as inputs to the cost model subroutine. This Fortran block calls the cost model subroutine, and passes the values of the performance variables via common blocks. This Fortran block is the last block to be executed in the ASPEN simulation.

```
;
;**********************************************************************
;**********************************************************************
;      FORTRAN BLOCK 'EFCOST' TO CALL SUBROUTINE 'USREFCC01'
;**********************************************************************
;**********************************************************************
;
;
;
FORTRAN EFCOST
;
; Description:  This block accesses the key variables required to
;               estimate capital and annual costs, and then calls
;               subroutine USREFCC cost model.
;
;
;  All key parameters in the cost model are assigned values in the main
;  program FORTRAN block.  Therefore, it is not necessary to make any
;  changes to the cost model subroutine when redefining case studies.
;
;
;
F      REAL*8 MCFCI, MPWDAI, MFGIDI, DELPID, DFGID, MAAPAI, DELPPA,
F     1       DAAPA, WCONDP, WBFWP, WFGDT, GTNET, STNET,HOC,
F     2       MFGCO, DFGCO, MMUCWI, MSCCWI, MFGFGI, DFGFGI, MSOFGI,
F     3       MASLO, MLFGI, DLFGI, MSFGO, MFGHRI, WHPT, WIPT,
F     4       WLP1T, WLP2T, MSLP2I, HST6A, HFGREH, HST7,
F     5       MSOIDO, MO2IDO, MSTIDO, MFGIDO, MASSFG, MFGFFI, DFGFFI,
F     6       MAGTI, MFGSSI, MSTSTI, MSTI, MAC3O, MAGTEI, MAGTEO,
F     7       MAT1I, MAGTO, MWFGO, LEVHPD, MFGFRO, MSSLI, MWCWW,
F     8       MWGTEI, MCO2SG, MO2CMI, MO2CMO, MSIPT, MSLP1T, MCAFGI
;
F      REAL*8 DCCBV, DCCBS, DCCBM, DCSSB, DCHXBC, DCHXBM, DCHRB,
F     1       DCFB81, DCFB82, DCFB83, DCIDB, AHXBC, AHXBM,
F     2       GFGSSB, MFGHRB, GFGFB, RACB, NBB, MAB, GFGIDB
;
F      REAL*8 CF, EREAL, FEHO, FGF, FICC, FPJ, INTRST, RTAX, ALABOR,
F     1       NSHIFTS, EINFL, RRD, RRP, RRE, DR, PR, TX, XITC, PTI, CEP
;
F      REAL*8 FPCCH, FPCC, FPCSS, FPCHX, FPCBF, FPCGT, FPCHR,
F     1       FPCST, FPCFGD, FPCFF, FPCID, FPCGF
;
F      REAL*8 FMCCH, FMCC, FMCSS, FMCHX, FMCBF, FMCGT, FMCHR,
F     1       FMCST, FMCFGD, FMCFF, FMCID, FMCGF
;
F      REAL*8 BCCOAL, BCRF, BCSA, BCSH, BCSP, BCHY, BCMO, BCLS, BCFB,
F     1       BCFO, BCAA, BCWAT, BCLPG, BCSLG, BCSLD, BCASH, UCLAND
;
F      REAL*8 VFGCO, RTC, TRFC, RCAS, RAC, NB, GENEFF, EF,
F     1       FRACAL, FRACAM, FRACAH, FRACSL, EFSSPL, EFSSPM, EFSSPH,
F     2       EFHXPL, EFHXPM, EFHXPH, EFFFPL, EFFFPM, EFFFPH, NOX,
F     3       FGDEFF, MOTEFF
;
F      REAL*8 SFSS, SFHXC, SFHXM, SFID,
F     1       F81, F82, F83, F84, A81, A82, A83, A84, B81, B82, B83,
F     2       B84, C81, C82, C83, C84
;
F      REAL*8 MWNETE, IC, MCOALIN, MTOTWIN, MTOTAIN, MSSLO, NOXEMS,
```

```
F     1         NOPERS
;
F      REAL*8 GTNETS, PHSHRO, MHSHRO, STNETS, ERRST
;
F      CHARACTER*4 REPOP
;
;   COMMON blocks for input performance variables:
;
;     USRSEN  passes process parameters which are set for sensitivity
;             analysis in the FORTRAN block STCTAIL
;     USRCST3 passes simulation variables accessed in block EFCOST.
;     USER    passes variables from the ASPEN simulation, including NRPT.
;     USRHOC  passes heat of combustion (coal heating value).
;     WRITEO  passes write option switch.
;     USRCST  passes cost model parameters.
;     USRFPC  passes capital cost uncertainty factors.
;     USRFMC  passes maintenance cost factors.
;     USRERR  passes regression model error terms.
;     USRBC   passes unit costs.
;     USRPER  passes design and performance assumptions.
;
;
;
F      COMMON/USRCST3/ MCFCI, MPWDAI, MFGIDI, DELPID, DFGID, MAAPAI,
F     1                DELPPA, DAAPA, WCONDP, WBFWP, WFGDT, GTNET,
F     2                STNET, MFGCO, DFGCO, MMUCWI, MSCCWI, MFGFGI,
F     3                DFGFGI, MSOFGI, MASLO, MLFGI, DLFGI, MSFGO,
F     4                MFGHRI, WHPT, WIPT, WLP1T, WLP2T, MSLP2I,
F     5                HST6A, HFGREH, HST7, MSOIDO, MO2IDO, MSTIDO,
F     6                MFGIDO, MASSFG, MFGFFI, DFGFFI, MAGTI, MFGSSI,
F     7                MSTSTI, FCAHXL, FHAHXL, HXHL, TSTG, MSTI,
F     8                AMBT, GTIT, MAC3O, MAGTEI, MAGTEO, MAT1I, MAGTO,
F     9                MWFGO, FILTPD, STMPD, HXCPD, TMX1PD, TMX2PD,
F     1           TMX3PD,TMX4PD, AIRPD, CMIXPD, COMBPD, SSPD, HALKPD,
F     2                HXHPD, CALKPD, REHHPD, SUPHPD, HEVHPD, HECHPD,
F     3                LEVHPD, FGDMPD, FGDRPD, FABPD, WSTIC, WGTC1,
F     4                WGTC2, WGTC3, MSSLI, RWFGD, TFGCO, COMBHL,
F     5                MWCWW, CWWHI, TFGHRI, TAHXI, TAHXO, MWGTEI,
F     6                PAGTCI, PAGTCO, COLCNV, MCO2SG,
F     7                MO2CMI, MO2CMO, WWINP, HFGAS2, HFGAS7, HCERHX,
F     8                TFGRHI, TFGRHO, MSIPT, MSLP1T, MCAFGI, SULEFF
;
F      COMMON/USRHOC/  HOC
;
F      COMMON/USRBASE/ DCCBV, DCCBS, DCCBM, DCSSB, DCHXBC, DCHXBM,
F     1                DCHRB, DCFB81, DCFB82, DCFB83, DCIDB, AHXBC,
F     2                AHXBM, GFGSSB, MFGHRB, GFGFB, RACB, NBB, MAB,
F     3                GFGIDB
;
F      COMMON/USRCST/ NCSTYR, CF, EREAL, FEHO, FGF, FICC, FPJ, INTRST,
F     1               NOYEARS, IBKLIF, RTAX, ALABOR, NSHIFTS, EINFL, RRD,
F     2               RRP, RRE, DR, PR, TX, XITC, PTI, CEP
;
F      COMMON/USRFPC/ FPCCH, FPCC, FPCSS, FPCHX, FPCBF, FPCGT, FPCHR,
F     1               FPCST, FPCFGD, FPCFF, FPCID, FPCGF
;
F      COMMON/USRFMC/ FMCCH, FMCC, FMCSS, FMCHX, FMCBF, FMCGT, FMCHR,
F     1               FMCST, FMCFGD, FMCFF, FMCID, FMCGF
```

```
;
F       COMMON/USRERR/ ERRST
;
F       COMMON/USRBC/  BCCOAL, BCRF, BCSA, BCSH, BCSP, BCHY, BCMO, BCLS,
F     1                BCFB, BCFO, BCAA, BCWAT, BCLPG, BCSLG, BCSLD,
F     2                BCASH, UCLAND
;
F       COMMON/USRPER/ NOC, NTC, NOHX, NTHX, NOGT, NTGT, NOHR, NTHR,
F     1                VFGCO, RTC, TRFC, RCAS, RAC, NB,
F     2                GENEFF, EF, FRACAL, FRACAM, FRACAH, FRACSL,
F     3                EFSSPL, EFSSPM, EFSSPH, EFHXPL, EFHXPM, EFHXPH,
F     4                EFFFPL, EFFFPM, EFFFPH, NOX, FGDEFF, MOTEFF,
F     5                HHVCOL
;
F       COMMON/USRSF/  SFSS, SFHXC, SFHXM, SFID,
F     1                F81, F82, F83, F84, A81, A82, A83, A84, B81, B82,
F     2                B83, B84, C81, C82, C83, C84
;
;
F       COMMON/USROUT/ HEATRT, EFFHHV, WGTE, WSTE, WAUXE, MWNETE,
F     1                DCCH, DCC, DCGT, DCSS, DCHX, DCHR, DCST, DCBF,
F     2                DCFGD, DCFF, DCID, DCGF, TDC, CICC, CTAX, CEHO,
F     3                TIC, CPC, CPJ, TPC, AF, TPI, AFDC, PPC, IC,
F     4                TCICC, TCLAND, TCR, OCL, OCM, OCAS, FOC, OCFUEL,
F     5                OCCONS, OCSLG, OCASH, OCSLD, VOC, DPERKW, FOCN,
F     6                VOCN, FUELN, VOCINC, CELEC,
F     7                MCOALIN, MTOTWIN, MTOTAIN, FAEMS, MSSLO, BATOT,
F     8                SLDG, SOEMSD, NOXEMS, CO2EMS, FCF, VCLF, NOPERS
;
F       COMMON/WRITEO/ REPOP
;
;-----------------------------------------------------------------------
;  VARIABLE DEFINITIONS OF SIMULATION VARIABLES ACCESSED
;
;
;   MCFCI  = Mass flow of coal to combustor, lb/hr
;   MFGIDI = Mass flow of flue gas to the ID fan inlet, lb/hr
;   DELPID = Pressure of the flue gas stream to the ID fan inlet, psia
;   DFGID  = Average density of flue gas to the ID fan inlet, lb/ft3
;   MAAPAI = Mass flow rate of aux. air to the primary air fan inlet, lb/hr
;   DELPPA = Pressure of the air stream from the primary air fan outlet,
;            psia
;   DAAPA  = Average density of aux. air into the primary air fan inlet,
;            lb/ft3
;   WBFWP  = Work requirement of the boiler feedwater pump, Btu/hr
;   WFGDT  = Work requirement of the FGD unit, Btu/hr
;   GTNET  = Net work output of the gas turbine, Btu/hr
;   STNET  = Net work output of steam turbines, Btu/hr
;   HOC    = Heat of combustion of coal, J/Kg
;   MPWDAI = Mass flow fo polished water to deaerator inlet, lb/hr
;   WCONDP = Work required by condenser pump, Btu/hr
;   HHV    = Higher heating value of coal, J/Kg
;   WGTE   = Work output of gas turbine, Btu/hr
;   WSTE   = Work output of steam turbine, Btu/hr
;   MFGCO  = Mass flow of flue gas at combustor outlet, lb/hr
;   DFGCO  = Density of flue gas at combustor outlet, lb/ft3
;   MMUCWI = Mass flow of makeup water at condenser wall outlet, lb/hr
;   MSCCWI = Mass flow of steam condensate at condenser wall outlet, lb/hr
```

```
;   MFGFGI = Mass flow of flue gas at FGD inlet, lb/hr
;   DFGFGI = Density of flue gas at FGD inlet, lb/hr
;   MSOFGI = Mass flow of SO2 at FGD inlet, lb/hr
;   MASLO  = Mass flow of ash at slagger outlet
;   MLFGI  = Mass flow of limestone slurry at FGD inlet, lb/hr
;   DLFGI  = Density of limestone slurry at FGD inlet, lb/ft3
;   MSFGO  = Mass flow of sludge at FGD outlet, lb/hr
;   MFGHRI = Mass flow of flue gas at HRSG inlet, lb/hr
;   WHPT   = Work output of high pressure steam turbine, Btu/hr
;   WIPT   = Work output of intermediate pressure steam turbine, Btu/hr
;   WLP1T  = Work output of first stage low pressure steam turbine, Btu/hr
;   WLP2T  = Work output of second stage low pressure steam turbine, Btu/hr
;   MSLP2I = Mass flow of steam at second stage low pressure
;            steam turbine inlet, lb/hr
;   HST6A  = Enthalpy of steam at second stage low pressure
;            steam turbine inlet, Btu/lb
;   HFGREH = Enthalpy (heat) requirement for FGD reheat, Btu/hr
;   HST7   = Enthalpy of steam at second stage low pressure
;            steam turbine outlet, Btu/hr
;   MSOIDO = Moleflow of SO2 at ID outlet (stack), lbmol/hr
;   MO2IDO = Moleflow of O2 at ID outlet (stack), lbmol/hr
;   MSTIDO = Moleflow of H2O (steam) at ID outlet (stack), lbmol/hr
;   MFGIDO = Moleflow of fluegas at ID outlet (stack), lbmol/hr
;   NOX    = NOx in fluegas at stack outlet, lb/MMBtu
;   MASSFG = Massflow of fluegas at stack outlet, lb/hr
;   MFGFFI = Massflow of fluegas at FF inlet, lb/hr
;   DFGFFI = Density of fluegas at FF inlet, lb/ft3
;-----------------------------------------------------------------------
;
;  DEFINE STATEMENTS TO ACCESS SIMULATION VARIABLES
;
;
 DEFINE MCFCI SUBSTREAM-VAR STREAM=COALFEED SUBS=NC VAR=MASSFLOW
 DEFINE MPWDAI STREAM-VAR STREAM=ST10 VAR=MASSFLOW
 DEFINE MFGIDI SUBSTREAM-VAR STREAM=FGAS12 SUBS=MIXED VAR=MASSFLOW
 DEFINE DELPID SUBSTREAM-VAR STREAM=FGAS12 SUBS=MIXED VAR=PRES
 DEFINE DFGID SUBSTREAM-VAR STREAM=FGAS12  SUBS=MIXED VAR=MASS-DENSITY
 DEFINE MAAPAI SUBSTREAM-VAR STREAM=AUXAIR SUBS=MIXED VAR=MASSFLOW
 DEFINE DELPPA SUBSTREAM-VAR STREAM=AIRIN SUBS=MIXED VAR=PRES
 DEFINE DAAPA SUBSTREAM-VAR STREAM=AUXAIR SUBS=MIXED VAR=MASS-DENSITY
 DEFINE WCONDP STRM-ATTR-VAR STREAM=WPUMP1 ATTR=WORK VAR=POWER
 DEFINE WBFWP STRM-ATTR-VAR STREAM=WPUMP2 ATTR=WORK VAR=POWER
 DEFINE WFGDT STRM-ATTR-VAR STREAM=WFGD ATTR=WORK VAR=POWER
 DEFINE GTNET STRM-ATTR-VAR STREAM=WGT ATTR=WORK VAR=POWER
 DEFINE STNET STRM-ATTR-VAR STREAM=WST ATTR=WORK VAR=POWER
 DEFINE MFGCO SUBSTREAM-VAR STREAM=FLUEGAS SUBS=MIXED VAR=MASSFLOW
 DEFINE DFGCO SUBSTREAM-VAR STREAM=FLUEGAS SUBS=MIXED VAR=MASS-DENSITY
 DEFINE MMUCWI STREAM-VAR STREAM=MAKEUP VAR=MASSFLOW
 DEFINE MSCCWI STREAM-VAR STREAM=ST7 VAR=MASSFLOW.
 DEFINE MFGFGI SUBSTREAM-VAR STREAM=FGAS7 SUBS=MIXED VAR=MASSFLOW
 DEFINE DFGFGI SUBSTREAM-VAR STREAM=FGAS7 SUBS=MIXED VAR=MASS-DENSITY
 DEFINE MSOFGI MASS-FLOW STREAM=FGAS7 SUBS=MIXED COMPONENT=SO2
 DEFINE MASLO SUBSTREAM-VAR STREAM=ASH SUBS=NC VAR=MASSFLOW
 DEFINE MLFGI SUBSTREAM-VAR STREAM=FGDLMST SUBS=MIXED VAR=MASSFLOW
 DEFINE DLFGI SUBSTREAM-VAR STREAM=FGDLMST SUBS=MIXED VAR=MASS-DENSITY
 DEFINE MSFGO SUBSTREAM-VAR STREAM=SLUDGE SUBS=CISOLID VAR=MASSFLOW
 DEFINE MFGHRI STREAM-VAR STREAM=FGAS2 VAR=MASSFLOW
 DEFINE WHPT  STRM-ATTR-VAR STREAM=WHPTURB  ATTR=WORK  VAR=POWER
```

```
 DEFINE WIPT  STRM-ATTR-VAR STREAM=WIPTURB  ATTR=WORK  VAR=POWER
 DEFINE WLP1T STRM-ATTR-VAR STREAM=WLPTURB1 ATTR=WORK  VAR=POWER
 DEFINE WLP2T STRM-ATTR-VAR STREAM=WLPTURB2 ATTR=WORK  VAR=POWER
 DEFINE MSLP2I STREAM-VAR STREAM=ST6A VAR=MASSFLOW
 DEFINE HST6A  STREAM-VAR STREAM=ST6A VAR=MASS-ENTH
 DEFINE HFGREH STRM-ATTR-VAR STREAM=QFGDREH ATTR=HEAT VAR=Q
 DEFINE HST7   STREAM-VAR STREAM=ST7 VAR=MASS-ENTH
 DEFINE MSOIDO MOLE-FLOW STREAM=STACKGAS SUBS=MIXED COMP=SO2
 DEFINE MO2IDO MOLE-FLOW STREAM=STACKGAS SUBS=MIXED COMP=O2
 DEFINE MSTIDO MOLE-FLOW STREAM=STACKGAS SUBS=MIXED COMP=H2O
 DEFINE MFGIDO STREAM-VAR STREAM=STACKGAS  VAR=MOLE-FLOW
 DEFINE MASSFG STREAM-VAR STREAM=STACKGAS VAR=MASSFLOW
 DEFINE MFGFFI SUBSTREAM-VAR STREAM=FGAS7B SUBS=MIXED VAR=MASSFLOW
 DEFINE DFGFFI SUBSTREAM-VAR STREAM=FGAS7B SUBS=MIXED VAR=MASS-DENSITY
 DEFINE MAGTI  STREAM-VAR STREAM=AIR   VAR=MASSFLOW
 DEFINE MFGSSI STREAM-VAR STREAM=FGAS1 VAR=MASSFLOW
 DEFINE MSTSTI STREAM-VAR STREAM=ST1   VAR=MASSFLOW
;
;
 DEFINE FCAHXL BVAR BLOCK=AIRLEAKC SENTENCE=FRAC VAR=FRAC &
                    ID1=MIXED ID2=HXALKC
 DEFINE FHAHXL BVAR BLOCK=AIRLEAKH SENTENCE=FRAC VAR=FRAC &
                    ID1=MIXED ID2=HXALKH
 DEFINE HXHL   BVAR BLOCK=CERHXHL  SENTENCE=PARAM VAR=FACTOR
 DEFINE TSTG SUBSTREAM-VAR STREAM=STACKGAS SUBS=MIXED VAR=TEMP
 DEFINE MSTI STREAM-VAR STREAM=IPSTEAM VAR=MASSFLOW
 DEFINE AMBT STREAM-VAR STREAM=AIR VAR=TEMPERATURE
 DEFINE GTIT STREAM-VAR STREAM=HTAR1I VAR=TEMPERATURE
 DEFINE MAC3O STREAM-VAR STREAM=HPAR3O VAR=MASSFLOW
 DEFINE MAGTEI STREAM-VAR STREAM=HPAIR VAR=MASSFLOW
 DEFINE MAGTEO STREAM-VAR STREAM=SATAIRC VAR=MASSFLOW
 DEFINE MAT1I STREAM-VAR STREAM=HTAR1I VAR=MASSFLOW
 DEFINE MAGTO STREAM-VAR STREAM=EXH-AIR VAR=MASSFLOW
 DEFINE MWFGO SUBSTREAM-VAR STREAM=SLUDGE SUBS=MIXED VAR=MASSFLOW
;
; Accessing the pressure drop along the air flow path
;
 DEFINE FILTPD  BVAR BLOCK=FILTER    SENTENCE=PARAM VAR=PRES
 DEFINE STMPD   BVAR BLOCK=GT-EVAP   SENTENCE=PARAM VAR=PRES
 DEFINE HXCPD   BVAR BLOCK=CERHXC    SENTENCE=PARAM VAR=PRES
 DEFINE TMX1PD  BVAR BLOCK=GT-TMIX1  SENTENCE=PARAM VAR=PRES
 DEFINE TMX2PD  BVAR BLOCK=GT-TMIX2  SENTENCE=PARAM VAR=PRES
 DEFINE TMX3PD  BVAR BLOCK=GT-TMIX3  SENTENCE=PARAM VAR=PRES
 DEFINE TMX4PD  BVAR BLOCK=GT-TMIX4  SENTENCE=PARAM VAR=PRES
;
; Accessing the pressure drop along the flue gas flow path
;
; 1. Combustor section
 DEFINE AIRPD   BVAR BLOCK=AIRBLOW   SENTENCE=PARAM VAR=PRES
 DEFINE CMIXPD  BVAR BLOCK=MIXING    SENTENCE=PARAM VAR=PRES
 DEFINE COMBPD  BVAR BLOCK=COMBUSTR  SENTENCE=PARAM VAR=PRES
 DEFINE SSPD    BVAR BLOCK=SLSCRN    SENTENCE=PARAM VAR=PRES
 DEFINE HALKPD  BVAR BLOCK=MIXHAIR   SENTENCE=PARAM VAR=PRES
 DEFINE HXHPD   BVAR BLOCK=CERHXH    SENTENCE=PARAM VAR=PRES
 DEFINE CALKPD  BVAR BLOCK=MIXCAIR   SENTENCE=PARAM VAR=PRES
;
; 2. HRSG section
 DEFINE REHHPD  BVAR BLOCK=REHEATH  SENTENCE=PARAM VAR=PRES
```

```
  DEFINE SUPHPD  BVAR BLOCK=SUPHTXH  SENTENCE=PARAM VAR=PRES
  DEFINE HEVHPD  BVAR BLOCK=HPEVAPH  SENTENCE=PARAM VAR=PRES
  DEFINE HECHPD  BVAR BLOCK=HPECONH  SENTENCE=PARAM VAR=PRES
  DEFINE LEVHPD  BVAR BLOCK=LPEVAPH  SENTENCE=PARAM VAR=PRES
;
;3. FGD
  DEFINE FGDMPD  BVAR BLOCK=FGDMIX SENTENCE=PARAM VAR=PRES
  DEFINE FGDRPD  BVAR BLOCK=FGDRXR SENTENCE=PARAM VAR=PRES
;
;4. Fabric Filter
  DEFINE FABPD   BVAR BLOCK=FFPD   SENTENCE=PARAM VAR=PRES
;
  DEFINE WSTIC STRM-ATTR-VAR STREAM=WGT-SI ATTR=WORK VAR=POWER
  DEFINE WGTC1 STRM-ATTR-VAR STREAM=WGT-C1 ATTR=WORK VAR=POWER
  DEFINE WGTC2 STRM-ATTR-VAR STREAM=WGT-C2 ATTR=WORK VAR=POWER
  DEFINE WGTC3 STRM-ATTR-VAR STREAM=WGT-C3 ATTR=WORK VAR=POWER
;
  DEFINE MSSLI SUBSTREAM-VAR STREAM=FGDLMST SUBS=CISOLID VAR=MASSFLOW
  DEFINE RWFGD SUBSTREAM-VAR STREAM=RECWAT SUBS=MIXED VAR=MASSFLOW
  DEFINE TFGCO SUBSTREAM-VAR STREAM=DIRTYGAS SUBS=MIXED VAR=TEMP
  DEFINE COMBHL STRM-ATTR-VAR STREAM=QCLOSSA ATTR=HEAT VAR=Q
  DEFINE MWCWW STREAM-VAR STREAM=WATWWI VAR=MASSFLOW
  DEFINE CWWHI STRM-ATTR-VAR STREAM=QCLOSSWW ATTR=HEAT VAR=Q
  DEFINE TFGHRI  STREAM-VAR STREAM=FGAS2 VAR=TEMP
  DEFINE TAHXI STREAM-VAR STREAM=SATAIRC VAR=TEMP
  DEFINE TAHXO STREAM-VAR STREAM=SATAIRH VAR=TEMP
  DEFINE MWGTEI STREAM-VAR STREAM=WATINJI VAR=MASSFLOW
;
  DEFINE PAGTCI STREAM-VAR STREAM=AIR VAR=PRES
  DEFINE PAGTCO STREAM-VAR STREAM=HPAR3O VAR=PRES
  DEFINE COLCNV BVAR BLOCK=COMBUSTR SENTENCE=CONV VAR=CONV ID1=1 &
                ID2=NC ID3=COAL
;
  DEFINE MCO2SG MASS-FLOW STREAM=STACKGAS SUBS=MIXED COMPONENT=CO2
  DEFINE MO2CMI MOLE-FLOW STREAM=CMIX SUBS=MIXED COMP=O2
  DEFINE MO2CMO MOLE-FLOW STREAM=DIRTYGAS SUBS=MIXED COMP=O2
  DEFINE WWINP STRM-ATTR-VAR STREAM=WWINJP ATTR=WORK VAR=POWER
;
  DEFINE HFGAS2 STREAM-VAR STREAM=FGAS2 VAR=MASS-ENTH
  DEFINE HFGAS7 SUBSTREAM-VAR STREAM=FGAS7 SUBS=MIXED VAR=MASS-ENTH
  DEFINE HCERHX STRM-ATTR-VAR STREAM=QCERHX ATTR=HEAT VAR=Q
;
  DEFINE TFGRHI SUBSTREAM-VAR STREAM=FGAS11 SUBS=MIXED VAR=TEMP
  DEFINE TFGRHO SUBSTREAM-VAR STREAM=FGAS12 SUBS=MIXED VAR=TEMP
;
  DEFINE MSIPT  STREAM-VAR STREAM=ST3 VAR=MASSFLOW
  DEFINE MSLP1T STREAM-VAR STREAM=ST5 VAR=MASSFLOW
;
  DEFINE MCAFGI MOLE-FLOW STREAM=FGDLMST SUBS=CISOLID COMP=CACO3
  DEFINE SULEFF BVAR BLOCK=FGDRXR SENTENCE=CONV VAR=CONV ID1=1    &
                ID2=MIXED ID3=SO2
;
  DEFINE COALMI COMPONENT-ATTR STREAM=COALFEED SUBSTREAM=NC         &
         COMPONENT=COAL ATTRIBUTE=COALMISC
;--------------------------------------------------------------------------
;----------------------VALUES FOR BASE CASE EQUIPMENT-----------------------
;
;  DCCBV  = Direct capital cost of base case combustor vessel, $1,000
```

```
;  DCCBS  = Direct capital cost of base case combustor structural supports,
;           $1,000
;  DCCBM  = Direct capital cost of base case combustor miscellaneous
;           features, $1,000
;  DCSSB  = Direct capital cost of base case slag screen, $1,000
;  DCHXBC = Direct capital cost of ceramic part of base case heat
;           exchanger, $1,000
;  DCHXBM = Direct capital cost of metallic part of base case heat
;           exchanger, $1000
;  DCHRB  = Direct capital cost of base case HRSG, $1,000
;  DCFB81 = Direct capital cost of process area 81 of base case fabric
;           filter, 1990  $1000
;  DCFB82 = Direct capital cost of process area 82 of base case fabric
;           filter, 1990  $1000
;  DCFB83 = Direct capital cost of process area 83 of base case fabric
;           filter, 1990  $1000
;  DCIDB  = Direct capital cost of base case ID fan, $1,000
;  AHXBC  = Surface area of ceramic part of base case heat exchanger, ft2
;  AHXBM  = Surface area of mettalic part of base case heat exchanger, ft2
;  GFGSSB = Volumetric flow rate of flue gas through base case slag screen,
;           ft3/hr
;  MFGHRB = Mass flow rate of flue gas through base case HRSG, lb/hr
;  GFGFB  = Volumetric flow rate of flue gas through base case fabric
;           filter, ft3/hr
;  RACB   = Air to cloth ratio of base case fabric filter, acfm/ft2
;  NBB    = Number of filter bags per compartment of base case fabric filter
;  MAB    = Flow rate of bottom ash from base case fabric filter, tons/hr
;  GFGIDB = Volumetric flue gas flow rate through base case ID fan, ft3/hr
;
F     DCCBV = 0.0
F     DCCBS = 0.0
F     DCCBM = 0.0
F     DCSSB = 0.0
F     DCHXBC = 0.0
F     DCHXBM = 0.0
F     DCHRB = 11050.0
F     DCFB81 = 7860.0
F     DCFB82 = 250.0
F     DCFB83 = 1590.0
F     DCIDB = 0.0
F     AHXBC = 1.0
F     AHXBM = 1.0
F     GFGSSB = 100000000.0
F     MFGHRB = 3842000.0
F     GFGFB = 950000.0
F     RACB = 2.0
F     NBB = 360.0
F     MAB = 10.44
F     GFGIDB = 100000000.0
;---------------------COST MODEL PARAMETER DEFINITION---------------------
;
;   NCSTYR = Year of plant construction
;   CF     = Annual plant capacity factor (decimal)
;   EREAL  = Annual real escalation rate (decimal)
;   FEHO   = Engineering and home office fee factor (decimal)
;   FGF    = General facilities cost (fraction of other direct cost)
;   FICC   = Indirect contruction cost factor (decimal)
;   FPJ    = Project contingency factor (decimal)
```

```
;    INTRST = Cost of capital during construction (decimal)
;    NOYEARS= Number of years of construction (years)
;    IBKLIF = Book life of project (years)
;    RATX   = Sales tax rate (decimal)
;    ALABOR = Average labor rate, including burdens ($/hour)
;    NSHIFTS= Number of work shifts per day (including vacations)
;    EINFL  = Inflation rate (decimal)
;    RRD    = Real rate of return on debt (decimal)
;    RRP    = Real rate of return on preferred stock (decimal)
;    RRE    = Real rate of return on equity (decimal)
;    DR     = Debt ratio (decimal)
;    PR     = Preferred stock ratio (decimal)
;    TX     = Federal and state income tax (decimal)
;    XITC   = Income tax credit (decimal)
;    PTI    = Property taxes and insurance (decimal)
;    CEP    = Cost of Environmental Permits (thousand $)
;
;
F      NCSTYR = 1990
F      CF     = 0.65
F      EREAL  = 0.0
F      FEHO   = 0.10
F      FGF    = 0.15
F      FICC   = 0.20
F      FPJ    = 0.30
F      INTRST = 0.05
F      NOYEARS= 4
F      IBKLIF = 30
F      RTAX   = 0.06
F      ALABOR = 19.70
F      NSHIFTS= 4.75
F      EINFL  = 0.00
F      RRD    = 0.046
F      RRP    = 0.052
F      RRE    = 0.087
F      DR     = 0.5
F      PR     = 0.15
F      TX     = 0.38
F      XITC   = 0.00
F      PTI    = 0.02
F      CEP    = 1000.0
;------------------------------------------------------------------------
;---------------------CAPITAL COST UNCERTAINTY FACTORS DEFINITION---------
;                              (PROCESS CONTINGENCY FACTORS)
;
;
;
F      FPCCH = 0.01
F      FPCC  = 0.7
F      FPCSS = 1.0
F      FPCHX = 1.0
F      FPCBF = 0.0
F      FPCGT = 0.25
F      FPCHR = 0.2
F      FPCST = 0.0
F      FPCFGD= 0.10
F      FPCFF = 0.10
F      FPCID = 0.10
```

```
F     FPCGF = 0.05
;
;-------------------------------------------------------------------------
;---------------------MAINTENANCE COST FACTORS---------------------------
;
;
;
F     FMCCH = 0.03
F     FMCC  = 0.04
F     FMCSS = 0.03
F     FMCHX = 0.05
F     FMCBF = 0.015
F     FMCGT = 0.015
F     FMCHR = 0.015
F     FMCST = 0.015
F     FMCFGD= 0.044
F     FMCFF = 0.0106
F     FMCID = 0.03
F     FMCGF = 0.03
;-------------------------------------------------------------------------
;---------------------REGRESSION MODEL ERROR TERMS-----------------------
;
;   ERRST = Regression model error term for steam turbine
;
F     ERRST = 0.0
;-------------------------------------------------------------------------
;--------------------UNIT COSTS AND PRICES-------------------------------
;
;  BCCOAL = Unit price of coal, $/MMBtu
;  BCRF   = Unit price of refractory, $/ft2
;  BCSA   = Unit price of sulfuric acid (93%), $/ton
;  BCSH   = Unit price of sodium hydroxide (50%), $/ton
;  BCSP   = Unit price of sodium phophate, $/lb
;  BCHY   = Unit price of hydrazine, $/lb
;  BCMO   = Unit price of morpholine, $/lb
;  BCLS   = Unit price of limestone, $/ton
;  BCFB   = Unit price of filter bags, $/ft2 bag
;  BCFO   = Unit price of fuel oil, $/bbl
;  BCAA   = Unit price of air adsorbent, $/lb
;  BCWAT  = Unit price of water, $/1000 gal
;  BCLPG  = Unit price of LPG, $/bbl
;  BCSLG  = Unit cost of slag disposal, $/ton
;  BCSLD  = Unit cost of sludge disposal, $/ton (dry)
;  BCASH  = Unit cost of ash disposal, $/ton
;
F     BCCOAL = 1.61
F     BCRF   = 0.0
F     BCSA   = 110.0
F     BCSH   = 220.0
F     BCSP   = 0.70
F     BCHY   = 3.20
F     BCMO   = 1.30
F     BCLS   = 18.0
F     BCFB   = 82.0
F     BCFO   = 42.0
F     BCAA   = 2.80
F     BCWAT  = 0.73
F     BCLPG  = 11.70
```

```
F     BCSLG  = 10.0
F     BCSLD  = 11.30
F     BCASH  = 10.0
;-----------------------UNIT COST OF LAND---------------------------------
;
;   UCLAND = Cost of land, $/acre
;
F     UCLAND = 6990.0
;---------------------------------------------------------------------------
;--------------------DESIGN AND PERFORMANCE ASSUMPTION---------------------
;
;  VFGCO  = Superficial flue gas velocity at combustor outlet, ft/sec
;  RTC    = Overall flue gas residence time in combustor, sec
;  TRFC   = Thickness of combustor refractory, ft
;  RCAS   = Calcium to sulfur molar ratio for FGD unit, decimal
;  RAC    = Air to cloth ratio for the FF, decimal
;  NB     = Number of filter bags per compartment of base case power
;           plant FF unit, integer
;  GENEFF = Gas turbine and Steam turbine generator efficiency, fraction
;  EF     = Centrigual fan efficiency, fraction
;  MOTEFF = Electric motor efficiency, fraction
;  FGDEFF = SO2 removal efficiency of FGD unit, decimal
;  NOX    = NOx concentration in stack gas, lb/MMBtu
;
F     NOC  = 1
F     NTC  = 1
F     NOHX = 1
F     NTHX = 1
F     NOHR = 1
F     NTHR = 1
F     NOGT = 1
F     NTGT = 1
F     VFGCO  = 4.0
F     RTC    = 8.0
F     TRFC   = 0.667
F     RCAS   = 1.05
F     RAC    = 2.0
F     NB     = 360.0
F     GENEFF = 0.986
F     EF     = 0.85
F     FGDEFF = 0.90
F     NOX    = 0.5
F     MOTEFF = 0.95
;
;  ASH SIZE DIST
;  FRACAL = Fraction of ash coming out of combustor in the low size range
;  FRACAM = Fraction of ash coming out of combustor in the medium size range
;  FRACAH = Fraction of ash coming out of combustor in the high size range
;  FRACSL = Fraction of ash in coal which leaves the combustor as slag
;
F     FRACAL = 0.25
F     FRACAM = 0.50
F     FRACAH = 0.25
F     FRACSL = 0.65
;
;  PARTICLE REMOVAL EFF
;
;  EFSSPL = Particle removing efficiency of slag screen in the low size range
```

```
;  EFSSPM = Particle removing efficiency of slag screen in the medium size
range
;  EFSSPH = Particle removing efficiency of slag screen in the high size range
;  EFHXPL = Particle removing efficiency of CerHx in the low size range
;  EFHXPM = Particle removing efficiency of CerHx in the medium size range
;  EFHXPH = Particle removing efficiency of CerHx in the high size range
;  EFFFPL = Particle removing efficiency of fabric filter in the low size
range
;  EFFFPM = Particle removing efficiency of fabric filter in the medium size
;           range
;  EFFFPH = Particle removing efficiency of fabric filter in the high size
range
;
F     EFSSPL = 0.001
F     EFSSPM = 0.99
F     EFSSPH = 0.001
F     EFHXPL = 0.001
F     EFHXPM = 0.001
F     EFHXPH = 0.99
F     EFFFPL = 0.99
F     EFFFPM = 0.99
F     EFFFPH = 0.99
;
F     HHVCOL = COALMI(1)
;---------------------------------------------------------------------------
;------------------------DC SCALING FACTORS AND FF EXPONENTS-------------
;
;
F     SFSS  = 1.0
F     SFHXC = 1.0
F     SFHXM = 1.0
F     SFID  = 1.0
F     F81   = 0.86
F     F82   = 0.75
F     F83   = 0.55
F     F84   = 0.0
F     A81   = 0.84
F     A82   = 0.0
F     A83   = 0.29
F     A84   = 0.0
F     B81   = 0.15
F     B82   = 0.0
F     B83   = 0.275
F     B84   = 0.0
F     C81   = 0.0
F     C82   = 0.0
F     C83   = 0.083
F     C84   = 0.0
;
;---------------------------------------------------------------------------
;-------------------------WRITE OPTION SWITCH--------------------------
;
;  REPOP controls whether the detailed cost model assumptions and output
;  summaries are printed in the report file.  For probabilistic analysis
;  with large sample sizes, the recommended value is 'NO', which suppresses
;  all written outputs, including range check warnings.  For each sample,
;  the cost model output is approximately three pages.
;
```

```
F      REPOP = 'YES'
;
;-----------------------------------------------------------------------
;
F      CALL USREFCC55
;------------------------------------------------------------------------;
```

D.2 Cost Model Subroutine USREFCC

The cost model subroutine calculates the capital and annual costs for the EFCC. This subroutine calls subroutines USRWCK, USRRCK, USRFCF, and USRVCLF. USRWCK and USRRCK are used for range-checking and are documented by Frey and Rubin (1990). USRFCF and USRVCLF are used to calculate the fixed charge and variable cost levelization factors required to estimate the cost of electricity, as documented by Frey and Rubin (1990).

```
      SUBROUTINE USREFCC54
      IMPLICIT DOUBLE PRECISION (A-H,O-Z)
C
C     Revision 1.0  P. Agarwal                 Date:  July 28, 1995
C
C     Short Title: Cost Model for the Externally Fired Combined Cycle
C     Description: Capital and Annual Cost Model of Externally Fired
C                  Combined Cycle.  Based on ASPEN input file developed
C                  by P. Agarwal.
C     Cost model documented in
C
C
C
      REAL*8 MCFCI, MPWDAI, MFGIDI, DELPID, DFGID, MAAPAI, DELPPA,
     1       DAAPA, WCONDP, WBFWP, WFGDT, GTNET, STNET, HOC, HHV,
     2       MWGRE, MWNETE, EFFHHV,
     3       MFGCO, DFGCO, MMUCWI, MSCCWI, MFGFGI, DFGFGI, MSOFGI,
     4       MASLO, MLFGI, DLFGI, MSFGO, MFGHRI, MSSLO, MFASLO,
     5       WHPT, WIPT, WLP1T, WLP2T, MSLP2I, HST6A, HFGREH, HST7,
     6       RAT, ENTHR, ENTHRA, MSLP2A, WLP2TA, MSOIDO, MO2IDO,
     7       MSTIDO, MFGIDO, MASSFG, MFGFFI, DFGFFI
     8       FATOT, ACTO2, FADO2, FAEMS, SOEMS, MAGTI, MFGSSI, MSTSTI,
     9       MSTI, MAC3O, MAGTEI, MAGTEO, MAT1I, MAGTO, MWFGO,
     1       LEVHPD, MSLFGI, MSSLI, MWCWW, MWGTEI, MCO2SG, MO2CMI,
     2       MO2CMO, MSIPT, MSLP1T, MCAFGI
C
      REAL*8 DCCBV, DCCBS, DCCBM, DCHXBC, DCHXBM, DCHRB,
     1       DCFB81, DCFB82, DCFB83, DCIDB, AHXBC, AHXBM,
     2       GFGSSB, MFGHRB, GFGFB, RACB, NBB, MAB, GFGIDB
C
      REAL*8 CF, EREAL, FEHO, FGF,  FICC, FPJ, INTRST,
     1       RTAX, ALABOR, NSHIFTS, EINFL, RRD, RRP, RRE, DR, PR, TX,
     2       XITC, PTI, CEP
C
      REAL*8 FPCCH, FPCC, FPCSS, FPCHX, FPCBF, FPCGT, FPCHR, FPCST,
     1       FPCFGD, FPCFF, FPCID, FPCGF
C
      REAL*8 FMCCH, FMCC, FMCSS, FMCHX, FMCBF, FMCGT, FMCHR, FMCST,
     1       FMCFGD, FMCFF, FMCID, FMCGF
C
      REAL*8 BCCOAL, BCRF, BCSA, BCSH, BCSP, BCHY, BCMO, BCLS, BCFB,
     1       BCFO, BCAA, BCWAT, BCLPG, BCSLG, BCSLD, BCASH, UCLAND
C
      REAL*8 VFGCO, RTC, TRFC, RCAS, RAC, NB, GENEFF, EF,
     1       FRACAL, FRACAM, FRACAH, FRACSL, EFSSPL, EFSSPM, EFSSPH,
     2       EFHXPL, EFHXPM, EFHXPH, EFFFPL, EFFFPM, EFFFPH, NOX,
     3       FGDEFF, MOTEFF
C
      REAL*8 SFSS, SFHXC, SFHXM, SFID,
     1       F81, F82, F83, F84, A81, A82, A83, A84, B81, B82, B83,
     2       B84, C81, C82, C83, C84
C
      REAL*8 GTNETS, PHSHRO, MHSHRO, STNETS,
     1       ERRST
C
C
      REAL*8 DCCH, DCC, DCSS, DCHX, DCGT, DCHR, DCST, DCBF,
     1       GFGCO, IRC, IRVC, HC, SAC, DCCV, DCCR, DCCS, DCCM,
```

```
     2        AHXC, AHXM, DCHXC, DCHXM, GFGFIDI
     3        WGTE, WSTE, CI, CICPPI, TPI, FCF, VCLF
C
      REAL*8 MPW, MRW, MAFFO, MSABFI, MSHBFI, MSPBFI, MHYBFI, MMOBFI,
     1        MSABFPI, MSHBFPI, MLS, MFGDW, MFOGTI, MAAGFI, MTOTW,
     2        MTWAT, MLPG, MSA, MSH, IC, MCOALIN, MFGDWIN, MRWIN,
     3        MTOTWIN, MGTAIN, MAUXAIN, MTOTAIN, NOEMS, NO2EMS, NOXEMS,
     4        FALEAKP
C
      CHARACTER*4 REPOP
C
C-----------------------------------------------------------------------
C
C   COMMON BLOCKS for input performance variables:
C
C   USRSEN  passes process parameters which are set for sensitivity analysis
C           from the FORTRAN block STCTAIL
C   USRCST3 passes simulation variables accessed in block EFCOST
C   USER    passes variables from the ASPEN simulation, including NRPT.
C   USRHOC  passes heat of combustion (coal heating value).
C   WRITEO  passes write option switch.
C   USRCST  passes cost model parameters.
C   USRFPC  passes capital cost uncertainty factors.
C   USRFMC  passes maintenance cost factors.
C   USRERR  passes regression model error terms.
C   USRBC   passes unit costs.
C   USRPER  passes design and performance assumptions.
C
C
C
      COMMON/USRCST3/ MCFCI, MFWDAI, MFGIDI, DELPID, DFGID, MAAPAI,
     1                DELPPA, DAAPA, WCONDP, WBFWP, WFGDT, GTNET,
     2                STNET, MFGCO, DFGCO, MMUCWI, MSCCWI, MFGFGI,
     3                DFGFGI, MSOFGI, MASLO, MLFGI, DLFGI, MSFGO,
     4                MFGHRI, WHPT, WIPT, WLP1T, WLP2T, MSLP2I,
     5                HST6A, HFGREH, HST7, MSOIDO, MO2IDO, MSTIDO,
     6                MFGIDO, MASSFG, MFGFFI, DFGFFI, MAGTI, MFGSSI,
     7                MSTSTI, FCAHXL, FHAHXL, HXHL, TSTG, MSTI,
     8                AMBT, GTIT, MAC3O, MAGTEI, MAGTEO, MAT1I, MAGTO,
     9                MWFGO, FILTPD, STMPD, HXCPD, TMX1PD, TMX2PD,
     1          TMX3PD,TMX4PD, AIRPD, CMIXPD, COMBPD, SSPD, HALKPD,
     2                HXHPD, CALKPD, REHHPD, SUPHPD, HEVHPD, HECHPD,
     3                LEVHPD, FGDMPD, FGDRPD, FABPD, WSTIC, WGTC1,
     4                WGTC2, WGTC3, MSSLI, RWFGD, TFGCO, COMBHL,
     5                MWCWW, CWWHI, TFGHRI, TAHXI, TAHXO, MWGTEI,
     6                PAGTCI, PAGTCO, COLCNV, MCO2SG,
     7                MO2CMI, MO2CMO, WWINP, HFGAS2, HFGAS7, HCERHX,
     8                TFGRHI, TFGRHO, MSIPT, MSLP1T, MCAFGI, SULEFF
C
      COMMON/USRHOC/  HOC
C
      COMMON/USRBASE/ DCCBV, DCCBS, DCCBM, DCSSB, DCHXBC, DCHXBM,
     1                DCHRB, DCFB81, DCFB82, DCFB83, DCIDB, AHXBC,
     2                AHXBM, GFGSSB, MFGHRB, GFGFB, RACB, NBB, MAB,
     3                GFGIDB
C
      COMMON/USRCST/ NCSTYR, CF, EREAL, FEHO, FGF, FICC, FPJ, INTRST,
     1               NOYEARS, IBKLIF, RTAX, ALABOR, NSHIFTS, EINFL, RRD,
```

```
     2                     RRP, RRE, DR, PR, TX, XITC, PTI, CEP
C
      COMMON/USRFPC/ FPCCH, FPCC, FPCSS, FPCHX, FPCBF, FPCGT, FPCHR,
     1               FPCST, FPCFGD, FPCFF, FPCID, FPCGF
C
      COMMON/ USRFMC/ FMCCH, FMCC, FMCSS, FMCHX, FMCBF, FMCGT, FMCHR,
     1                FMCST, FMCFGD, FMCFF, FMCID, FMCGF
C
      COMMON/USRERR/  ERRST
C
      COMMON/USRBC/   BCCOAL, BCRF, BCSA, BCSH, BCSP, BCHY, BCMO, BCLS,
     1                BCFB, BCFO, BCAA, BCWAT, BCLPG, BCSLG, BCSLD,
     2                BCASH, UCLAND
C
      COMMON/USRPER/  NOC, NTC, NOHX, NTHX, NOGT, NTGT, NOHR, NTHR,
     1                VFGCO, RTC, TRFC, RCAS, RAC, NB,
     2                GENEFF, EF, FRACAL, FRACAM, FRACAH, FRACSL,
     3                EFSSPL, EFSSPM, EFSSPH, EFHXPL, EFHXPM, EFHXPH,
     4                EFFFPL, EFFFPM, EFFFPH, NOX, FGDEFF, MOTEFF,
     5                HHVCOL
C
      COMMON/USRSF/   SFSS, SFHXC, SFHXM, SFID,
     1                F81, F82, F83, F84, A81, A82, A83, A84, B81,B82,
     2                B83, B84, C81, C82, C83, C84
C
      COMMON/USROUT/  HEATRT, EFFHHV, WGTE, WSTE, WAUXE, MWNETE,
     1                DCCH, DCC, DCGT, DCSS, DCHX, DCHR, DCST, DCBF,
     2                DCFGD, DCFF, DCID, DCGF, TDC, CICC, CTAX, CEHO,
     3                TIC, CPC, CPJ, TPC, AF, TPI, AFDC, PPC, IC,
     4                TCICC, TCLAND, TCR, OCL, OCM, OCAS, FOC, OCFUEL,
     5                OCCONS, OCSLG, OCASH, OCSLD, VOC, DPERKW, FOCN,
     6                VOCN, FUELN, VOCINC, CELEC,
     7                MCOALIN, MTOTWIN, MTOTAIN, FAEMS, MSSLO, BATOT,
     8                SLDG, SOEMSD, NOXEMS, CO2EMS, FCF, VCLF, NOPERS
C
      COMMON/WRITEO/  REPOP
C
      COMMON/USER/    RMISS, IMISS, NGBAL, IPASS, IRESTR, ICONVG,
     1                LMSG, LPMSG, KFLAG, NHSTRY, NRPT, NTRMNL
C
C
C-----------------------------------------------------------------------
C     Defining the value of NRPT because not being able to access via
C       the COMMON block
C
C
      NRPT = 14
C-----------------------------------------------------------------------
C  Defining factor for converting by 1000.0
C
      FACN = 1000.0
C-----------------------------------------------------------------------
C
C   Calculating the steam turbine output to account for power output
C   reduction due to FGD reheat steam consumption
C
      RAT     = WLP2T/MSLP2I
      ENTHR   = MSLP2I*(HST7 - HST6A)
```

```
      ENTHRA = ENTHR - HFGREH
      MSLP2A = ENTHRA/(HST7 - HST6A)
      WLP2TA = RAT*MSLP2A
C Estimating the work requirement for compression of steam injection flow
C
      WSTIC = WSTIC/MOTEFF
      STNET = GENEFF*(WHPT + WIPT + WLP1T + WLP2TA + WSTIC)
C
C   Estimating the equivalent power consumption due to FGD reheat
C
      REHMWE = -0.293071*1.0D-6*(GENEFF*(WLP2T - WLP2TA))
C-----------------------------------------------------------------
C  Estimating the work requirement of the gas turbine compressors
C
      WGTC = WGTC1 + WGTC2 + WGTC3
C
C---------------------------------------------------------------------
C  Estimating the heat input from the HRSG to the steam cycle
      ENHRSG = MFGFGI*HFGAS7 - MFGHRI*HFGAS2
C
C---------------------------------------------------------------------
C   Calculating the gas turbine and the steam turbine efficiencies
C
      GTEFF = (GTNET/(0.99*HCERHX))*100.0
      STEFF = (STNET/ENHRSG)*100.0
C---------------------------------------------------------------------
C   Converting units of work output from gas and steam turbine from
C   Btu/hr to MW, and calculating the gross electrical output of the
C   EFCC plant.
C
C
      WGTE = -0.293071*GTNET/1000000.0
C
      WSTE = -0.293071*STNET/1000000.0
C
      MWGRE = WGTE + WSTE
C
C   Converting the units of higher heating value of coal from J/Kg
C   to Btu/lb coal.
C
      HHV = 0.000430210*HOC
C
C   Rate of heat input, MMBtu/hr
      HTINP = (HHV*MCFCI)/1000000.0
C---------------------------------------------------------------------
C   Caculating the excess air ratio, auxiliary air ratio, combustor heat
C   percentage, and combustor heat directed to the water wall as a percent
C   of the heat of reation
C
      EXARAT = 1 + MO2IDO/(MO2CMI-MO2CMO)
      AUARAT = MAAPAI/MCFCI
      COMHLP = (COMBHL/(HHVCOL*MCFCI))*100.0
      CHLWWP = (CWWHI /(HHVCOL*MCFCI))*100.0
C---------------------------------------------------------------------
C   Calculating the total air leakage from the ceramic heat exchanger tubes,
C   the CerHx heat loss percentage, and mass flow of steam injection in
C   lb/sec
C
```

```
      FALEAKP =(FCAHXL + FHAHXL)*100.0
      HXHLP   = (HXHL - 1.0)*100.0
C-----------------------------------------------------------------------
C  Estimating the gas turbine compressor pressure ratio
C
      GTPR = PAGTCO/PAGTCI
C
C  Calculating the CerHx effectiveness
C
      HXEFF = (TAHXO-TAHXI)/(TFGCO-TAHXI)
C-----------------------------------------------------------------------
C  Calculating the slurry flow andd sludge flow (solid and water) for FGD,
C  calcium to sulfur ratio, flue gas flow in Kacfm
C
      MSLFGI = MLFGI + MSSLI
      SLDG   = MSFGO + MWFGO
      CASR   = MCAFGI/(MSOFGI/64.0)
      FGKCFM = (MFGFGI/DFGFGI)/(1000.0*60.0)
C-----------------------------------------------------------------------
C   Input Variable Definitions:
C
C
C   MCFCI  = Mass flow of coal to combustor, lb/hr
C   MFGIDI = Mass flow of flue gas to the ID fan inlet, lb/hr
C   DELPID = Pressure of the flue gas stream to the ID fan inlet, psia
C   DFGID  = Average density of flue gas to the ID fan inlet, lb/ft3
C   MAAPAI = Mass flow rate of aux. air to the primary air fan inlet, lb/hr
C   DELPPA = Pressure of the air stream from the primary air fan outlet,
C            psia
C   DAAPA  = Average density of aux. air into the primary air fan inlet,
C            lb/ft3
C   WBFWP  = Work requirement of the boiler feedwater pump, Btu/hr
C   WFGDT  = Work requirement of the FGD unit, Btu/hr
C   GTNET  = Net work output of the gas turbine, Btu/hr
C   STNET  = Net work output of steam turbines, Btu/hr
C   HOC    = Heat of combustion of coal, J/Kg
C   MPWDAI = Mass flow fo polished water to deaerator inlet, lb/hr
C   WCONDP = Work required by condenser pump, Btu/hr
C   HHV    = Higher heating value of coal, J/Kg
C   WGTE   = Work output of gas turbine, Btu/hr
C   WSTE   = Work output of steam turbine, Btu/hr
C   MFGCO  = Mass flow of flue gas at combustor outlet, lb/hr
C   DFGCO  = Density of flue gas at combustor outlet, lb/ft3
C   MMUCWI = Mass flow of makeup water at condenser wall outlet, lb/hr
C   MSCCWI = Mass flow of steam condensate at condenser wall outlet, lb/hr
C   MFGFGI = Mass flow of flue gas at FGD inlet, lb/hr
C   DFGFGI = Density of flue gas at FGD inlet, lb/hr
C   MSOFGI = Mass flow of SO2 at FGD inlet, lb/hr
C   MASLO  = Mass flow of ash at slagger outlet
C   MLFGI  = Mass flow of limestone slurry water at FGD inlet, lb/hr
C   DLFGI  = Density of limestone slurry at FGD inlet, lb/ft3
C   MSFGO  = Mass flow of sludge at FGD outlet, lb/hr
C   MFGHRI = Mass flow of flue gas at HRSG inlet, lb/hr
C   WHPT   = Work output of high pressure steam turbine, Btu/hr
C   WIPT   = Work output of intermediate pressure steam turbine, Btu/hr
C   WLP1T  = Work output of first stage low pressure steam turbine, Btu/hr
C   WLP2T  = Work output of second stage low pressure steam turbine, Btu/hr
C   MSLP2I = Mass flow of steam at second stage low pressure
```

```
C           steam turbine inlet, lb/hr
C   HST6A  = Enthalpy of steam at second stage low pressure
C            steam turbine inlet, Btu/lb
C   HFGREH = Enthalpy (heat) requirement for FGD reheat, Btu/hr
C   HST7   = Enthalpy of steam at second stage low pressure
C            steam turbine outlet, Btu/hr
C   MSOIDO = Moleflow of SO2 at ID outlet (stack), lbmol/hr
C   MO2IDO = Moleflow of O2 at ID outlet (stack), lbmol/hr
C   MSTIDO = Moleflow of H2O (steam) at ID outlet (stack), lbmol/hr
C   MFGIDO = Moleflow of fluegas at ID outlet (stack), lbmol/hr
C   NOX    = NOx in fluegas at stack outlet, lb/MMBtu
C   MASSFG = Massflow of fluegas at stack outlet, lb/hr
C   MFGFFI = Massflow of fluegas at FF inlet, lb/hr
C   DFGFFI = Density of fluegas at FF inlet, lb/ft3
C
C-------------------------------------------------------------------
C
C  The following statement is used to avoid printing the report file for
C  analyisis cases
C
      IF(REPOP.EQ.'NO') GOTO 710
C
C--------------------------------------------------------------------
C
C   Print out the input performance parameters accessed from the
C   performance simulation model (ASPEN input file)
C
C
      WRITE(NRPT,100)
 100  FORMAT('       EXTERNALLY-FIRED COMBINED CYCLE POWER PLANT',//,
     &      '                  " Version 1.0"',//,
     &      '  ***  SUMMARY OF SAMPLED ASPEN FLOWSHEET ',
     & 'PERFORMANCE PARAMETERS   ***',//,
     & 'PLANT SECTION    ','FLOWSHEET PERFORMANCE PARAMETERS',
     &  10X, 'VALUE', 2X, 'UNITS',/,
     & '-------------    ','--------------------------------',
     &  10X, '-----', 2X, '-----',/)
C
      WRITE(NRPT, 110)COMHLP, MWCWW, CHLWWP, COLCNV*100.0, AUARAT,
     &                FALEAKP, HXHLP, HXEFF, GTIT, GTPR,
     &                MSTI, MWGTEI, AMBT, TSTG
 110  FORMAT(/,
     & 'General          ',
     & 'Combustor heat loss                   '    , F12.2,' %',/,
     & 'Process',9X,'Mass flow of water to comb. ',
     &               'water wall',            2X, F8.2,' lb/hr',/,
     & 'Parameters',6X,'Comb. heat directed to water walls',
     &                                        2X, F12.2,' %',/,
     & 16X,'Carbon conversion of coal          ', F12.2,' %    ',/,
     & 16X,'Auxiliary air flow per lb of coal ', F14.2,' lb/lb coal',/,
     & 16X,'Air leakage in CerHx               ', F12.2,' %',/,
     & 16X,'CerHx heat loss                    ', F12.2,' %',/,
     & 16X,'CerHx  Effectiveness               ', F12.2,     /,
     & 16X,'Gas turbine inlet temperature      ', F12.2,' deg F',/,
     & 16X,'GT compressor pressure ratio       ', F12.2,         /,
     & 16X,'Steam injection mass flow          ', F12.2,' lb/hr',/,
     & 16X,'Water injection mass flow          ', F12.2,' lb/hr',/,
     & 16X,'Ambient temperature                ', F12.2,' deg F',/,
```

```
     & 16X,'Stack gas temperature                    ', F12.2,' deg F',//)
C
      WRITE(NRPT, 120) MCFCI, HTINP, MAAPAI,
     &                 DELPPA, DAAPA, MFGCO, EXARAT, DFGCO, TFGCO,
     &                 COMBHL, CWWHI
 120  FORMAT(/,
     & 'Coal Feed       ',
     & 'Mass flow of coal to combustor       '    , F12.2,' lb/hr',/,
     & 16X,'Plant thermal input                  ', E12.4,' MMBtu/hr',/,
     & 16X,'Auxiliary air mass flow rate         ', F12.2,' lb/hr',/,
     & 16X,'Pressure of aux.air at PA fan outlet', F12.2,' psia ',/,
     & 16X,'Avg. den. of aux.air at PA fan outlet',F11.4,' lb/ft3',//,
C
     & 'Coal Combustor  ',
     & 'Massflow of fluegas at combustor    ',     F12.2,' lb/hr' ,/,
     & 16X,'Excess air ratio                     ', F12.2,          /,
     & 16X,'Density of fluegas at combustor      ', F12.4,' lb/ft3',/,
     & 16X,'Temp. of fluegas at combustor outlet', F12.2,' deg F' ,/,
     & 16X,'Combustor heat loss                  ', E12.6,' Btu/hr',/,
     & 16X,'Combustor heat to water wall         ', E12.6,' Btu/hr',/)
C
      WRITE(NRPT, 130) GTNET, WGTC, TAHXI, TAHXO, MAGTI, MAC3O,
     &                 MAT1I, MAGTO, HCERHX, GTEFF
 130  FORMAT(/,
     & 'Gas Turbine     ',
     & 'Net GT power output                 ',     E12.6,' Btu/hr',/,
     & 16X,'GT compressor work requirement      ', E12.6,' Btu/hr',/,
     & 16X,'Temperature of air at CerHx inlet   ', F12.2,' deg F',/,
     & 16X,'Temperature of air at CerHx outlet  ', F12.2,' deg F',/,
     & 16X,'Massflow of air to GT comp inlet    ', F12.2,' lb/hr',/,
     & 16X,'Massflow of air at GT comp outlet   ', F12.2,' lb/hr',/,
     & 16X,'Massflow of air at GT turb inlet    ', F12.2,' lb/hr',/,
     & 16X,'Massflow of air at GT turb outlet   ', F12.2,' lb/hr',/,
     & 16X,'Heat duty of the CerHx              ', E12.6,' Btu/hr',/,
     & 16X,'Net Efficiency of the Gas Turbine   ', F12.2,' %',/)
C
      WRITE(NRPT, 140)MFGHRI, TFGHRI, ENHRSG,
     &                MSTSTI, MSIPT, MSLP1T, MSLP2I,
     &                MMUCWI, MSCCWI, MPWDAI,
     &                WCONDP, WBFWP, WHPT, WIPT, WLP1T, WLP2T,
     &                HST6A, HFGREH, REHMWE, HST7, WSTIC, WWINP,
     &                STNET, STEFF
 140  FORMAT(/,
     & 'HRSG            ',
     & 'Massflow of flue gas at HRSG inlet  ',     F12.2,' lb/hr',/,
     & 16X,'Temperature of flue gas at HRSG inlet',F11.2,' deg F',/,
     & 16X,'Enthalpy of HRSG to steam cycle     ', E12.6,' Btu/hr',//,
C
     & 'Steam Cycle     ',
     & 'Mass flow of steam to HP steam turb ',     F12.2,' lb/hr',/,
     & 16X,'Mass flow of steam to IP steam turb ', F12.2,' Btu/hr',/,
     & 16X,'Mass flow of steam to LP1 steam turb', F12.2,' Btu/hr',/,
     & 16X,'Mass flow of steam to LP2 steam turb', F12.2,' Btu/hr',/,
     & 16X,'Massflow of makeup water            ', F12.2,' lb/hr',/,
     & 16X,'Massflow of steam condensate        ', F12.2,' lb/hr',/,
     & 16X,'Massflow of polished water          ', F12.2,' lb/hr',/,
     & 16X,'Work req. of condenser pump         ', E12.6,' Btu/hr',/,
     & 16X,'Work req. of boiler feed water pump ', E12.6,' Btu/hr',/,
```

```
     & 16X,'Work output of HP steam turbine     ', E12.6,' Btu/hr',/,
     & 16X,'Work output of IP steam turbine     ', E12.6,' Btu/hr',/,
     & 16X,'Work output of LP1 steam turbine    ', E12.6,' Btu/hr',/,
     & 16X,'Work of output LP2 steam turbine    ', E12.6,' Btu/hr',/,
     & 16X,'Enth. of inlet stream to LP2 turb   ', F12.2,' Btu/hr',/,
     & 16X,'Heat requirement for FGD reheat     ', E12.6,' Btu/hr',/,
     & 16X,'Heat requirement for FGD reheat     ', F12.2,' MW    ',/,
     & 16X,'Enth. of exit stream of LP2 turb    ', F12.2,' Btu/hr',/,
     & 16X,'Work req. of steam inj. compressor  ', E12.6,' Btu/hr',/,
     & 16X,'Work req. of water inj. pump        ', E12.6,' Btu/hr',/,
     & 16X,'Net ST power output                 ', E12.6,' Btu/hr',/,
     & 16X,'Net Efficiency of the Steam Turbine ', F12.2,' %',/)
C
      WRITE(NRPT, 145) SULEFF, CASR, MFGFGI, MSOFGI, MSLFGI, RWFGD,
     &                 SLDG, DFGFGI, DLFGI
 145  FORMAT(/,
     & 'FGD Parameters  ',
     & 'Sulfur removal efficiency              ',     F12.2,/,
     & 16X,'Calcium to sulfur ratio             ', F12.2,' Ca/S',/,
     & 16X,'Massflow of flue gas at FGD inlet   ', F12.2,' lb/hr',/,
     & 16X,'Massflow of SO2 at FGD inlet        ', F12.2,' lb/hr',/,
     & 16X,'Massflow of slurry to FGD           ', F12.2,' lb/hr',/,
     & 16X,'Massflow of water recir. to FGD     ', F12.2,' lb/hr',/,
     & 16X,'Massflow of sludge from FGD (wet)   ', F12.2,' lb/hr',/,
     & 16X,'Density of flue gas at FGD inlet    ', F12.4,' lb/ft3',/,
     & 16X,'Density of slurry to FGD            ', F12.4,' lb/ft3',/)
C
      WRITE(NRPT, 150)MASLO, MFGFFI,
     &               MFGIDI, MASSFG, MCO2SG, MSOIDO, NOX, MO2IDO,
     &               MSTIDO, MFGIDO,  DELPID, TFGRHI, TFGRHO,
     &               DFGFFI, DFGID
 150  FORMAT(/,
     & 'Environmental   ',
     & 'Massflow of ash at slagger             ',     F12.2,' lb/hr',/,
     & 16X,'Massflow of flue gas at FF inlet    ', F12.2,' lb/hr',/,
     & 16X,'Massflow of flue gas to ID fan      ', F12.2,' lb/hr',/,
     & 16X,'Massflow of flue gas exiting stack  ', F12.2,' lb/hr',/,
     & 16X,'Massflow of CO2 in stack gas        ', F12.2,' lb/hr',/,
     & 16X,'Moleflow of SO2 in stack gas        ', F12.2,' lbmol/hr',/,
     & 16X,'NOx in stack gas                    ', F12.2,' lb/MMBtu',/,
     & 16X,'Moleflow of O2 in stack gas         ', F12.2,' lbmol/hr',/,
     & 16X,'Moleflow of H2O in stack gas        ', F12.2,' lbmol/hr',/,
     & 16X,'Moleflow of fluegas exiting stack   ', F12.2,' lbmol/hr',/,
     & 16X,'Pressure at ID fan inlet            ', F12.2,' psia ',/,
     & 16X,'Temp. of flue gas at reheater inlet ', F12.2,' deg F',/,
     & 16X,'Temp. of flue gas at reheater outlet', F12.2,' deg F',/,
     & 16X,'Density of flue gas at FF inlet     ', F12.4,' lb/ft3',/,
     & 16X,'Average density of flue gas         ', F12.4,' lb/ft3',/)
C
      WRITE(NRPT, 160)HOC
 160  FORMAT(/,
     & 'Miscellaneous   ',
     & 'Heat of Combustion of coal             ',     F12.2,' J/Kg',/)
C
      WRITE(NRPT,170)FILTPD, STMPD, HXCPD, TMX1PD, TMX2PD, TMX3PD,
     &              TMX4PD, AIRPD, CMIXPD, COMBPD, SSPD, HALKPD,
     &              HXHPD, CALKPD, REHHPD, SUPHPD, HEVHPD, HECHPD,
     &              LEVHPD, FGDMPD, FGDRPD, FABPD
```

```
 170  FORMAT(/,
     &  'PROCESS SECTION                           ','PRESSURE DROP (psi)',/,
     &  '---------------                           ','-------------------',/,
     &  'Air filter                                ', F10.4,/,
     &  'Steam injection mixer                     ', F10.4,/,
     &  'CerHx-air side                            ', F10.4,/,
     &  'Gas turbine Mixer 1                       ', F10.4,/,
     &  'Gas turbine Mixer 2                       ', F10.4,/,
     &  'Gas turbine Mixer 3                       ', F10.4,/,
     &  'Gas turbine Mixer 4                       ', F10.4,/,
     &  'Combustor air blower                      ', F10.4,/,
     &  'Combustor mixer                           ', F10.4,/,
     &  'Coal combustor                            ', F10.4,/,
     &  'Slag screen                               ', F10.4,/,
     &  'CerHx airleak-hot side                    ', F10.4,/,
     &  'CerHx-flue gas side                       ', F10.4,/,
     &  'CerHx airleak-cold side                   ', F10.4,/,
     &  'HRSG reheater                             ', F10.4,/,
     &  'HRSG superheater                          ', F10.4,/,
     &  'HRSG-High pressure evaporator             ', F10.4,/,
     &  'HRSG-high pressure economizer             ', F10.4,/,
     &  'HRSG-low pressure evaporator              ', F10.4,/,
     &  'FGD mixer                                 ', F10.4,/,
     &  'FGD reactor                               ', F10.4,/,
     &  'Fabric filter                             ', F10.4,//,
     &  'Note:  A negative number or 0 implies a pressure drop',/,
     &  '       whereas a positive number implies outlet pressure')
C
 710  CONTINUE
C
C-------------------COST MODEL PARAMETERS---------------------------
C
C     The following variables are cost model parameters.
C
C     Set the Chemical Engineering Plant Cost Index (CI) and Industrial
C     Chemicals Producer Price Indicator (CICPPI) to January of the cost
C     year NCSTYR.  The default is 1989 dollars unless otherwise specified.
C
C
C
C
      IF (NCSTYR.EQ.1994) THEN
         CI=368.0
      ELSEIF (NCSTYR.EQ.1993) THEN
         CI=359.2
      ELSEIF (NCSTYR.EQ.1992) THEN
         CI=358.2
      ELSEIF (NCSTYR.EQ.1991) THEN
         CI=361.3
      ELSEIF (NCSTYR.EQ.1990) THEN
         CI=354.7
         CICPPI=391.87
      ELSEIF (NCSTYR.EQ.1989) THEN
         CI=351.5
         CICPPI=411.25
      ELSEIF (NCSTYR.EQ.1988) THEN
         CI=336.3
         CICPPI=349.9
```

```
      ELSEIF (NCSTYR.EQ.1987) THEN
         CI=318.3
         CICPPI=323.9
      ELSEIF (NCSTYR.EQ.1986) THEN
         CI=323.5
         CICPPI=341.7
      ELSEIF (NCSTYR.EQ.1985) THEN
         CI=324.7
         CICPPI=337.7
      ELSEIF (NCSTYR.EQ.1984) THEN
         CI=320.3
         CICPPI=347.4
      ELSEIF (NCSTYR.EQ.1983) THEN
         CI=315.5
         CICPPI=339.9
      ELSE
         NCSTYR=1989
         CI=351.5
         CICPPI=411.25
      ENDIF
C
C
C-----------------------------------------------------------------------
C
C  Call financial subroutines to calculate the fixed charge and variable
C  cost levelization factors.
C
C  FCF   = Fixed charge factor (decimal)
C  DISC  = Discount rate (decimal)
C  VCLF  = Variable cost levelization factor (decimal)
C
      CALL USRFCF(EINFL, RRD, RRP, RRE, DR, PR, TX, XITC, PTI,
     1            IBKLIF, FCF, DISC)
      CALL USRVCLF(EINFL, EREAL, DISC, IBKLIF, VCLF)
C
C---------------------UNIT COST FOR FUEL, CHEMICALS, AND CATALYST---------
C
C   The following are unit costs for fuels, chemicals, and catalysts.
C   The costs are adjusted to other years using the Chemical Engineering
C   magazine Industrial Chemicals Producers Prices Index.
C
      UCCOAL = BCCOAL * (CICPPI/411.25)
      UCRF   = BCRF   * (CICPPI/411.25)
      UCSA   = BCSA   * (CICPPI/411.25)
      UCSH   = BCSH   * (CICPPI/411.25)
      UCSP   = BCSP   * (CICPPI/411.25)
      UCHY   = BCHY   * (CICPPI/411.25)
      UCMO   = BCMO   * (CICPPI/411.25)
      UCLS   = BCLS   * (CICPPI/411.25)
      UCFB   = BCFB   * (CICPPI/411.25)
      UCFO   = BCFO   * (CICPPI/411.25)
      UCAA   = BCAA   * (CICPPI/411.25)
      UCWAT  = BCWAT  * (CICPPI/411.25)
      UCLPG  = BCLPG  * (CICPPI/411.25)
      UCSLG  = BCSLG  * (CICPPI/411.25)
      UCSLD  = BCSLD  * (CICPPI/411.25)
      UCASH  = BCASH  * (CICPPI/411.25)
C-----------------------------------------------------------------------
```

```
C
C---------------------DIRECT CAPITAL COST--------------------------------
C
C     The direct capital cost for the following process areas are
C     estimated:
C
C     1.  Coal Handling.
C       2.  Coal Combustor
C       3.  Slag Screen
C     4.  Ceramic Heat Exchanger (CerHx).
C     5.  Boiler Feedwater System.
C     6.  Heat Recovery Steam Generator (HRSG).
C     7.  Steam Turbine.
C     8.  Gas Turbine
C     9.  Flue Gas Desulfurization (FGD) unit
C     10. Fabric Filter (FF).
C     11. ID Fan.
C     12. General Facilites.
C
C----------------------------------------------------------------------
C  Direct Cost Calculation for Coal Handling.  Coal Feed rate
C  is in lb/hr. Original regression based on tons/day. Coefficient
C  adjusted accordingly.
C
C
C
      CALL USRWCK (4.16D4,MCFCI,3.0D5,'MCFCI','DCCH')
C
C  The following relationship is based on coal handling of a PCFS
C  plant
C
      NOCH = 1
      NTCH = 1
      DCCH = 3.3163*(0.012*MCFCI)*NTCH*(CI/320.3)
C
C---------------------------------------------------------------------
C
C   Direct Capital Cost of Coal Combustor.
C
      NOC = 1
      NTC = 1
C
      GFGCO = MFGCO/DFGCO
C
      IRC = SQRT(GFGCO/(3600.0*3.414*NOC*VFGCO))
C
      IRVC = IRC + TRFC
C
      HC = VFGCO*RTC
C
C
      SAC = 2.0*3.414*(IRVC**2.0) + 2.0*3.414*IRVC*HC
C
C
      DCCV = DCCBV*(SAC/520.0)*NTC*(CI/351.5)
C
C
      DCCR = 2.0*3.414*IRVC*HC*UCRF*NTC*(CI/351.5)
```

```
C
C
      DCCS = DCCBS*NTC*((SAC/520.0)**0.6)*(CI/351.5)
C
C
      DCCM = DCCBM*NTC*((SAC/520.0)**0.6)*(CI/351.5)
C
C   Direct capital cost of the coal combustor is the sum of direct cost
C   of the different components.
C
      DCC = (DCCV + (DCCR/1000.0) + DCCS + DCCM) * NTC
C
C-------------------------------------------------------------------
C   Direct Capital Cost of Gas Turbine.  The gross electrical
C   output is estimated in MW.
C
C   The number of gas turbines is estimated based on the air flow rate
C   to the gas turbine
C
      NOGT = INT(MAGTI/3000000.0+0.5)
      IF(NOGT.EQ.0) NOGT=1
      NTGT=NOGT
C
      DCGT = 28300.0*NTGT*(CI/351.5)
C
C-------------------------------------------------------------------
C   Direct Capital Cost of Slag Screen.
C
      NOSS = NOGT
      NTSS = NTGT
      DCSS = DCSSB*((GFGCO/GFGSSB)**SFSS)*NTSS*(CI/351.5)
C
C-------------------------------------------------------------------
C
C   Direct Capital cost of Ceramic Heat Exchanger.
C
C   Surface areas of heat exchanger tubes are calculated.
C
      AHXC = 1.0
      AHXM = 1.0
C
      DCHXC = DCHXBC*((AHXC/AHXBC)**SFHXC)*(CI/351.5)
C
      DCHXM = DCHXBM*((AHXM/AHXBM)**SFHXM)*(CI/351.5)
C
C   Total Direct Cost of Ceramic heat exchanger is the sum of the direct
C   cost of the ceramic and the metallic part.
C
      NOHX = NOGT
      NTHX = NTGT
      DCHX = (DCHXC + DCHXM)*NTHX*(CI/351.5)
C
C-------------------------------------------------------------------
C---------------------------------------------------------------------
C
C  Direct Capital Cost of HRSG.
C
      NOHR=NOGT
```

```
      NTHR=NTGT
C
      CALL USRWCK (66.0D3, (MFGHRI/NOHR), 6.0D6, 'MFGHRI', 'DCHR')
      DCHR = DCHRB*(MFGHRI/(NOHR*MFGHRB))*NTHR*(CI/368.0)
C
C-----------------------------------------------------------------------
C
C  Direct Capital Cost of Steam Turbine.
C
C  The gross electrical output of the Steam Turbine is estimated in MW.
C
      NOST = 1
      NTST = 1
C
C
      DCST = (158.7*(WSTE) + ERRST)*NTST*(CI/351.5)
C
C-----------------------------------------------------------------------
C
C   Direct cost of boiler feed water system.
C
C   Raw and polished water flow rate are calculated
C
      MPW = MPWDAI
      MRW = MMUCWI + MWGTEI
C
      NOBF = NOHR
      NTBF = NTHR
      CALL USRWCK(24.0D3, MRW, 0.614D6, 'MRW', 'DCBF')
      CALL USRWCK(2.34D5, MPW, 3.88D6,  'MPW', 'DCBF')
      DCBF = (0.145*(MRW**0.307)*(MPW**0.435)*NTHR*(CI/351.5))
C
C-----------------------------------------------------------------------
C
C   Direct cost of FGD.
C
      GFGFGI = MFGFGI/(1000.0*DFGFGI*60.0)
C
C
      YSO2 = (1000000.0*MSOFGI)/MFGFGI
C
C
      RLG = (0.00025*(YSO2 - 2000.0) - 0.725 - (10.0*RCAS - 10.3) -
     1         LOG10(1.0-FGDEFF))/0.0175
C
C
      C71 = 3.05-(0.79*LOG10(GFGFGI))+(0.22*LOG10(YSO2))+
     1       (0.137*LOG10(RLG))+(1.664*LOG10(RCAS))
      DC71 = 10.0**C71
C
      C72 = 2.67-(0.44*LOG10(GFGFGI))+(0.019*LOG10(YSO2))+
     1       (0.137*LOG10(RLG))+(1.372*LOG10(RCAS))
      DC72 = 10.0**C72
C
      C73 = 2.422-(0.398*LOG10(GFGFGI))+(1.372*LOG10(RCAS))
      DC73 = 10.0**C73
C
      C74 = 1.32-(0.619*LOG10(GFGFGI))+(0.379*LOG10(YSO2))+
```

```
     1       (1.655*LOG10(RCAS))
      DC74 = 10.0**C74
C
      C75 = 2.717-(0.899*LOG10(GFGFGI))+(1.433*LOG10(RCAS))+
     1       (0.022*LOG10(YSO2))
      DC75 = 10.0**C75
C
      C76 = 2.523-(0.697*LOG10(GFGFGI))+(0.016*LOG10(YSO2))-
     1       (0.012*LOG10(RLG))+(0.958*LOG10(RCAS))
C
C
C
      NOFGD = 1
      NTFGD = 1
      DCFGD = (DC71+DC72+DC73+DC74+DC75+DC76)*NTFGD*MWGRE
C
C-----------------------------------------------------------------
C
C   Direct cost of fabric filter.
C
C
      GFGFFI = MFGIDI/(60.0*DFGID)
C
      MAFFO = (0.65*MASLO)/2000.0
C
C
      DCFF81 = DCFB81*((GFGFFI/GFGFB)**F81)*((RACB/RAC)**A81)*
     1         ((NBB/NB)**B81)*((MAFFO/MAB)**C81)
C
      DCFF82 = DCFB82*((GFGFFI/GFGFB)**F82)*((RACB/RAC)**A82)*
     1         ((NBB/NB)**B82)*((MAFFO/MAB)**C82)
C
      DCFF83 = DCFB83*((GFGFFI/GFGFB)**F83)*((RACB/RAC)**A83)*
     1         ((NBB/NB)**B83)*((MAFFO/MAB)**C83)
C
      NOFF = 1
      NTFF = 1
      DCFF = (DCFF81 + DCFF82 + DCFF83)*NTFF*(CI/357.6)
C
C-----------------------------------------------------------------
C
C   Direct cost of ID fan.
C
      GFGIDI = MFGIDI/DFGID
C
      NOID = NOFF
      NTID = NTFF
      DCID = DCIDB*(GFGIDI/GFGIDB)**SFID*NTID
C
C-----------------------------------------------------------------
C
C--------------------CAPITAL COSTS: TOTAL PLANT INVESTMENT------------
C
C   The total plant investment is estimated in this section.  The total
C   capital requirement, which includes items dependent on operating
C   costs, is not calculated until after the operating costs are
C   estimated.
C
```

```
C   Total Direct Cost (TDC)
C
      TDC = DCCH+DCC+DCSS+DCHX+DCGT+DCHR+DCST+DCBF+DCFGD+DCFF+DCID
C
C   Total Process Facility Cost
C
      CICC = FICC*TDC
      CTAX = RTAX*(0.8*TDC + 0.1*CICC)
      TPFC = TDC + CICC + CTAX
C  Engineering and home office cost
      CEHO = FEHO*TPFC
C  General Facilities Cost
      CGF  = FGF*TPFC
C
C   Process Contingency Costs
C
      CPCCH  = FPCCH *DCCH* (1.0 + (CICC+CTAX)/TDC)
      CPCC   = FPCC  *DCC*  (1.0 + (CICC+CTAX)/TDC)
      CPCSS  = FPCSS *DCSS* (1.0 + (CICC+CTAX)/TDC)
      CPCHX  = FPCHX *DCHX* (1.0 + (CICC+CTAX)/TDC)
      CPCBF  = FPCBF *DCBF* (1.0 + (CICC+CTAX)/TDC)
      CPCGT  = FPCGT *DCGT* (1.0 + (CICC+CTAX)/TDC)
      CPCHR  = FPCHR *DCHR* (1.0 + (CICC+CTAX)/TDC)
      CPCST  = FPCST *DCST* (1.0 + (CICC+CTAX)/TDC)
      CPCFGD = FPCFGD*DCFGD*(1.0 + (CICC+CTAX)/TDC)
      CPCFF  = FPCFF *DCFF* (1.0 + (CICC+CTAX)/TDC)
      CPCID  = FPCID *DCID* (1.0 + (CICC+CTAX)/TDC)
      CPCGF  = FPCGF *DCGF* (1.0 + (CICC+CTAX)/TDC)
      CPC    = CPCCH+CPCC+CPCSS+CPCHX+CPCBF+CPCGT+CPCHR+CPCST+CPCFGD+
     1         CPCFF+CPCID+CPCGF
C
C   Project Contingency Costs
C
      CPJ = FPJ*TPFC
C
C   Total Plant Cost (TPC)
C
      TPC = TPFC + CGF + CEHO + CPC + CPJ + CEP
C
C   Total Plant Investment (TPI)
C
      Z = (1+INTRST)/(1+EREAL)
      IF(Z.LE.1) THEN
        AF = 1
      ELSE
        AF = (Z**NOYEARS - 1)/(NOYEARS*(Z-1))
      ENDIF
      TPI = AF*TPC
      AFDC = TPI-TPC
C
C---------------------NET PLANT ELECTRICAL OUTPUT----------------------
C
C     The net electrical output must be determined in order to estimate
C     a number of plant costs.  The gross power plant electrical output
C     is reduced by plant auxiliary loads, including coal handling, the
C     primary air fan, the ID fan, the FGD unit, and power consumption
C     in the steam cycle.  The pump and compressor power consumption is
C     estimated directly from the ASPEN simulation.
```

```
C
C
C     Auxiliary power requirement of the coal handling unit.  The factor
C     0.012 converts coal mass flow rate from lb/hr to tons/day
C
      WCHE = 0.39*0.012*MCFCI
C
C     Auxiliary power requirement of the boiler feedwater treatment
C     section in the steam cycle
C
      WBFE = 20.8 + 0.000213*MPWDAI
C
C     Electricity requirement of the ID fan in watts, converted directly
C     to kW by multiplying by 0.001
C
      WIDE = 0.001*0.0542*((MFGIDI*(14.7-DELPID))/(DFGID*EF))
C
C     Electricity requirement of the primary air fan in watts, converted
C     directly to kW by multiplying by 0.001
C
      WPAE = 0.001*0.0542*((MAAPAI*(DELPPA-14.7))/(DAAPA*EF))
C
C     Electricity requirement of the steam cycle is the sum of the
C     electricity requirment of the condenser pump, the boiler feedwater
C     pump, and the water injecton pump in kW
C
      WCONPE = 0.001*0.293071*WCONDP
      WBFWPE = 0.001*0.293071*WBFWP
      WWINPE = 0.001*0.293071*WWINP
      WSCE   = WCONPE + WBFWPE + WWINPE
C
C     Electricity requirement of the FGD unit in kW
C
      WFGD71 = 0.00234*(MLFGI)
      WFGD72 = 0.865*YSO2
      WFGD74 = 0.0023*SLDG
      WFGD75 = 0.006*(WFGD71+WFGD72+WFGD74)
C
      WFGDE = WFGD71+WFGD72+WFGD74+WFGD75
C
C
C     Electricity requirement for general facilities
C
      WGFE = 0.093*(WCHE+WBFE+WIDE+WPAE+WSCE+WFGDE)
C
C     Total auxiliary power requirement, converted to MW
C
      WAUXE = 0.001*(WCHE+WBFE+WIDE+WPAE+WSCE+WFGDE+WGFE)
C
C     Net plant power output
C
      MWNETE = WGTE + WSTE - WAUXE
      KWNETE  = MWNETE*1000.0
C
C-----------------------------------------------------------------------
C-----------------------INITIAL CATALYSTS AND CHEMICALS-----------------
C
      FOI = 80000.0/(MWNETE/550.0)
```

```
      TCICC = (UCFO*FOI)/1000.0
C-----------------------------------------------------------------------
C----------------------FIXED OPERATING COST-----------------------------
C
C   Fixed Operating Costs:  Operating Labor
C
      NOPERS = 18 + 4*NOGT
      OCL    = ALABOR*2080*NSHIFTS*NOPERS
C
C   Fixed Operating Costs: Maintenance, multiplying by 1000 to get the
C   cost in $/year
C
      OCMCH = FMCCH * DCCH * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMC  = FMCC  * DCC  * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMSS = FMCSS * DCSS * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMHX = FMCHX * DCHX * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMGT = FMCGT * DCGT * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMHR = FMCHR * DCHR * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMST = FMCST * DCST * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMBF = FMCBF * DCBF * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMFGD= FMCFGD* DCFGD* (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMFF = FMCFF * DCFF * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMID = FMCID * DCID * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
      OCMGF = FMCGF * DCGF * (1.0 + (CICC+CTAX)/TDC)  * 1000.0
C Maintenance cost for combined cycle
      OCMCC = OCMGT + OCMHR + OCMST
C
      OCM   = OCMCH+OCMC+OCMSS+OCMHX+OCMCC+OCMBF+OCMFGD+OCMFF+OCMID+
     1        OCMGF
C
      OCMM = 0.60*OCM
      OCML = 0.40*OCM
C
C   Fixed Operating Costs: Administration and Supervision
C
      OCAS = 0.30*(OCL + OCML)
C
C Total fixed operating costs
C
      FOC = OCL+OCM+OCAS
C-----------------------------------------------------------------------
C----------------------VARIABLE OPERATING COSTS-------------------------
C
C   Fuel (coal) consumption.
C
      OCFUEL = CF*8760*HTINP*UCCOAL
C
C   Consumables: which include boiler feedwater, FGD, and fabric filter
C                consumables.
C
C
C   Boiler feedwater (raw and polished water) treatment chemical requirement
C
      MSABFI  = CF*(47.0  + 2.09D-3*MRW)
      MSHBFI  = CF*(9.50  + 4.2D-4*MRW)
      MSPBFI  = CF*(115.0 + 3.61D-3*MRW)
      MHYBFI  = CF*(529.0 + 0.0174*MRW)
      MMOBFI  = CF*(420.0 + 0.0163*MRW)
```

```
      MSABFPI = CF*(15.0  + 5.4D-5*MPW)
      MSHBFPI = CF*(30.0  + 1.07D-4*MPW)
C
      CALL USRWCK (1.2D6, MPW, 2.2D6, 'MPW', 'MSABFP')
C
C  FGD water requirement is the difference between the water coming in with
C  the slurry and the recirculated water.  The Limestone slurry feed has been
C  estimated earlier as the sum of the water and the solids in the slurry
feed.
C
      MFGDW = MLFGI-RWFGD
C
C  FF consumables requirement (area of filter bags)
C
      ABAG = 0.25*CF*(((MFGFFI/DFGFFI)/60.0)/RAC)
C
C  Other consumables:  which include gas turbine startup fuel oil, plant
C                      and instrument air adsorbents, raw water for steam
C                      cycle, and LPG for plant flare.
C
      MFOGTI = 48000.0*CF*(MWNETE/550.0)
      MAAGFI = 3600.0*CF*(MWNETE/550.0)
      MTOTW  = (MRW+MFGDW)*1.191545D-4
      MTWAT  = 8760.0*CF*MTOTW
      MLPG   = 7200.0*CF*(MWNETE/550.0)
C
C  Estimate of consumable operating costs
C
      MSA  = MSABFI + MSABFPI
      MSH  = MSHBFI + MSHBFPI
      OCSA = UCSA * MSA
      OCSH = UCSH * MSH
      OCSP = UCSP * MSPBFI
      OCHY = UCHY * MHYBFI
      OCMO = UCMO * MMOBFI
C
      OCLMST = UCLS * MSSLI
      OCFFB  = UCFB  * NB
C
      OCFO = UCFO * MFOGTI
      OCAA = UCAA * MAAGFI
      OCWAT= UCWAT* MTWAT
      OCLPG= UCLPG* MLPG
C
      OCCONS = OCSA+OCSH+OCSP+OCHY+OCMO+OCLMST+OCFFB+OCFO+OCAA+OCWAT
     1          +OCLPG
C
C   Costs for slag, sludge, and ash disposal.
C
C
      MSSLO = FRACSL*MASLO
      MFASLO = (1.0 - FRACSL)*MASLO
C
      FALSLO = FRACAL*MFASLO
      FAMSLO = FRACAM*MFASLO
      FAHSLO = FRACAH*MFASLO
C
C
```

```
      BALSSO = EFSSPL*FALSLO
      BAMSSO = EFSSPM*FAMSLO
      BAHSSO = EFSSPH*FAHSLO
C
      BATSSO = BALSSO + BAMSSO + BAHSSO
C
      FALSSO = (1.0 - EFSSPL)*FALSLO
      FAMSSO = (1.0 - EFSSPM)*FAMSLO
      FAHSSO = (1.0 - EFSSPH)*FAHSLO
C
      BALHXO = EFHXPL*FALSSO
      BAMHXO = EFHXPM*FAMSSO
      BAHHXO = EFHXPH*FAHSSO
C
C
      BATHXO = BALHXO + BAMHXO + BAHHXO
C
      FALHXO = (1.0 - EFHXPL)*FALSSO
      FAMHXO = (1.0 - EFHXPM)*FAMSSO
      FAHHXO = (1.0 - EFHXPH)*FAHSSO
C
      BALFFO = EFFFPL*FALHXO
      BAMFFO = EFFFPM*FAMHXO
      BAHFFO = EFFFPH*FAHHXO
C
      BATFFO = BALFFO + BAMFFO + BAHFFO
C
      FALFFO = (1.0 - EFFFPL)*FALHXO
      FAMFFO = (1.0 - EFFFPM)*FAMHXO
      FAHFFO = (1.0 - EFFFPH)*FAHHXO
C
      FATOT = FALFFO + FAMFFO + FAHFFO
      BATOT = BATSSO + BATHXO + BATFFO
C
      OCSLG = 8760.0*CF*UCSLG*(MSSLO/2000.0)
      OCASH = 8760.0*CF*UCASH*(BATOT/2000.0)
      OCSLD = 8760.0*CF*UCSLD*(SLDG/2000.0)
C
C    TOTAL VARIABLE OPERATING COST
C
      VOC = OCFUEL + OCCONS + OCSLG + OCASH + OCSLD
C
C---------------------------------------------------------------
C----------------------PREPRODUCTION CAPITAL COSTS--------------------
C
      PPFC   = FOC/(12000.0)
      PPOC   = (0.083/CF)*(OCCONS + OCSLG + OCASH + OCSLD)/1000.0
      PPFUEL = (0.021/CF)*(OCFUEL/1000.0)
      PPC    =  PPFC + PPOC + PPFUEL + 0.02*TPI
C
C----------------------INVENTORY CAPITAL COSTS------------------------
C
      IC = (0.164/CF)*((OCFUEL + OCCONS)/1000.0)
C---------------------------------------------------------------
C----------------------TOTAL CAPITAL REQUIREMENT----------------------
C
      ALAND  = -93.0 + 0.065*HTINP
      TCLAND = ALAND*UCLAND/1000.0
```

```
      TCR    = 1.01*TPI+PPC+IC+TCICC+TCLAND
C-----------------------------------------------------------------------
C-----------------------NORMALIZED COST AND COST OF ELECTRICITY---------
C
      DPERKW = TCR/MWNETE
      FOCN   = FOC/(MWNETE*1000.0)
      VOCN   = VOC/(MWNETE*8760.0*CF)
      FUELN  = OCFUEL/(MWNETE*8760.0*CF)
      VOCINC = VOCN - FUELN
      CELEC  = (1000.0*FCF*TCR + VCLF*(FOC+VOC))/(MWNETE*8760.0*CF)
C
C
C-----------------------PLANT EFFICIENCY--------------------------------
C
      HEATRT = (MCFCI*HHV)/KWNETE
      EFFHHV = (3414.98/HEATRT)*100.0
      EFFHHV1 = (3414980*MWNETE)/(MCFCI*HHV)*100.0
C
C-----------------------------------------------------------------------
C-----------------------ENVIRONMENTAL SUMMARY---------------------------
C
C Calculating the material inputs to the EFCC plant, which includes coal,
C air, and water
C
      MCOALIN = MCFCI/KWNETE
      MFGDWIN = MFGDW/KWNETE
      MRWIN   = MRW/KWNETE
      MTOTWIN = MFGDWIN + MRWIN
      MGTAIN  = MAGTI/KWNETE
      MAUXAIN = MAAPAI/KWNETE
      MTOTAIN = MGTAIN + MAUXAIN
C
C Massflow of flyash on 15% O2 and dry basis (ppm)
C
      FADRY = (FATOT/(MASSFG - (MSTIDO*18.0)))*1000000.0
      ACTO2 = (MO2IDO/MFGIDO)*100.0
      FADO2 = FADRY*((20.9-15.0)/(20.9 - ACTO2))
C
C Massflow of ash in lb/MMBtu
C
      FAEMS = FATOT/HTINP
C
C Emission of SO2 based on direct estimation of SO2 in stack gas from
C ASPEN simulation in lb/MMBtu and by assuming an FGD SO2 removal
C efficiency in lb/MMBtu
C
      SOEMSD = (MSOIDO*64.0)/HTINP
      SOEMSF = (1.0-FGDEFF)*MSOFGI/HTINP
C
C Emission of NO and NO2, assuming 95% NO on a molar basis (92.5% on a
C mass basis)
C
      NOEMS   = 0.925*NOX
      NO2EMS  = 0.075*NOX
      NOXEMS  = NOEMS + NO2EMS
C
C Emissions of carbon dioxide
C
```

```
      CO2EMS = MCO2SG/KWNETE
C-----------------------------------------------------------------------
C
C-----------------------COST SUMMARY REPORT----------------------------
C
      IF(REPOP.EQ.'NO') GOTO 720
C
      WRITE(NRPT,200)
 200  FORMAT(/,33X, 'COST SUMMARY',
     & /, 21X, 'Externally Fired Combined Cycle')
C
C------------------Report Cost Model Parameters---------------------
C
      WRITE(NRPT, 210)
 210  FORMAT(/,'A. KEY INPUT ASSUMPTIONS----------------------------',
     & '---------------')
C
C
      WRITE(NRPT, 220)VFGCO, RTC, TRFC, RCAS, RAC, NB, GENEFF, EF,
     &                MOTEFF,
     &                FGDEFF, FRACSL, FRACAL, FRACAM, FRACAH, EFSSPL,
     &                EFSSPM, EFSSPH, EFHXPL, EFHXPM, EFHXPH, EFFFPL,
     &                EFFFPM, EFFFPH
 220  FORMAT(/,'Performance Assumptions:',
     &//,' Superficial gas velocity in combustor: ', F6.3,' ft/sec'
     & /,'       Gas residence time in combustor: ', F6.3,' sec',
     & /,'         Combustor refractory thicknes: ', F6.3,' ft',
     & /,'       Calcium to sulfur ratio for FGD: ', F6.3,
     & /,'                    Air to cloth ratio: ', F6.3,' acfm/ft2',
     & /,'              No. of filter bags/comp.: ', F6.2,
     & /,'                  Generator efficiency: ', F6.3,
     & /,'             Centrigual fan efficiency: ', F6.3,
     & /,'             Electric motor efficiency: ', F6.3,
     & /,'            FGD SO2 removal efficiency: ', F6.3,
     & /,'                Frac. coal ash as slag: ', F6.3,
     & /,'     Frac. of flyash in low size range: ', F6.3,
     & /,'     Frac. of flyash in med size range: ', F6.3,
     & /,'    Frac. of flyash in high size range: ', F6.3,
     & /,'         Low size flyash rem eff in SS: ', F6.3,
     & /,'         Med size flyash rem eff in SS: ', F6.3,
     & /,'        High size flyash rem eff in SS: ', F6.3,
     & /,'         Low size flyash rem eff in HX: ', F6.3,
     & /,'         Med size flyash rem eff in HX: ', F6.3,
     & /,'        High size flyash rem eff in HX: ', F6.3,
     & /,'         Low size flyash rem eff in FF: ', F6.3,
     & /,'         Med size flyash rem eff in FF: ', F6.3,
     & /,'        High size flyash rem eff in FF: ', F6.3,
     & //,'Notes:  SS = Slag screen,   HX  = Ceramic heat exchanger',/,
     &    '        FF = Fabric filter, FGD = Flue gas desulfurization')
C
      WRITE(NRPT, 230)NCSTYR, EINFL, CI, EREAL, CICPPI, IBKLIF,
     &                CF, RTAX, FGF, RRD, FICC, RRP, FEHO, RRE, FPJ,
     &                DR, PR, ALABOR, TX, NSHIFTS, XITC, INTRST, PTI,
     &                NOYEARS, CEP
 230  FORMAT(/'Economic Assumptions:',
     & //,'                    Cost Year: January', I6,
     &    '                     Inflation Rate: ', F6.3,
     & /,'                   Plant Cost Index: ', F6.2,
```

```
     &           '        Real Escalation Rate: ', F6.3,
     &  /,'           Chemical Cost Index: ', F6.2,
     &           '              Plant Booklife: ', I6,
     &  /,'        Plant Capacity Factor: ', F6.3,
     &           '              Sales Tax Rate: ', F6.3,
     &  /,'    General Facilities Factor: ', F6.3,
     &           '         Real Return on Debt: ', F6.3,
     &  /,'  Indirect Construction Factor: ', F6.3,
     &           '          Real Ret. on Pref.: ', F6.3,
     &  /,'      Engr & Home Office Fees: ', F6.3,
     &           '         Real Ret. on Equity: ', F6.3,
     &  /,'    Project Contingency Factor: ', F6.3,
     &           '                  Debt Ratio: ', F6.3,
     &  /,'            Pref. Stock Ratio: ', F6.3,
     &           '         Avg. Op. Labor Rate: ', F6.3,
     &  /,'          Fed. & State Taxes: ', F6.3,
     &           '           Number of Shifts: ', F6.3,
     &  /,'       Investment Tax Credit: ', F6.3,
     &           '        Const. Interest Rate: ', F6.3,
     &  /,'      Prop. Taxes & Insurance: ', F6.3,
     &           '        Year of Construction: ', I6,
     &  /,'  Cost of Environmental Permit: ', F6.1,//)
C
C
      WRITE (NRPT, 240)
 240  FORMAT('Process Contingency and Maintenance Cost Factors:  ',
     &       //26X,'                   Process     Maintenance',
     &        /26X,'Plant Section    Contingency    Cost Factor',
     &        /26X,'-------------    -----------    -----------')
C
      WRITE (NRPT, 245)FPCCH, FMCCH, FPCC, FMCC, FPCSS, FMCSS, FPCHX,
     &                 FMCHX, FPCBF, FMCBF, FPCGT, FMCGT, FPCHR,
     &                 FMCHR, FPCST, FMCST, FPCFGD, FMCFGD, FPCFF,
     &                 FMCFF, FPCID, FMCID, FPCGF, FMCGF
 245  FORMAT(10X,'               Coal Handling ',6X,F6.3,10X,F6.3,
     &       /10X,'              Coal Combustor ',6X,F6.3,10X,F6.3,
     &       /10X,'                 Slag Screen ',6X,F6.3,10X,F6.3,
     &       /10X,'      Ceramic Heat Exchanger ',6X,F6.3,10X,F6.3,
     &       /10X,'            Boiler Feedwater ',6X,F6.3,10X,F6.3,
     &       /10X,'                 Gas Turbine ',6X,F6.3,10X,F6.3,
     &       /10X,'Heat Recovery Steam Generator ',6X,F6.3,10X,F6.3,
     &       /10X,'               Steam Turbine ',6X,F6.3,10X,F6.3,
     &       /10X,'    Flue Gas Desulfurization ',6X,F6.3,10X,F6.3,
     &       /10X,'               Fabric Filter ',6X,F6.3,10X,F6.3,
     &       /10X,'                      ID Fan ',6X,F6.3,10X,F6.3,
     &       /10X,'          General Facilities ',6X,F6.3,10X,F6.3)
C
C----------------------------CAPITAL COSTS---------------------------
C
      WRITE (NRPT, 305)
 305  FORMAT(/,'B. CALCULATED DIRECT CAPITAL AND PROCESS CONTINGENCY ',
     &  'COSTS ($1,000) ----',
     & //16X,'               Number of Units       Direct     Process',
     &  /16X,'Plant Section  Operating Total  Capital Cost Contingency',
     &  /16X,'-------------  ---------------  ------------ -----------')
C
      WRITE (NRPT, 310)NOCH, NTCH, DCCH, CPCCH, NOC,  NTC,  DCC,  CPCC,
     &                 NOSS, NTSS, DCSS, CPCSS, NOHX, NTHX, DCHX, CPCHX,
```

```
     &                  NOGT, NTGT, DCGT, CPCGT, NOHR, NTHR, DCHR, CPCHR,
     &                  NOST, NTST, DCST, CPCST, NOBF, NTBF, DCBF, CPCBF,
     &                  NOFGD, NTFGD, DCFGD, CPCFGD,
     &                  NOFF, NTFF, DCFF, CPCFF, NOID, NTID, DCID, CPCID,
     &                  DCGF, CPCGF
 310  FORMAT(
     & /,'                Coal Handling ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'               Coal Combustor ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'                  Slag Screen ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'      Ceramic Heat Exchanger ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'                  Gas Turbine ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'Heat Recovery Steam Generator ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'                Steam Turbine ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'             Boiler Feed Water ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'     Flue Gas Desulfurization ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'                Fabric Filter ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'                       ID Fan ',2X,I5,2X,I5,3X,F10.1,3X,F10.1,
     & /,'          General Facilities ',5X,'N/A',4X,'N/A',2X,
     &                                    F10.1,3X,F10.1,/)
C
      WRITE (NRPT, 320)TDC, CICC, CTAX, TPFC, CGF, CEHO, CEP, CPC, CPJ,
     &                 TPC, AFDC, TPI, PPC, IC, TCICC, TCLAND, TCR
 320  FORMAT(/,'C. CALCULATED TOTAL CAPITAL REQUIREMENET ($1,000)----',
     &         '--------------------',
     &//, 10X, '    Description                               ',
     &     7X, 'Capital Cost',
     & /, 10X, '    -----------                               ',
     &     7X, '------------',
     & /, 10X, '    Total Direct Cost                        ',10X,F10.0,
     & /, 10X, '        Indirect Construction Cost           ',10X,F10.0,
     & /, 10X, '        Sales Tax                            ',10X,F10.0,
     & /, 10X, '    Total Plant Facilities Cost              ',10X,F10.0,
     & /, 10X, '        General Facilities Cost              ',10X,F10.0,
     & /, 10X, '        Engineering & Home Office Fees       ',10X,F10.0,
     & /, 10X, '        Environmental Permitting             ',10X,F10.0,
     & /, 10X, '        Total Process Contingency            ',10X,F10.0,
     & /, 10X, '        Project Contingency                  ',10X,F10.0,
     & /, 10X, '    Total Plant Cost                         ',10X,F10.0,
     & /, 10X, '        AFDC                                 ',10X,F10.0,
     & /, 10X, '    Total Plant Investment                   ',10X,F10.0,
     & /, 10X, '        Preproduction (startup) cost         ',10X,F10.0,
     & /, 10X, '        Inventory Capital                    ',10X,F10.0,
     & /, 10X, '        Initial Catalysts and Chemicals      ',10X,F10.0,
     & /, 10X, '        Land                                 ',10X,F10.0,
     &//, 10X, 'TOTAL CAPITAL REQUIREMENT ($1,000)  ------------>',
     &     2X,F10.0)
C
C
C------------------------OPERATING COSTS-----------------------------
C
C---------------------FIXED OPERATING COSTS----------------------------
C
      WRITE (NRPT, 400) OCL, OCM, OCAS, FOC
 400  FORMAT(/'D. CALCULATED FIXED OPERATING COSTS ($/year)  -------',
     & '----------------',
     & //,10X, '    Description                   ', 15X, 'Annual Cost',
     &  /,10X, '    -----------                   ', 15X, '-----------',
     &  /,10X, '    Operating Labor                    ',10X, F11.0,
```

```
     &  /,10X,'    Maintenance Cost                      ',10X, F11.0,
     &  /,10X, '    Administration and Supervision    ',10X, F11.0,
     & //,10X, 'TOTAL FIXED OPERATING COST',
     &     '($/year) ------------>', F11.0)
C
C----------------------VARAIBLE OPERATING COSTS---------------------
C                                    and
C---------------Unit Costs for Fuel and Chemicals-------------------
C
      WRITE (NRPT, 450)
 450  FORMAT(/,
     &    'E. CALCULATED VARIABLE OPERATING COSTS  ---------',
     &    '---------------------',
     &   //,'   1. CONSUMABLES ($/year)',
     &   //,6X,'                Assumed         ',
     &    ' Calc. Material    Calc. Annual ',
     &    /,6X,'Description     Unit Cost       ',
     &    ' Requirement      Operating Cost',
     &    /,6X,'-----------     ---------       ',
     &     '-------------    -------------')
C
      WRITE (NRPT, 460) UCSA, MSA, OCSA, UCSH, MSH, OCSH,
     &                  UCSP, MSPBFI, OCSP, UCHY, MHYBFI, OCHY,
     &                  UCMO, MMOBFI, OCMO, UCLS, MSSLI, OCLMST,
     &                  UCFB, NB, OCFFB, UCFO, MFOGTI, OCFO,
     &                  UCAA, MAAGFI, OCAA, UCWAT, MTWAT, OCWAT,
     &                  UCLPG, MLPG, OCLPG, OCCONS
 460  FORMAT(
     & /,4X,' Sulfuric Acid: ',F7.2,' $/ton   ',F10.1,' ton/yr   '
     &  ,F14.0,
     & /,4X,'          NaOH: ',F7.2,' $/ton   ',F10.1,' ton/yr   '
     &  ,F14.0,
     & /,4X,'  Sodium Phos.: ',F7.2,' $/lb    ',F10.1,'  lb/yr   '
     &  ,F14.0,
     & /,4X,'     Hydrazine: ',F7.2,' $/lb    ',F10.1,'  lb/yr   '
     &  ,F14.0,
     & /,4X,'    Morpholine: ',F7.2,' $/lb    ',F10.1,'  lb/yr   '
     &  ,F14.0,
     & /,4X,'     Limestone: ',F7.2,' $/ton   ',F10.1,' ton/yr   '
     &  ,F14.0,
     & /,4X,'       FF bags: ',F7.2,' $/bag   ',F10.1,' bags     '
     &  ,F14.0,
     & /,4X,'      Fuel Oil: ',F7.2,' $/bbl   ',F10.1,' bbl/yr   '
     &  ,F14.0,
     & /,4X,'     Adsorbent: ',F7.2,' $/lb    ',F10.1,'  lb/yr   '
     &  ,F14.0,
     & /,4X,'         Water: ',F7.2,' $/1000gal ',F8.1,
     & ' (1000 gal)/yr' ,F10.0,
     & /,4X,'     LPG-Flare: ',F7.2,' $/ton   ',F10.1,' ton/yr   '
     &  ,F14.0,
     &//,4X, 'TOTAL CONSUMABLES ($/year) -------------',
     &       '----------->', F14.0)
C
      WRITE (NRPT, 470) UCCOAL, MCFCI, OCFUEL, UCSLG, MSSLO, OCSLG,
     &                  UCASH, BATOT, OCASH, UCSLD, SLDG, OCSLD, VOC
 470  FORMAT(
     &/,3X,' 2.  FUEL COST, and SLAG, ASH, and SLUDGE DISPOSAL',
     &            '($/year)',
```

```
     &//,6X,'           Coal: ',F7.2,' $/MMBtu ',F10.1,' lb/hr    '
     &  ,F14.0,
     & /,6X,'           Slag: ',F7.2,' $/ton   ',F10.1,' ton/yr   '
     &  ,F14.0,
     & /,6X,'            Ash: ',F7.2,' $/ton   ',F10.1,' ton/yr   '
     &  ,F14.0,
     & /,6X,'         Sludge: ',F7.2,' $/ton   ',F10.1,' ton/yr   '
     &  ,F14.0,
     & //,6X, 'TOTAL VARIABLE OPERATING COST ($/year) -------',
     &        '-------> ',F11.0)
C
C-----------------Normalized Costs and Cost of Electricity------------
C
      WRITE(NRPT, 500) WGTE, WCHE/FACN, WSCE/FACN, WSTE, WBFE/FACN,
     &                 WFGDE/FACN, WAUXE, WIDE/FACN, WGFE/FACN,
     &                 WPAE/FACN, MWNETE,
     &                 DPERKW, FOCN, VOCINC, FUELN,
     &                 VOCN, CELEC, FCF, VCLF
 500  FORMAT (/,
     &   'F.  CALCULATED COST OF ELECTRICITY-----------------------',
     &   '-----------------',
     & //,'      Power Summary (MWe)              ',
     &   'Auxiliary Electrical Loads (MWe)',
     & /,'-----------------------------',
     &   '     ------------------------------------------',
     & /,'Gas Turbine Output', 4X, F7.2, 5X, 'Coal Handling',4X,F4.2,
     &    2X, 'Steam Cycle',   3X, F4.2,
     & /,'Steam Turbine Output',2X, F7.2,5X,'Boilfeed Water',3X,F4.2,
     &    2X, 'Flue Gas Des.', 1X, F4.2,
     & /,'Total Auxiliary Loads',3X, F5.2, 5X, 'ID Fan', 11X,F4.2,
     &    2X, 'General Fac. ', 1X, F4.2,
     & /,'-----------------------------',
     &                                   5X,'Prim. AirFan',5X,F4.2,
     & /,'Net Electricity',     7X, F7.2, 5X,
     &//,40X, 'Capital Cost:', 3X, F9.2, ' $/kW',
     & /,32X, 'Fixed Operating Cost:',5X, F7.2, ' $/(kW-yr)',
     &//,8X, 'Incremental Variable Costs:', 4X, F6.2, ' mills/kWh',
     & /,8X, 'Fuel Cost:', 21X, F6.2, ' mills/kWh',
     &//,29X, 'Variable Operating Cost:',5X, F7.2, ' mills/kWh',
     &//,'COST OF ELECTRICITY  ------------------------------------>',
     &    F7.2, ' mills/kWh',
     & //, 'Fixed Charge Factor: ', 2X, F6.4, 5X
     &    'Variable Cost Levelization Factor: ', F6.4)
C
      WRITE(NRPT, 510) HEATRT, EFFHHV
 510  FORMAT (//, 'THE PLANT HEATRATE (HHV):', 8X, F7.0, 'BTU/KWh',
     &          /, 'THE PLANT EFFICIENCY (%,HHV):', 6X, F7.2,/)
C
C--------------------------Environmental Summary----------------------
C
      WRITE(NRPT, 520)MCOALIN, MFGDWIN, MRWIN, MTOTWIN, MGTAIN,
     &                MAUXAIN, MTOTAIN,
     &                FADO2, FAEMS, SOEMSD,
     &                NOXEMS, CO2EMS, MSSLO, BATOT, SLDG
 520  FORMAT(/'G. ENVIRONMENTAL SUMMARY -----------------------------',
     & '--------------------',//
     &  ,15X, 'INPUTS:', 5X, 'Coal',             25X, F10.4, ' lb/kWh',
     & /,            27X, 'FGD slurry water',13X, F10.4, ' lb/kWh',
```

```
     & /,            27X, 'Raw water',        20X, F10.4, ' lb/kWh',
     & /,            27X, 'Total water',      18X, F10.4, ' lb/kWh',
     & /,            27X, 'Gas Turbine air', 14X, F10.4, ' lb/kWh',
     & /,            27X, 'Auxiliary air',   16X, F10.4, ' lb/kWh',
     & /,            27X, 'Total air',        20X, F10.4, ' lb/kWh',
     & //,15X,'OUTPUTS:',4X, 'Flyash(dry,15%O2)',13X, F10.4, ' ppm',
     & /,            27X, 'Flyash',             23X,F10.4,' lb/MMBtu',
     & /,            27X,'SO2 emissions       ',11X,F10.4,' lb/MMBtu',
     & /,            27X, 'NOx emissions (as NO2)',7X,F10.4,' lb/MMBtu',
     & /,            27X, '(Mole fraction of NOx as NO = 0.95)',
     & /,            27X, 'CO2 emissions             ',7X,F10.4,' lb/kWh ',
     & /,            27X, 'Slag from combustor',10X, F10.4, ' lb/hr',
     & /,            27X, 'Collected fly ash', 12X, F10.4, ' lb/hr',
     & /,            27X, 'sludge (wet)',       17X, F10.4, ' lb/hr',//)
C
C-----------------------------------------------------------------
C
C
      WRITE (NRPT, 600)
 600  FORMAT(/,'H. BASE CASE DIRECT CAPITAL COSTS ($1,000) ---------',
     &       '------------',
     & //,16X,'                                  Direct              ',
     &  /,16X,'Plant Section                   Capital Cost',
     &  /,16X,'-------------                   ------------ ')
C
      WRITE (NRPT, 610)DCCBV, DCCBS, DCCBM, DCSSB, DCHXBC, DCHXBM,
     &                 DCHRB, DCFB81, DCFB82, DCFB83, DCIDB
 610  FORMAT(
     & /,'                      Coal Combustor Vessel ', 11X,F10.1,
     & /,'                  Coal Combustor Structural ', 11X,F10.1,
     & /,'               Coal Combustor Miscellaneous ', 11X,F10.1,
     & /,'                                Slag Screen ', 11X,F10.1,
     & /,'               Heat Exchanger, ceramic part ', 11X,F10.1,
     & /,'              Heat Exchanger, mettalic part ', 11X,F10.1,
     & /,'              Heat Recovery Steam Generator ', 11X,F10.1,
     & /,'              Fabric Filter process area 81 ', 11X,F10.1,
     & /,'              Fabric Filter process area 82 ', 11X,F10.1,
     & /,'              Fabric Filter process area 83 ', 11X,F10.1,
     & /,'                                     ID Fan ', 11X,F10.1)
C
C
C
      WRITE (NRPT, 620)
 620  FORMAT(/,'I. BASE CASE DIRECT CAPITAL COST PARAMETERS ---------',
     &       '-------',
     & //,16X,'                                          ',
     &  /16X,'Plant Section                      Parameter Value',
     &  /16X,'-------------                      --------------- ')
C
      WRITE (NRPT, 630)BCRF, GFGSSB,  AHXBC, AHXBM, MFGHRB,
     &                 GFGFB, RACB, NBB, MAB, GFGIDB,
     &                 F81, F82, F83, F84, A81, A82, A83, A84,
     &                 B81, B82, B83, B84, C81, C82, C83, C84
 630  FORMAT(
     & /,'                 Coal Combustor Refractory ', 11X,F10.1, ' $/ft2',
     & /,'  Fluegas vol.flowrate to Slag Screen ', 5X,F16.1, ' ft3/hr',
     & /,'    Heat Exchanger area, ceramic part ', 11X,F10.1, ' ft2',
     & /,'   Heat Exchanger area, mettalic part ', 11X,F10.1, ' ft2',
```

```
      & /,'         Fluegas mass flowrate  to HRSG ', 11X,F10.1, ' lb/hr',
      & /,'Fluegas vol.flowrate to Fabric Filter ', 11X,F10.1,' ft3/hr',
      & /,'  Air to Cloth ratio of Fabric Filter ', 11X,F10.1,
      & /,'       Number of bags per compartment ', 11X,F10.1,
      & /,'               Massflow of ash to silo ', 11X,F10.1, ' lb/hr',
      & /,'       Fluegas vol.flowrate to ID Fan ', 5X, F16.1,' ft3/hr',
      & /,'      Factor f for FF process area 81 (see note below)',
      &          2X,F3.1,
      & /,'      Factor f for FF process area 82 ', 11X,F10.1,
      & /,'      Factor f for FF process area 83 ', 11X,F10.1,
      & /,'      Factor f for FF process area 84 ', 11X,F10.1,
      & /,'      Factor a for FF process area 81 ', 11X,F10.1,
      & /,'      Factor a for FF process area 82 ', 11X,F10.1,
      & /,'      Factor a for FF process area 83 ', 11X,F10.1,
      & /,'      Factor a for FF process area 84 ', 11X,F10.1,
      & /,'      Factor b for FF process area 81 ', 11X,F10.1,
      & /,'      Factor b for FF process area 82 ', 11X,F10.1,
      & /,'      Factor b for FF process area 83 ', 11X,F10.1,
      & /,'      Factor b for FF process area 84 ', 11X,F10.1,
      & /,'      Factor c for FF process area 81 ', 11X,F10.1,
      & /,'      Factor c for FF process area 82 ', 11X,F10.1,
      & /,'      Factor c for FF process area 83 ', 11X,F10.1,
      & /,'      Factor c for FF process area 84 ', 11X,F10.1,
      & //,'Note: Factors are described in model documentation,
      & Section 5.10')
C
C----------------------------------------------------------------------
C
 720  CONTINUE
      RETURN
      END
C
```

D.3 Sample Output

The following is a sample output from an ASPEN simulation of the EFCC system with the integrated performance and newly developed cost model. The output echos the values of the performance variables accessed in the ASPEN flowsheet that are inputs to the cost model. Key cost parameter assumptions are also summarized, including cost model parameters, process contingency factors, maintenance cost factors, and the unit costs of consumables, byproducts, fuel, and ash disposal. Total direct, total capital, fixed operating, and variable operating costs are summarized. A summary of plant power output and internal auxiliary power consumptions is given. Based on the total capital and annual costs, the fixed charge and variable cost levelization factors, and the net plant power output, the cost of electricity is estimated. The plant thermal efficiency and a summary of selected plant emissions are also printed.

Because cost data were not available for the combustor, slag screen, and ceramic heat exchanger, the direct costs of these process areas are not computed. Because these costs are not calculated, any totals based upon these are also not reported. These include total capital costs, maintenance costs, and cost of electricity. However, the cost model contains all variables and equations necessary to calculate these totals.

EXTERNALLY-FIRED COMBINED CYCLE POWER PLANT

" Version 1.0"

*** SUMMARY OF SAMPLED ASPEN FLOWSHEET PERFORMANCE PARAMETERS ***

PLANT SECTION	FLOWSHEET PERFORMANCE PARAMETERS	VALUE	UNITS
General Process Parameters	Combustor heat loss	0.50	%
	Mass flow of water to comb. water wall	0.00	lb/hr
	Comb. heat directed to water walls	0.00	%
	Carbon conversion of coal	99.00	%
	Auxiliary air flow per lb of coal	0.80	lb/lb coal
	Air leakage in CerHx	0.50	%
	CerHx heat loss	1.00	%
	CerHx Effectiveness	0.85	
	Gas turbine inlet temperature	2300.00	deg F
	GT compressor pressure ratio	13.50	
	Steam injection mass flow	0.00	lb/hr
	Water injection mass flow	0.00	lb/hr
	Ambient temperature	59.00	deg F
	Stack gas temperature	181.00	deg F
Coal Feed	Mass flow of coal to combustor	189533.16	lb/hr
	Plant thermal input	0.2132E+04	MMBtu/hr
	Auxiliary air mass flow rate	151624.54	lb/hr
	Pressure of aux.air at PA fan outlet	16.40	psia
	Avg. den. of aux.air at PA fan outlet	0.0763	lb/ft3
Coal Combustor	Massflow of fluegas at combustor	3823914.29	lb/hr
	Excess air ratio	2.30	
	Density of fluegas at combustor	0.0130	lb/ft3
	Temp. of fluegas at combustor outlet	2702.64	deg F
	Combustor heat loss	0.121054E+08	Btu/hr
	Combustor heat to water wall	0.000000E+00	Btu/hr
Gas Turbine	Net GT power output	-.471502E+09	Btu/hr
	GT compressor work requirement	0.556309E+09	Btu/hr
	Temperature of air at CerHx inlet	724.29	deg F
	Temperature of air at CerHx outlet	2405.89	deg F
	Massflow of air to GT comp inlet	3500074.77	lb/hr
	Massflow of air at GT comp outlet	3293220.35	lb/hr
	Massflow of air at GT turb inlet	3081155.07	lb/hr
	Massflow of air at GT turb outlet	3501164.05	lb/hr
	Heat duty of the CerHx	-.134407E+10	Btu/hr
	Net Efficiency of the Gas Turbine	35.43	%
HRSG	Massflow of flue gas at HRSG inlet	3838386.35	lb/hr
	Temperature of flue gas at HRSG inlet	1506.49	deg F
	Enthalpy of HRSG to steam cycle	-.129815E+10	Btu/hr
Steam Cycle	Mass flow of steam to HP steam turb	827650.55	lb/hr
	Mass flow of steam to IP steam turb	827650.55	Btu/hr
	Mass flow of steam to LP1 steam turb	827650.55	Btu/hr
	Mass flow of steam to LP2 steam turb	780138.40	Btu/hr
	Massflow of makeup water	26854.82	lb/hr
	Massflow of steam condensate	780138.40	lb/hr

	Massflow of polished water	806993.22	lb/hr
	Work req. of condenser pump	0.494103E+05	Btu/hr
	Work req. of boiler feed water pump	0.692764E+07	Btu/hr
	Work output of HP steam turbine	-.967743E+08	Btu/hr
	Work output of IP steam turbine	-.133728E+09	Btu/hr
	Work output of LP1 steam turbine	-.173772E+09	Btu/hr
	Work of output LP2 steam turbine	-.128059E+09	Btu/hr
	Enth. of inlet stream to LP2 turb	-5700.00	Btu/hr
	Heat requirement for FGD reheat	-.433491E+08	Btu/hr
	Heat requirement for FGD reheat	12.53	MW
	Enth. of exit stream of LP2 turb	-5864.14	Btu/hr
	Work req. of steam inj. compressor	0.168199E-07	Btu/hr
	Work req. of water inj. pump	0.335242E-08	Btu/hr
	Net ST power output	-.482138E+09	Btu/hr
	Net Efficiency of the Steam Turbine	37.14	%
FGD Parameters	Sulfur removal efficiency	0.90	
	Calcium to sulfur ratio	1.05	Ca/S
	Massflow of flue gas at FGD inlet	3838386.35	lb/hr
	Massflow of SO2 at FGD inlet	12734.44	lb/hr
	Massflow of slurry to FGD	211442.06	lb/hr
	Massflow of water recir. to FGD	66709.81	lb/hr
	Massflow of sludge from FGD (wet)	55188.32	lb/hr
	Density of flue gas at FGD inlet	0.0518	lb/ft3
	Density of slurry to FGD	53.2650	lb/ft3
Environmental	Massflow of ash at slagger	18407.46	lb/hr
	Massflow of flue gas at FF inlet	3838386.35	lb/hr
	Massflow of flue gas to ID fan	3927930.37	lb/hr
	Massflow of flue gas exiting stack	3927930.37	lb/hr
	Massflow of CO2 in stack gas	430395.10	lb/hr
	Moleflow of SO2 in stack gas	19.88	lbmol/hr
	NOx in stack gas	0.50	lb/MMBtu
	Moleflow of O2 in stack gas	14841.23	lbmol/hr
	Moleflow of H2O in stack gas	12226.20	lbmol/hr
	Moleflow of fluegas exiting stack	136358.63	lbmol/hr
	Pressure at ID fan inlet	13.27	psia
	Temp. of flue gas at reheater inlet	116.40	deg F
	Temp. of flue gas at reheater outlet	160.31	deg F
	Density of flue gas at FF inlet	0.0528	lb/ft3
	Average density of flue gas	0.0575	lb/ft3
Miscellaneous	Heat of Combustion of coal	26147069.94	J/Kg

PROCESS SECTION	PRESSURE DROP (psi)
Air filter	-0.5000
Steam injection mixer	0.0000
CerHx-air side	-3.5000
Gas turbine Mixer 1	0.0000
Gas turbine Mixer 2	0.0000
Gas turbine Mixer 3	0.0000
Gas turbine Mixer 4	15.1300
Combustor air blower	16.4000
Combustor mixer	15.1300
Coal combustor	-0.1100
Slag screen	-0.3560

CerHx airleak-hot side	0.0000
CerHx-flue gas side	-0.7600
CerHx airleak-cold side	0.0000
HRSG reheater	-0.0280
HRSG superheater	-0.0280
HRSG-High pressure evaporator	-0.0280
HRSG-high pressure economizer	-0.0280
HRSG-low pressure evaporator	-0.0280
FGD mixer	-0.2000
FGD reactor	-0.0200
Fabric filter	-0.2700

Note: A negative number or 0 implies a pressure drop whereas a positive number implies outlet pressure

COST SUMMARY
Externally Fired Combined Cycle

A. KEY INPUT ASSUMPTIONS--

Performance Assumptions:

Superficial gas velocity in combustor:	4.000 ft/sec
Gas residence time in combustor:	8.000 sec
Combustor refractory thicknes:	0.667 ft
Calcium to sulfur ratio for FGD:	1.050
Air to cloth ratio:	2.000 acfm/ft2
No. of filter bags/comp.:	360.00
Generator efficiency:	0.986
Centrigual fan efficiency:	0.850
Electric motor efficiency:	0.950
FGD SO2 removal efficiency:	0.900
Frac. coal ash as slag:	0.650
Frac. of flyash in low size range:	0.250
Frac. of flyash in med size range:	0.500
Frac. of flyash in high size range:	0.250
Low size flyash rem eff in SS:	0.001
Med size flyash rem eff in SS:	0.990
High size flyash rem eff in SS:	0.001
Low size flyash rem eff in HX:	0.001
Med size flyash rem eff in HX:	0.001
High size flyash rem eff in HX:	0.990
Low size flyash rem eff in FF:	0.990
Med size flyash rem eff in FF:	0.990
High size flyash rem eff in FF:	0.990

Notes: SS = Slag screen, HX = Ceramic heat exchanger
FF = Fabric filter, FGD = Flue gas desulfurization

Economic Assumptions:

Cost Year:	January 1990	Inflation Rate:	0.000
Plant Cost Index:	354.70	Real Escalation Rate:	0.000
Chemical Cost Index:	391.87	Plant Booklife:	30
Plant Capacity Factor:	0.650	Sales Tax Rate:	0.060
General Facilities Factor:	0.150	Real Return on Debt:	0.046
Indirect Construction Factor:	0.200	Real Ret. on Pref.:	0.052
Engr & Home Office Fees:	0.100	Real Ret. on Equity:	0.087

Project Contingency Factor:	0.300	Debt Ratio:	0.500
Pref. Stock Ratio:	0.150	Avg. Op. Labor Rate:	19.700
Fed. & State Taxes:	0.380	Number of Shifts:	4.750
Investment Tax Credit:	0.000	Const. Interest Rate:	0.050
Prop. Taxes & Insurance:	0.020	Year of Construction:	4
Cost of Environmental Permit:	1000.0		

Process Contingency and Maintenance Cost Factors:

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.010	0.030
Coal Combustor	0.700	0.040
Slag Screen	1.000	0.030
Ceramic Heat Exchanger	1.000	0.050
Boiler Feedwater	0.000	0.015
Gas Turbine	0.250	0.015
Heat Recovery Steam Generator	0.200	0.015
Steam Turbine	0.000	0.015
Flue Gas Desulfurization	0.100	0.044
Fabric Filter	0.100	0.011
ID Fan	0.100	0.030
General Facilities	0.050	0.030

B. CALCULATED DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS ($1,000) ----

Plant Section	Number of Operating	Units Total	Direct Capital Cost	Process Contingency
Coal Handling	1	1	8352.7	104.3
Coal Combustor	1	1	0.0	0.0
Slag Screen	1	1	0.0	0.0
Ceramic Heat Exchanger	1	1	0.0	0.0
Gas Turbine	1	1	28557.6	8918.6
Heat Recovery Steam Generator	1	1	10640.6	2658.5
Steam Turbine	1	1	22628.6	0.0
Boiler Feed Water	1	1	1243.1	0.0
Flue Gas Desulfurization	1	1	28791.4	3596.6
Fabric Filter	1	1	11057.2	1381.3
ID Fan	1	1	0.0	0.0
General Facilities	N/A	N/A	0.0	0.0

C. CALCULATED TOTAL CAPITAL REQUIREMENET ($1,000)----------------------

Description	Capital Cost
Total Direct Cost	------.
Indirect Construction Cost	-----.
Sales Tax	-----.
Total Plant Facilities Cost	------.
General Facilities Cost	------.
Engineering & Home Office Fees	------.
Environmental Permitting	------.
Total Process Contingency	------.

Project Contingency	------.
Total Plant Cost	------.
AFDC	------.
Total Plant Investment	------.
Preproduction (startup) cost	------.
Inventory Capital	------.
Initial Catalysts and Chemicals	------.
Land	------.
TOTAL CAPITAL REQUIREMENT ($1,000) ------------>	------.

[NOTE: Total capital costs are not reported because direct capital cost data are not yet available for the combustor, slag screen, or ceramic heat exchanger]

D. CALCULATED FIXED OPERATING COSTS ($/year) ----------------------

Description	Annual Cost
Operating Labor	4281992.
Maintenance Cost	-------.
Administration and Supervision	-------.
TOTAL FIXED OPERATING COST($/year) ------------>	-------.

[NOTE: Maintenance costs are not reported because direct capital cost data are not yet available for the combustor, slag screen, or ceramic heat exchanger. Maintenance costs depend on facilities costs for each major process area.]

E. CALCULATED VARIABLE OPERATING COSTS ------------------------------

1. CONSUMABLES ($/year)

Description	Assumed Unit Cost	Calc. Material Requirement	Calc. Annual Operating Cost
Sulfuric Acid:	104.82 $/ton	105.1 ton/yr	11017.
NaOH:	209.63 $/ton	89.1 ton/yr	18685.
Sodium Phos.:	0.67 $/lb	137.8 lb/yr	92.
Hydrazine:	3.05 $/lb	647.6 lb/yr	1975.
Morpholine:	1.24 $/lb	557.5 lb/yr	691.
Limestone:	17.15 $/ton	21144.2 ton/yr	362660.
FF bags:	78.14 $/bag	360.0 bags	28129.
Fuel Oil:	40.02 $/bbl	15047.5 bbl/yr	602211.
Adsorbent:	2.67 $/lb	1128.6 lb/yr	3011.
Water:	0.70 $/1000gal	102070.3 (1000 gal)/yr	71000.
LPG-Flare:	11.15 $/ton	2257.1 ton/yr	25164.
TOTAL CONSUMABLES ($/year) ---------------------->			1124634.

2. FUEL COST, and SLAG, ASH, and SLUDGE DISPOSAL($/year)

Coal:	1.53 $/MMBtu	189533.2 lb/hr	18623795.
Slag:	9.53 $/ton	11964.8 ton/yr	324587.
Ash:	9.53 $/ton	6426.1 ton/yr	174328.

Sludge:	10.77 $/ton	55188.3 ton/yr	1691801.

TOTAL VARIABLE OPERATING COST ($/year) --------------> 21939145.

F. CALCULATED COST OF ELECTRICITY--

Power Summary (MWe)	
Gas Turbine Output	138.18
Steam Turbine Output	141.30
Total Auxiliary Loads	14.22
Net Electricity	265.26

Auxiliary Electrical Loads (MWe)			
Coal Handling	0.89	Steam Cycle	2.04
Boilfeed Water	0.19	Flue Gas Des.	3.46
ID Fan	6.21	General Fac.	1.21
Prim. AirFan	0.22		

Capital Cost: ----.-- $/kW
Fixed Operating Cost: --.-- $/(kW-yr)

Incremental Variable Costs: 2.20 mills/kWh
Fuel Cost: 12.33 mills/kWh

Variable Operating Cost: --.-- mills/kWh

COST OF ELECTRICITY ------------------------------------> --.-- mills/kWh

[NOTE: Total costs are not reported because direct capital cost data are not yet available for the combustor, slag screen, or ceramic heat exchanger]

Fixed Charge Factor: 0.1034 Variable Cost Levelization Factor: 1.0000

THE PLANT HEATRATE (HHV): 8037.BTU/KWh
THE PLANT EFFICIENCY (%,HHV): 42.49

G. ENVIRONMENTAL SUMMARY --

INPUTS:	Coal	0.7145 lb/kWh
	FGD slurry water	0.4659 lb/kWh
	Raw water	0.1012 lb/kWh
	Total water	0.5672 lb/kWh
	Gas Turbine air	13.1949 lb/kWh
	Auxiliary air	0.5716 lb/kWh
	Total air	13.7665 lb/kWh
OUTPUTS:	Flyash(dry,15%O2)	2.6304 ppm
	Flyash	0.0078 lb/MMBtu
	SO2 emissions	0.5967 lb/MMBtu
	NOx emissions (as NO2)	0.5000 lb/MMBtu
	(Mole fraction of NOx as NO = 0.95)	
	CO2 emissions	1.6225 lb/kWh
	Slag from combustor	11964.8493 lb/hr
	Collected fly ash	6426.0541 lb/hr
	sludge (wet)	55188.3191 lb/hr

H. BASE CASE DIRECT CAPITAL COSTS ($1,000) ---------------------

Plant Section	Direct Capital Cost
Coal Combustor Vessel	0.0
Coal Combustor Structural	0.0
Coal Combustor Miscellaneous	0.0
Slag Screen	0.0
Heat Exchanger, ceramic part	0.0
Heat Exchanger, mettalic part	0.0
Heat Recovery Steam Generator	11050.0
Fabric Filter process area 81	7860.0
Fabric Filter process area 82	250.0
Fabric Filter process area 83	1590.0
ID Fan	0.0

I. BASE CASE DIRECT CAPITAL COST PARAMETERS ----------------

Plant Section	Parameter Value	
Coal Combustor Refractory	0.0	$/ft2
Fluegas vol.flowrate to Slag Screen	100000000.0	ft3/hr
Heat Exchanger area, ceramic part	1.0	ft2
Heat Exchanger area, mettalic part	1.0	ft2
Fluegas mass flowrate to HRSG	3842000.0	lb/hr
Fluegas vol.flowrate to Fabric Filter	950000.0	ft3/hr
Air to Cloth ratio of Fabric Filter	2.0	
Number of bags per compartment	360.0	
Massflow of ash to silo	10.4	lb/hr
Fluegas vol.flowrate to ID Fan	100000000.0	ft3/hr
Factor f for FF process area 81 (see note below)	0.9	
Factor f for FF process area 82	0.8	
Factor f for FF process area 83	0.6	
Factor f for FF process area 84	0.0	
Factor a for FF process area 81	0.8	
Factor a for FF process area 82	0.0	
Factor a for FF process area 83	0.3	
Factor a for FF process area 84	0.0	
Factor b for FF process area 81	0.2	
Factor b for FF process area 82	0.0	
Factor b for FF process area 83	0.3	
Factor b for FF process area 84	0.0	
Factor c for FF process area 81	0.0	
Factor c for FF process area 82	0.0	
Factor c for FF process area 83	0.1	
Factor c for FF process area 84	0.0	

Note: Factors are described in model documentation, Section 5.10