

**Stochastic Modeling of Coal Gasification
Combined Cycle Systems: Cost Models
for Selected Integrated Gasification
Combined Cycle (IGCC) Systems**

Task 2 Topical Report

**H.C. Frey
E.S. Rubin**

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Morgantown Energy Technology Center
P.O. Box 880
Morgantown, West Virginia 26507-0880**

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

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NOMENCLATURE

English Letter Symbols

A	=	Heat transfer surface area (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 (\$)
$CAT_{i,j}$	=	Initial catalyst requirement for chemical i in plant section j (mass units)
$CHEM_{i,j}$	=	Initial chemical requirement for chemical i in plant section j (mass units)
CTP	=	Coal throughput per gasifier (lb/hr/gasifier)
d	=	Diameter (ft)
d_{min}	=	Minimum diameter (ft)
D_b	=	Sorbent bulk density (lb/ft ³)
DC_j	=	Direct capital cost of plant section j (\$1,000 January 1989)
e_a	=	annual escalation rate for plant equipment (fraction)
$E(x)$	=	Expected (mean) value of variable x .
f_{ash}	=	Fraction of ash in coal (fraction)
f_{att}	=	Zinc ferrite attrition rate per 80 cycles (fraction)
f_{CaCO_3}	=	Purity of limestone (fraction)
f_{cr}	=	Capital recovery factor (fraction)
f_{EHO}	=	Engineering and home office cost factor (fraction)
f_{HS}	=	Volume fraction of hydrogen sulfide in syngas (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_{SO_2}	=	Fraction of off-gas volume flow rate that is sulfur dioxide
f_{vcif}	=	Variable cost levelization factor (ratio)
FOC	=	Fixed operating cost (\$/year)
i	=	Interest cost for spent funds (fraction)
I_{PCI}	=	<i>Chemical Engineering</i> 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
H	=	Height (ft)
HV_{coal}	=	Heating value of coal (Btu/lb)
IC	=	Inventory capital (\$1,000)
L_s	=	Sulfur loading in sorbent (weight fraction)

NOMENCLATURE (Continued)

$m_{i,j,k}$	=	Mass flow rate of species i at the plant section j inlet or outlet k , (lb/hr in all cases except for coal, where units are tons/day)
$M_{i,j,k}$	=	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)
$N_{o,j}$	=	Number of 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)
OC_i	=	Operating cost for category i (\$/year)
$P_{j,k}$	=	Pressure at the plant section j inlet or outlet k (psia)
PP_i	=	Preproduction cost of category i (\$1,000)
PPC	=	Total preproduction cost (\$1,000)
q	=	heat flux (Btu/hr)
Q_{coal}	=	Energy flow of coal (MMBtu/hr)
$R_{\text{Ca/S}}$	=	Calcium-to-sulfur molar ratio (lbmole Ca/lbmole S)
R^2	=	Coefficient of Determination (decimal)
r_{tax}	=	sales tax (fraction)
$R_{\text{Ca/S}}$	=	Molar ratio of calcium to sulfur (lbmole Ca/lbmole S)
s	=	sample standard deviation estimated from a data set
S_c	=	Fresh sorbent charge (lb)
S_s	=	Spent sorbent charge (lb)
SF	=	Shift factor (labor shifts/day)
t_a	=	Absorber cycle time (hours)
$T_{j,k}$	=	Temperature at the plant section j inlet or outlet k ($^{\circ}\text{F}$)
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- $^{\circ}\text{F}$))
UC_i	=	Unit cost of item i (\$/mass unit)
V_c	=	Volume of sorbent charge (ft ³)
$V_{i,j,k}$	=	Volume flow rate of species i at plant section j inlet or outlet k (acfm)
V_s	=	Superficial velocity (ft/s)
$\text{Var}(x)$	=	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 (kW)

NOMENCLATURE (Continued)

Greek Letter Symbols

η_{HS}	=	Hydrogen sulfide removal efficiency (fraction)
η_{Ox}	=	Fraction of oxidant that is oxygen, molar basis (fraction)
ΔT_{LM}	=	Log mean temperature difference (°F)
σ	=	population standard deviation

Subscripts

a	=	ambient
AS	=	Administrative and support labor
EHO	=	engineering and home office costs
EP	=	environmental permitting
f	=	fresh
FC	=	fixed operating cost
Fuel	=	fuel
i	=	inlet
IC&C	=	initial catalyst and chemicals
ICC	=	indirect capital cost
M	=	maintenance
ML	=	maintenance labor
MM	=	maintenance materials
OC	=	variable operating cost
PC	=	process contingency
PJ	=	project contingency
o	=	outlet
s	=	spent

Species

ads	=	adsorbent
ash	=	ash
bent	=	bentonite
byp	=	byproduct (either sulfur or sulfuric acid)
cat	=	catalyst
cf	=	coal feed
chem	=	chemical
ci	=	corrosion inhibitor
cm	=	coal feed, moisture and ash free basis
coke	=	coke
cons	=	consumables
Cl	=	chlorine
COS	=	carbonyl sulfide
CW	=	cooling water
finer	=	finer
fo	=	fuel oil

NOMENCLATURE (Continued)

hps	=	high pressure steam
hy	=	hydrazine
HS	=	hydrogen sulfide
lime	=	lime
L	=	limestone
LPG	=	liquified petroleum gas
mo	=	morpholine
O	=	oxygen
Ox	=	oxidant
pw	=	polished water
rw	=	raw water
s	=	sulfur
sa	=	sulfuric acid
sbd	=	scrubber blowdown
sh	=	sodium hydroxide
so	=	soda ash
sp	=	sodium phosphate
ss	=	Selexol solvent
ssy	=	saturated syngas
su	=	surfactant
syn	=	syngas
ww	=	waste water
zf	=	zinc ferrite

Equipment/Plant Sections

BF	=	Boiler Feed Water System
BS	=	Beavon-Stretford Tailgas Treating System
C	=	Claus Plant Sulfur Recovery Plant
CH	=	Coal Handling
CM	=	Air Boost Compressor
CW	=	Cooling Water
CY	=	High Temperature, High Pressure Cyclones
G	=	Gasification, High Temperature Gas Cooling, Ash Removal, and Particulate Scrubbing
GF	=	General Facilities
GT	=	Gas Turbine
HR	=	Heat Recovery Steam Generator
L	=	Limestone Handling
LT	=	Low Temperature Gas Cooling and Fuel Gas Saturation
OF	=	Oxidant Feed
PC	=	Process Condensate Treatment System
PR	=	Air Boost Compressor Precooler
S	=	Selexol Acid Gas Removal
SA	=	Sulfuric Acid Plant
SC	=	SCOT Tail Gas Treating System
SF	=	Sulfation

NOMENCLATURE (Continued)

SS	=	Sub-Systems for Lurgi Gasifier
ST	=	Steam Turbine
SU	=	Sulfation
ZF	=	Zinc Ferrite

1.0 INTRODUCTION

This report documents cost models developed for selected integrated gasification combined cycle (IGCC) systems. The objective is to obtain a series of capital and operating cost models that can be integrated with an existing set of IGCC process performance models developed at the U.S. Department of Energy (DOE) Morgantown Energy Technology Center (METC). These models are implemented in ASPEN, a Fortran-based process simulator. Under a separate task, 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 cost models presented here is to explicitly characterize uncertainties in capital and annual costs, supplanting the traditional approach of incorporating uncertainty via a "contingency factor."

The IGCC systems selected by DOE/METC for cost model development include the following:

- KRW gasifier with cold gas cleanup
- KRW gasifier with hot gas cleanup
- Lurgi gasifier with hot gas cleanup.

For each technology, 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 the 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, the costs of utilities and reagent consumption, and credits from byproduct sales. Uncertainties in cost parameters are identified for both capital and operating cost models.

1.1 Overview of IGCC 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 (SO₂) 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 SO₂ 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 not yet been applied with U.S. coals (Cichanowicz and Offen, 1987; Damon and Giovanni, 1987). Furthermore, DOE and others have supported development of more advanced alternatives for control of SO₂ and NO_x emissions from pulverized coal (PC) power plants.

Integrated coal gasification combined cycle (IGCC) systems are an alternative to conventional pulverized coal (PC) combustion. IGCC systems are capable of NO_x emissions comparable to or less than those of PC plants equipped with SCR, as well as high (over 95 percent) levels of SO₂ control (Cool Water, 1986). Compared to other advanced technology options, such as fluidized bed combustion and fuel cells, IGCC offers advantages in terms of phased construction, fuel flexibility, and extensive use of conventional components (with the exception of a few key process sections). Political concern over acid rain (for which SO₂ and NO_x are precursors) may accelerate the timetable for more 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 IGCC 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 synthetic fuel gas ("syngas") derived from coal for natural gas or oil, a coal-fueled gas turbine combined cycle power plant results. By integrating the steam cycle to generate steam from the high temperature coal gasification process, the overall thermal efficiency can be optimized. Advantages of IGCC plants over PC plants include higher thermal efficiency, higher sulfur removal efficiency, lower NO_x emissions, lower particulate emissions, reduced solid waste due to byproduct recovery of elemental sulfur, reduced cooling water requirements (because gas turbines, rather than boiler/furnaces, generate a large portion of the power), reduced land requirements and a capability to burn coal, oil, or natural gas (SFA, 1983).

IGCC systems are characterized by a modular design which allows phased construction of the system, flexibility in fuel use, and flexibility in design. A phased approach to IGCC construction might be based on an initial natural gas- or oil-fired simple cycle gas turbine. In later phases, a combined cycle and a coal gasification system are added, gradually increasing the electrical output of the facility and resulting ultimately in an IGCC power plant. The advantage of phased construction is that a utility can add new capacity incrementally, reducing the amount of capital that is at risk at any given time due to uncertain electric load

growth forecasts. The lead time required for simple cycle and combined cycle power plants is significantly less than that for PC power plants. Furthermore, the utility is not committed to natural gas or oil as a long-term fuel in a phased IGCC project (Fluor, 1986). Perhaps equally important, the plant is not dependent on a single fuel (e.g., coal) in the event of a gasification system failure. Because IGCC plants are modular, they can be built in a wide range of sizes and with a variety of options for specific equipment, such as gasifiers and gas turbines.

1.1.1 IGCC Designs

A number of variations of IGCC power plant designs exist, based primarily on differences in the coal gasifier technology. Both oxygen and steam are necessary reactants in the coal gasification process, which produces a syngas containing carbon monoxide and hydrogen. Alternate gasifier designs may use either oxygen or air as the oxidant. The primary difference in gasifier design is the type of bed in which the coal is gasified. The three generic types of gasifiers are moving-bed, fluidized-bed, and entrained-flow. In a moving bed gasifier, coal flows downward counter-current with the steam and oxidant, and the highest temperatures are reached toward the bottom of the reactor. A prominent example of this type of gasifier is the Lurgi design. In a fluidized bed reactor, the coal is well-mixed with steam and oxidant, leading to a more uniform temperature distribution in the gasifier. The Kellogg-Rust-Westinghouse (KRW) gasifier is an example of a fluidized bed design. In an entrained flow gasifier, the coal is gasified in a plug flow reactor in which the coal and reactants move co-currently through the reactor. The Texaco gasifier is the most common entrained-flow design. The gasifier design affects the temperature of the fuel gas, composition of the fuel gas (e.g., methane content, presence of tars and oils), ability to handle certain coals (e.g., caking coals), ability to handle fines, and oxidant and steam requirements, among other factors (SFA, 1983).

The Electric Power Research Institute (EPRI), a privately funded research consortium in the electric utility industry, has sponsored a number of performance and cost evaluations of IGCC technologies. These include technologies based on the Texaco, KRW, Shell, British Gas Corporation/Lurgi (BGC/L), and Dow gasifier technologies (e.g., Fluor, 1983a, 1983b; Fluor, 1984; Fluor, 1985; Parsons, 1985; Fluor Daniel, 1989). Most of the studies sponsored by EPRI have focused on the entrained-flow Texaco gasifier technology. This is partly because the Texaco gasifier has more operating experience on the demonstration plant level than other technologies such as Shell, KRW, and British Gas/Lurgi (SFA, 1983). EPRI, in cooperation with others, has co-sponsored the Cool Water gasification program, the first IGCC demonstration plant, based on the Texaco technology. The Cool Water IGCC plant was in service for five years. Emissions of SO_2 and NO_x were well below both NSPS and the more stringent emission permit limits for Cool Water (Cool Water, 1988).

An oxygen-blown (oxygen used as the oxidant) KRW-based system may offer some advantages over a comparable Texaco-based system, including reduced oxygen consumption, a lower temperature and pressure gasifier, a syngas with a higher heating value, and fewer parasitic loads (due largely to reduced oxygen consumption). Initial comparisons of Texaco and KRW based systems indicate that the heat rates (efficiency), capital costs, and levelized costs for these systems are nearly identical, even though a larger capital cost contingency factor was used for the KRW cost estimate (Fluor, 1985).

Lurgi gasification technology is the oldest of the technologies most commonly considered for IGCC systems. The Lurgi dry ash coal gasification process was developed in the 1930s in Germany, and over 150 gasifiers have been installed internationally since, most notably nearly 90 gasifiers in South Africa. Because the Lurgi gasifier is a moving-bed design in which the coal flows countercurrently with the steam and oxidant, the temperature varies throughout the reactor. The exiting gas temperature is lower than for other gasifier designs, and the "cold gas efficiency" (percent of chemical energy in the coal contained in the syngas) is higher than for other gasifier designs (SFA, 1983). However, the syngas typically contains oils and tars, which must be removed from the syngas in conventional IGCC designs to avoid deposition on downstream equipment. The removal of tars and oils reduces the heating value of the syngas, and requires additional scrubbing equipment, increasing capital costs (Parsons, 1985; Bechtel et al, 1988). Alternatives to conventional gas cleanup systems may eliminate the requirement for syngas cooling in Lurgi-based systems, therefore preventing the condensation of tars and oils and eliminating the associated gas scrubbing equipment, resulting in significant cost savings (Corman, 1986).

1.1.2 DOE/METC Research

DOE/METC has sponsored, and continues to sponsor, a number of system analysis studies to identify potentially promising process configurations and to provide performance and, in some cases, cost data for these. These include conceptual design studies of new and repowered Texaco-based IGCC plants (UTC, 1983), a phased Westinghouse (now KRW)-based plant (WEC, 1983), oxygen-blown Lurgi-based systems for power generation and synthetic natural gas production (Cincotta, 1984; Zahnstecher, 1984), simplified air-blown (air used as the oxidant) systems using Lurgi gasifiers and "hot" gas cleanup (Corman, 1986), and the performance and cost of "hot" gas cleanup and sulfur recovery systems (Klett et al., 1986; O'Hara, Chow, and Findley, 1987), as opposed to the lower temperature "cold" gas sulfur removal systems assumed in EPRI studies.

Conventional IGCC designs are based on "cold" gas cleanup, in which the fuel gas from the gasifier is cooled to a sufficiently low temperature (e.g., 100 °F) that the Selexol or similar sulfur removal process can be used to separate H₂S from the fuel gas. A focus of

research at the METC is the development of "hot" gas cleanup systems, in which sulfur compounds may be removed from the gasifier or the fuel gas at high temperature (e.g., 1,000 °F).

One of the most promising hot gas cleanup configurations is an air-blown Kellogg-Rust-Westinghouse (KRW) IGCC system. The hot gas cleanup system features in-bed desulfurization in the fluidized bed gasifier with limestone or dolomite, subsequent sulfur removal from the fuel gas with a zinc ferrite sorbent, and high efficiency cyclones and ceramic filters for particulate removal. The off-gas from the zinc ferrite reactor, which contains sulfur compounds, is recycled to the gasifier. The advantage of such a system compared to a base case oxygen-blown system with cold gas cleanup is that, (1) it does not require an expensive and energy consuming oxygen plant, (2) it eliminates the capital costs associated with sulfur recovery (all sulfur is disposed with the spent limestone or dolomite), and (3) it reduces the amount of fuel gas cooling required prior to combustion in the gas turbine, thereby improving the plant thermal efficiency.

Testing of an air-blown KRW-based system with hot gas cleanup at the process development unit (PDU) level has been conducted (KRW, 1988). M.W. Kellogg has presented some results of a performance and cost analysis of such a system, although no detail was provided on costs (Banchik and Cover, 1988). M.W. Kellogg and Bechtel, under a cooperative agreement with DOE as part of the clean coal technology program, are designing and constructing a 63.5 MW demonstration plant of a KRW gasifier IGCC system with hot gas cleanup using the fixed-bed zinc ferrite process, due for startup in 1991 (KRW, 1988).

In the fixed-bed zinc ferrite process, sulfur is removed from the syngas by reaction with a sorbent consisting of zinc ferrite pellets. Absorption occurs until just before "breakthrough", at which point the sorbent is saturated. The absorber is then taken off-line, and the syngas is diverted to another zinc ferrite reactor vessel containing regenerated sorbent. Sulfided sorbent is regenerated using air as a reactant and steam as a diluent, to prevent the heat released in the exothermic regeneration reactions from sintering the sorbent. The regeneration off-gas containing sulfur dioxide is then recycled to the gasifier, in KRW-based designs. METC has developed a recommended design basis for a fixed-bed zinc ferrite hot gas desulfurization system, which is useful for further analyses (Kasper, 1988). Because a significant amount of METC resources have been committed to developing the air-blown KRW-IGCC in-situ desulfurization hot gas cleanup concept, it is appropriate to consider this as an important case to include in cost model development.

METC has also sponsored development of IGCC systems with hot gas cleanup based on the Lurgi gasification technology. The higher cold gas efficiency of Lurgi gasifiers

compared to other gasifiers assumed for IGCC design can result in a higher plant efficiency, because a larger portion of the energy input enters the combined cycle system through the fuel gas rather than only through the steam cycle. The conversion efficiency of energy entering the gas turbine is much higher than that of energy entering the steam cycle. The exit temperature of syngas from a Lurgi or similar gasifier provides a more direct match with the temperature window of hot gas cleanup systems, thereby eliminating any requirement for syngas cooling. Lurgi-based IGCC systems with hot gas cleanup therefore offer the potential for simplified plant designs. General Electric, under contract to METC, has been involved in analysis, testing, and development of hot gas cleanup systems for simplified Lurgi-based IGCC plants. These efforts include conceptual cost and design studies (Cincotta, 1984; Corman, 1986), proof-of-concept system design studies (e.g., Smith, 1987), and construction of a proof-of-concept system for a moving bed gasifier with hot gas cleanup (e.g., Cook, 1989).

The focus of the General Electric research program is testing of a moving-bed zinc ferrite desulfurization system, in which sorbent circulates continuously between an absorber and regenerator vessel, as opposed to the fixed-bed system in which the sorbent remains in one vessel which is cycled between absorption and regeneration duty. The moving bed design offers advantages in terms of a steady flow of regeneration off-gases and the elimination of a requirement for steam as a diluent (Smith, 1987). However, at this time, only limited design data and no detailed cost data are publicly available for this system.

1.2 Availability of IGCC Performance Models

METC has developed a number of performance simulations of IGCC systems in the ASPEN (Advanced System for Process ENgineering) modeling environment. 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.

The METC IGCC performance models are used to calculate mass and energy balances for IGCC systems, to track environmental species, to conduct sensitivity analyses of performance parameters, and to evaluate design modifications. The IGCC designs that have been modeled to date include oxygen-blown systems with cold gas cleanup based on the

Texaco, BGC/L, and KRW gasifier technologies, an air-blown KRW-based system with in-bed limestone desulfurization and hot gas zinc ferrite sulfur removal, and an air-blown Lurgi-based system with hot gas zinc ferrite sulfur removal. In some cases, such as the air-blown KRW and Lurgi based systems, documentation is not available for these models, other than the ASPEN input file. These models typically consist of about 80 unit operation blocks and 4 to 8 flowsheet sections. While the bulk of the models are comprised of generalized unit operation blocks, there are a number of Fortran blocks and design specifications which are specific to IGCC systems or to a flowsheet. There also are 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 flowsheets have been developed in a modular approach to allow sections to be "borrowed" from other flowsheets, substantially reducing development time of new IGCC simulation models.

1.2.1 Oxygen-Blown KRW IGCC With Cold Gas Cleanup

The ASPEN simulation of the oxygen-blown fluidized-bed KRW IGCC system with cold gas cleanup was originally based on a conceptual design of a 570 MW plant (Bechtel and WE, 1983c), and is documented in a METC report (Stone, 1985). An updated version is based on a more recent study (Fluor, 1985), but is not documented. The only information about the recent flowsheet is the input file itself (Stone and Craig, 1988). The main difference between the earlier and recent version of the flowsheet is in the pressure levels assumed in the steam cycle. Figure 1 shows a diagram of the flowsheet sections for the ASPEN flowsheet. The components or characteristics included in each flowsheet section, the number of unit operation blocks used in the ASPEN simulation, and a brief description of the function of each section are listed in Table 1. The material stream flows that link each flowsheet section are shown in Figure 1. Not shown in detail are "information" streams that show the flow of heat from various flowsheet sections to the steam cycle.

1.2.2 Air-Blown KRW IGCC With Hot Gas Cleanup

An ASPEN flowsheet of a KRW-IGCC with in-bed limestone sorbent, zinc ferrite hot gas cleanup, an advanced gas turbine (GE MS7000F) and a reheat steam cycle has recently been developed at METC. The only current documentation of this model, however, is the ASPEN computer code (Craig, 1988). The computer model includes 73 unit operation blocks in four major plant sections. From examination of the ASPEN input file, a simple conceptual diagram of the flowsheet was developed and is given in Figure 2. This flowsheet is considerably simpler than the system with cold gas cleanup, reflecting the simpler design of hot gas cleanup configurations of IGCC systems. Some features of the model are listed in Table 2.

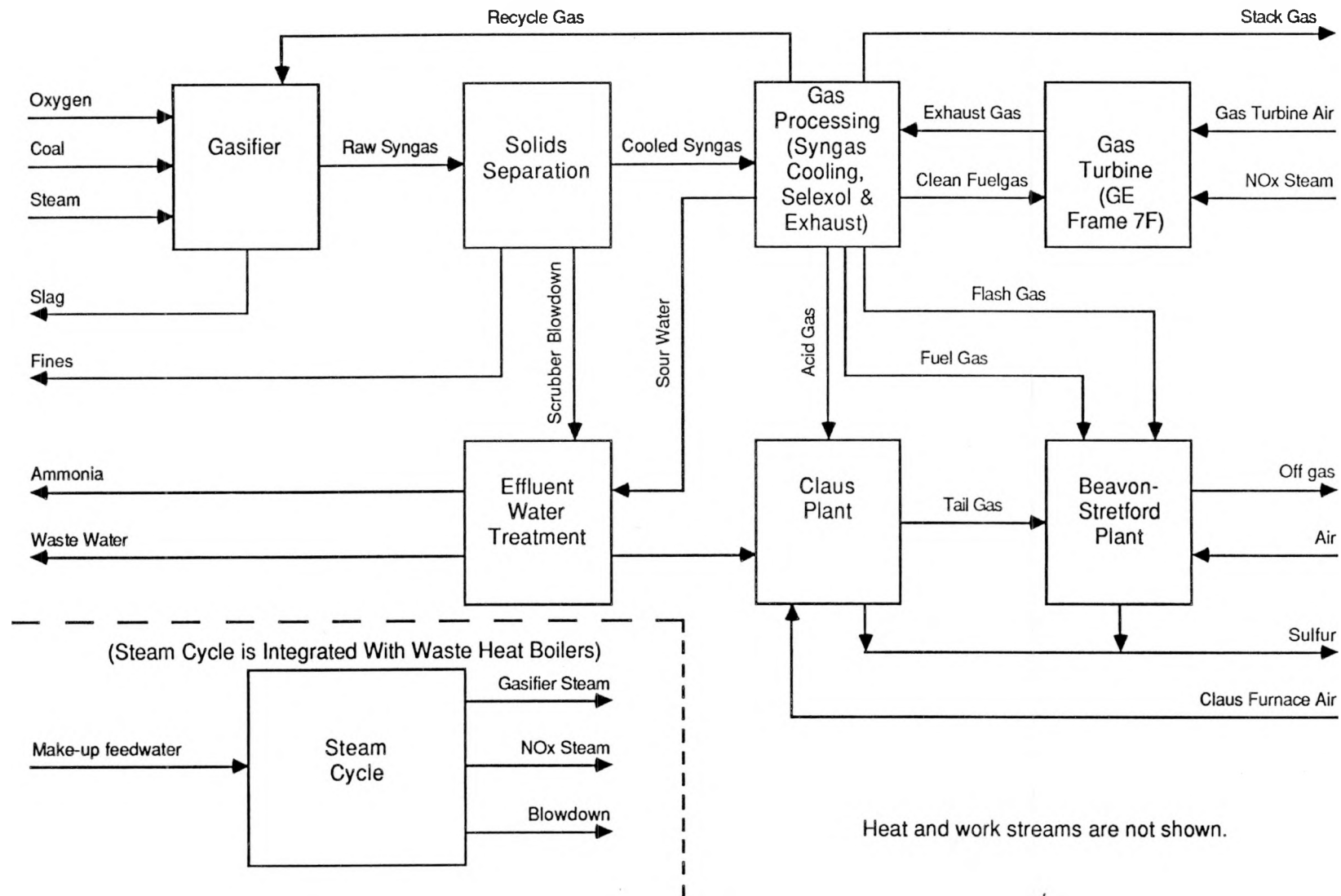


Figure 1. Conceptual Diagram of ASPEN Flowsheet for Oxygen-Blown KRW IGCC With Cold Gas Cleanup

Table 1. METC ASPEN Model of KRW-Based IGCC System: Plant Sections and Components

Flowsheet Section ^a	Components/Processes/Comments
Gasifier (6)	Coal decomposed based on ultimate analysis; Stoichiometric reactor model for slag and fines; Mix decomposed coal, steam, oxygen, recycle; Equilibrium reactor model for gasification; Ash removal after cyclones.
Solids Separation (4)	Gas cooling Particulate scrubbing to separate liquids and solids from gas, and split effluent to fines and purge water; Cooling (to reheat clean fuel gas); Water vapor knock-out, recycled to scrubber.
Gas Processing (8) (Cleanup and cooling)	Ammonia separation; Fuel gas recycle (to gasifier); ammonia/H ₂ S to effluent water treatment system; Selexol: acid gas removal to Claus plant, flash gases to Beavon-Stretford unit; Fuel split: fuel gas to Stretford unit; Reheat fuel gas (from Solid Sep. cooling); Cooling of gas turbine exhaust in high-pressure and low pressure HRSG.
Gas Turbine (14)	Reheated fuel gas mixed with steam and compressed air; Mixture burned in stoichiometric reactor; Multi-stage compression of air, bleeds for turbine blade cooling; Multi-stage expansion of combustion gases; Heat loss and pressure drop considered.
Effluent Water Treatment (4)	Purge water (scrubber) stream class change; Mix purge water with ammonia/H ₂ S stream; Cool mixture, separate wastewater, H ₂ S (to Claus), and ammonia.

(Continued on next page)

Table 1 (Continued). METC ASPEN Model of KRW-Based IGCC System: Plant Sections and Components

Flowsheet Section ^a	Components/Processes/Comments
Claus Plant (7)	Compress air, mix with acid gas and H ₂ S from water treatment section; Three stoichiometric reactor models to convert 95 percent of H ₂ S to elemental sulfur; Waste heat boiler; Separate sulfur and tail gas.
Beavon-Stretford Unit (7)	Compress air, compress Claus tail gas; Mix air, tail gas, fuel gas, and flash gas from the Selexol unit; Three stoichiometric reactor models to eliminate remaining H ₂ S; Separate sulfur and off-gas.
Steam Cycle (33)	Linked to other sections by heat streams;
HRSG (9)	Condenser, makeup water, pump, deaerator;
Auxiliaries (11)	Steam sides of HRSG and associated pumps; Heat recovery from Claus, Stretford, effluent water treatment;
Steam Turbine (7)	Four stage, no reheat; Steam from or split to Selexol, water treatment, and ammonia removal.

^a Numbers in parenthesis are number of unit operation blocks in each ASPEN flowsheet section

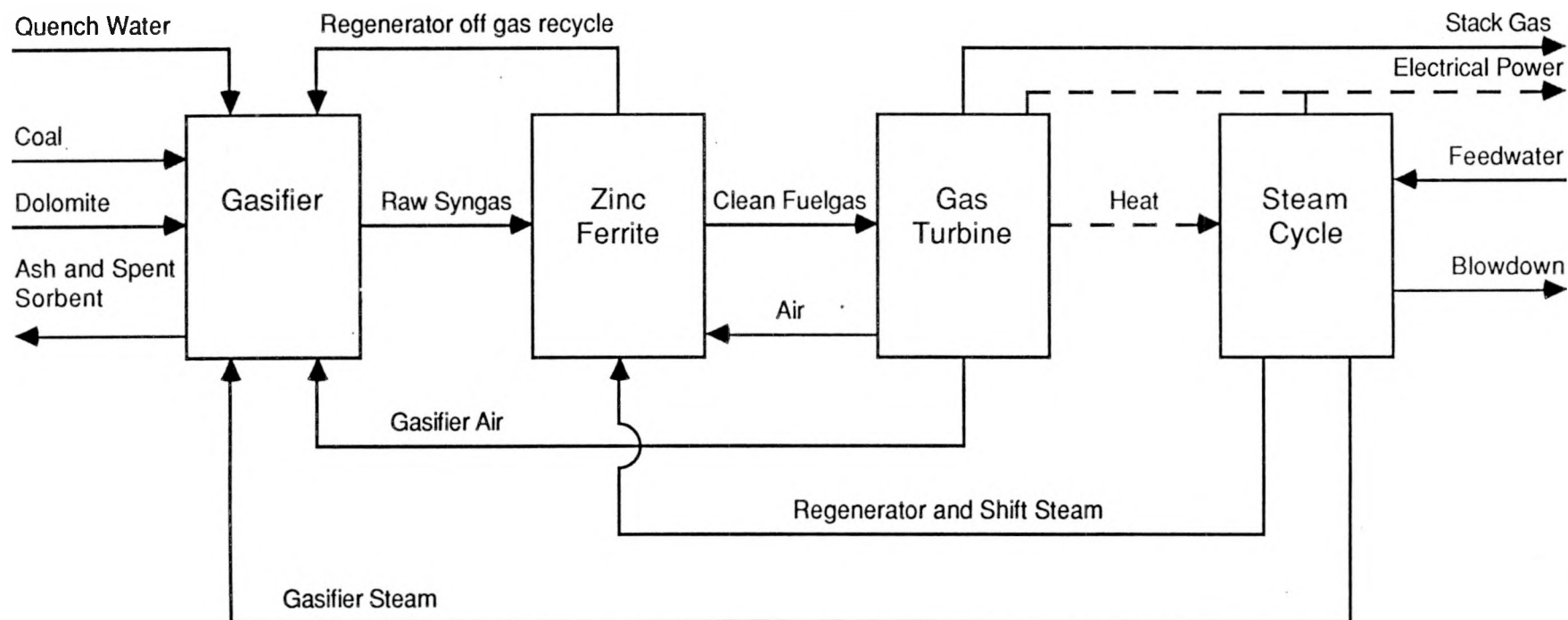


Figure 2. Conceptual Diagram of ASPEN Flowsheet of Air-Blown KRW IGCC System with In-Bed Desulfurization and Zinc Ferrite Process

Table 2. METC ASPEN Model of an Air-Blown KRW IGCC System With Hot Gas Cleanup: Plant Sections and Components

Flowsheet Section ^a	Components/Processes/Comments
Gasifier (10)	Air-blown KRW gasifier with in-bed desulfurization; characterization of fines, slag, and dolomite based on chemical analysis, in-bed desulfurization reaction based on specified conversion rate, gasification reactions based on equilibrium reaction model, recycle gas, raw gas quench cooling.
Zinc Ferrite (17)	Two-vessel zinc ferrite desulfurization system;
Stream Class Change (6)	ASPEN feature for handling gases and solids;
Absorption (5)	Mixing of steam and raw syngas for shift equilibrium to protect sorbent, sulfidation reactions;
Regeneration (6)	Mixing of diluent steam, regeneration air, and spent sorbent, preheating of inlet gas streams against outlet gas streams, regeneration reactions.
Gas Turbine (24)	General Electric model MS7000F Gas Turbine with air extraction; boost compressor for gasifier and zinc ferrite regeneration air; and gas-side heat exchange in HRSG;
Compressors (6)	Three stage compression, pressure ratio of 12, extraction for turbine vane cooling;
Combustion (4)	Stoichiometric reactor model for combustion reactions, heat loss, temperature control for gas turbine firing temperature (e.g., 2,300 °F);
Turbines (4)	Two stage expansion, inlet vane cooling air;
Booster (4)	Booster compressor for gasification air, compressor precoolers;
HRSG (4)	Exhaust gas cooling.
Steam Cycle (22)	Steam cycle, including steam-side HRSG, steam turbine, condenser, and deaerator.
HRSG (6)	1545 psia and 570 psia steam generation, 1465 psia superheated steam at 975 °F, 510 psia reheated steam;
Steam Turbine (8)	Reheat steam turbine;
Miscellaneous (12)	Deaerator, condenser, pumps, splitting of steam flows to gasifier, zinc ferrite regenerator, and zinc ferrite absorber shift reaction.

^a Numbers in parenthesis are number of unit operation blocks in each ASPEN flowsheet section

The gasifier section of the flowsheet models processes associated with coal gasification including the gasifier reactor vessel, ash removal from the gasifier, recycle of syngas, and high temperature gas cooling. The gasification section includes models to represent coal in terms of its chemical components, fines, slag, and ash, and a model to decompose dolomite into components of CaO, MgO, and CO₂. Inputs to the gasifier include coal, limestone, steam, and air. An equilibrium reaction model is used to represent the reactions occurring in the gasifier. A separate model is used to represent the in-bed sulfur adsorption, in which the sulfur removal efficiency is specified as an input. All solids are assumed to be removed from the syngas prior to the desulfurization system.

The zinc ferrite section of the flowsheet includes models of sulfur absorption and sorbent regeneration based on a fixed-bed design. Prior to absorption, the raw syngas is mixed with steam to promote the water-gas shift reaction for improved catalyst performance. The shift reaction is modeled using an equilibrium reaction model. The raw syngas and steam mixture is then mixed with the zinc ferrite sorbent, and the desulfurization reactions are modeled using a stoichiometric reactor model, in which the reaction conversion rates are specified. The clean fuel gas is an output of this flowsheet section.

The regeneration of the sorbent, which involves exothermic chemical reactions, requires air as a reactant and dilution with steam to prevent sintering of the sorbent. A stoichiometric reactor model, in which conversion rates are specified, is used to represent the regeneration reaction. The off-gas from the regenerator is cooled against the incoming diluent and recycled to the gasifier.

The flowsheet section for the MS7000F gas turbine includes three stages for the compressor (including air extraction for turbine vane cooling), combustion of the syngas (including heat losses and combustor pressure drop), two stages of turbines with a 2,300 °F turbine inlet temperature, and a gas-side heat recovery steam generator model. The steam cycle model includes pumps, economizers, boilers, a superheater, a reheater, a deaerator, condensate pumps, condenser, and a reheat steam turbine with high pressure, intermediate pressure, low pressure, and very low pressure stages.

1.2.3 Air-Blown Fixed Bed Gasifier with Hot Gas Cleanup

An ASPEN model of a fixed bed gasifier-based IGCC system with hot gas cleanup has been developed at METC. Like the KRW-based hot gas cleanup IGCC model, the only model documentation is the ASPEN input file (Klara, Rastogi, and Craig, 1988). The flowsheet includes a gasifier section, hot gas desulfurization using a moving bed zinc ferrite process, a model of General Electric MS7000F gas turbine, and a model of the steam cycle. The model

assumes that the off-gas from the zinc ferrite process is converted to sulfuric acid. There is no in-bed desulfurization in this model. A simple diagram of the ASPEN flowsheet is shown in Figure 3. Table 3 lists the features of the model.

The available performance models will be adopted as the basis for performance modeling of IGCC systems in ASPEN.

1.3 Selection of IGCC Systems for Cost Model Development

Based on METC's research interest and the availability of performance models of IGCC systems, three IGCC systems have been selected for cost model development. These systems are:

- Oxygen-blown KRW-based IGCC with cold gas cleanup (baseline)
- Air-blown KRW-based IGCC with hot gas cleanup
- Air-blown Lurgi-based IGCC with hot gas cleanup.

The cost models developed in this report are applicable to only these systems. The models are not intended for direct application to cost estimating for any other IGCC system configurations, nor are they intended for application to other types of coal gasification systems, such as fuel cell, coal-to-SNG, or general coal refinery systems.

1.4 Modeling Philosophy

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 definitions of

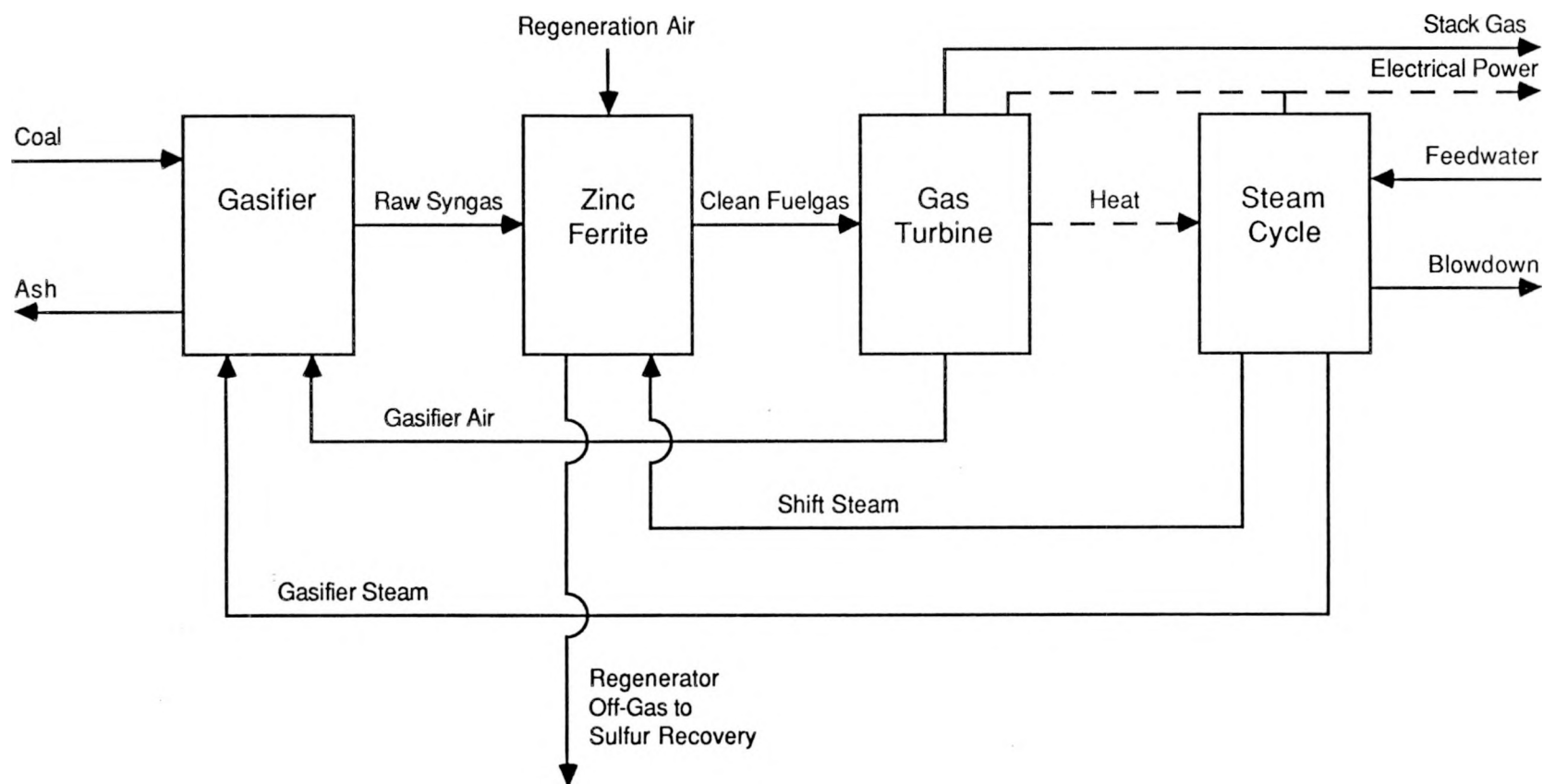


Figure 3. Conceptual Diagram of ASPEN Flowsheet of Air-Blown Fixed-Bed Gasifier IGCC System with Zinc Ferrite Process

Table 3. METC ASPEN Model of an Air-Blown Lurgi IGCC System With Hot Gas Cleanup: Plant Sections and Components

Flowsheet Section ^a	Components/Processes/Comments
Gasifier (11)	Air-blown generic fixed-bed dry-ash gasifier; characterization of coal, fines based on chemical analysis, gasification reactions based on equilibrium reaction model, allowance for specification of carbon conversion and fines carryover, provisions for in-bed desulfurization but not used in this input file.
Zinc Ferrite (26)	Two-vessel zinc ferrite desulfurization system based on General Electric moving bed design;
Stream Class Change (4)	ASPEN feature for handling gases and solids;
Absorption (6)	Mixing of steam and raw syngas for shift equilibrium to protect sorbent, sulfidation reactions;
Regeneration (13)	Mixing of recycle offgas, regeneration air, and spent sorbent, preheating of inlet gas streams against outlet gas streams, regeneration reactions.
Acid Plant Feed (3)	Simple model to account for power consumption
Gas Turbine (24)	General Electric model MS7000F Gas Turbine with air extraction and boost compressor for gasifier and zinc ferrite regeneration air and gas-side heat exchange in HRSG;
Compressors (6)	Three stage compression, pressure ratio of 12, extraction for turbine vane cooling;
Combustion (4)	Stoichiometric reactor model for combustion reactions, heat loss, temperature control for gas turbine firing temperature (e.g., 2,300 °F);
Turbines (4)	Two stage expansion, inlet vane cooling air;
Booster (4)	Booster compressor for gasification air, compressor precoolers;
HRSG (4)	Exhaust gas cooling.
Steam Cycle (22)	Steam cycle, including steam-side HRSG, steam turbine, condenser, and deaerator.
HRSG (6)	1545 psia and 570 psia steam generation, 1465 psia superheated steam at 975 oF, 510 psia reheated steam;
Steam Turbine (8)	Reheat steam turbine;
Miscellaneous (12)	Deaerator, condenser, pumps, splitting of steam flows to gasifier, and zinc ferrite absorber shift reaction.

^a Numbers in parenthesis are number of unit operation blocks in each ASPEN flowsheet section

these types of estimates are further discussed in Section 5.1. 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 three selected IGCC systems 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. Limitations of cost data availability for the three selected IGCC systems are discussed in the next section. Data from published studies can be used to develop cost models for specific process areas using regression analysis. Regression analysis is used extensively for cost model development in this study (and elsewhere). An overview of the key

concepts of regression analysis, and the philosophy of this study in applying regression analysis, is given in Section 1.7.

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. This approach was taken in several cases, such as for developing a model for zinc ferrite system direct capital cost. Such cases are noted in the text.

1.5 Cost Data Availability

The primary constraint on cost model development is the availability of performance and cost data from which to develop correlations between process performance and equipment capital cost. A summary of performance and cost studies of IGCC systems and components that are used for cost model development is given in Table 4. Limitations of these studies include disaggregation of capital costs to only about ten major process areas, differences in the battery limits of process areas from one study to another, differences in process technology assumed for a given plant section, and differences in the cost method employed.

A number of studies have been performed for DOE, including conventional designs with cold gas cleanup and advanced alternatives for hot gas cleanup. Studies of conventional systems include a series of performance and cost studies for Texaco, Lurgi, and KRW systems prepared by Bechtel for Argonne National Laboratory (Bechtel, 1983a; 1983b; 1983c), and a detailed study of Lurgi-based systems for synthetic fuel gas production by Foster Wheeler (Zahnstecher, 1984). A study of a KRW-based system developed by MW. Kellogg (Bostwick et al, 1981) was not used because it contained insufficient performance data and the cost estimates were inconsistent with more recent studies. Detailed performance and cost studies of hot gas cleanup systems include studies by General Electric (Cincotta, 1984; Corman, 1986) and an overview of sulfur recovery methods by Parsons (O'Hara, Chow, and Findley, 1987). A study by Gilbert/Commonwealth (Klett et al, 1987) was not used for cost model development because it presents only normalized (unitless) relative costs. A study by Westinghouse was specific to a phased addition at a particular site, and the sections of the report that describe the gasification section were deleted pending patent action (WEC, 1983). A study of a Texaco-based IGCC (UTC, 1983) contains little detail on the cost estimate. The available costs in the DOE studies do not use any consistent cost method or categorization of capital cost sections, and many are partially or not applicable to the cases of interest in this research.

Table 4. Guide to Studies Used for Cost Model Development

Report No.	Process ^a	Contractor ^b	Sponsor ^c	Date
ANL/FE-83-15	Texaco	Bechtel Group	DOE	1983
ANL/FE-83-16	BGC/L	Bechtel Group	DOE	1983
ANL/FE-83-17	KRW	Bechtel Group	DOE	1983
AP-2207	Texaco & Lurgi	Ralph M. Parsons	EPRI	1982
AP-3084	Texaco	Fluor Engineers	EPRI	1983
AP-3129	Shell	Fluor Engineers	EPRI	1983
AP-3486	Texaco	Fluor Engineers	EPRI	1984
AP-3980	BGC/L	Ralph M. Parsons	EPRI	1985
AP-4018	KRW	Fluor Engineers	EPRI	1985
AP-4395	Texaco	Fluor Engineers	EPRI	1986
AP-4826	Texaco	Fluor Technology	EPRI	1986
AP-5950	Texaco	Bechtel Group	EPRI	1988
AP-6011	BGC/L	Bechtel et al.	EPRI	1988
DOE/ET/14928	Lurgi	General Electric	DOE	1986
DOE/FE/05081	Lurgi	Foster Wheeler	DOE	1984
DOE/MC/20315	Lurgi	General Electric	DOE	1984
DOE/MC/21097	Sulfur Recovery	Ralph M. Parsons	DOE	1987
GRI-86/0009	KRW & Lurgi	Kellogg Rust Synfuels	GRI	1985
GRI-87/0154	KRW	Fluor Technology	GRI	1986
GRI-87/0155	KRW	Fluor Technology	GRI	1986
GRI-87/0156	Lurgi	Fluor Technology	GRI	1986
GRI-87/0159	Lurgi	Fluor Technology	GRI	1986
GRI-87/0160	KRW	Fluor Technology	GRI	1987
GRI-87/0169	KRW	Fluor Technology	GRI	1988
GS-6160	Texaco	Fluor Daniel	EPRI	1988
GS-6161	Shell	NUSCo	EPRI	1988
GS-6176	Shell	Florida Power & Light	EPRI	1989
GS-6283	Shell	BG&E	EPRI	1989
GS-6318	Dow	Fluor Daniel	EPRI	1989

^a BGC/L=British Gas Corporation/Lurgi; KRW = Kellogg Rust Westinghouse

^b BG&E = Baltimore Gas and Electric; NUSCo = Northeast Utilities Services Company

^c DOE = U.S. Department of Energy; EPRI = Electric Power Research Institute; GRI = Gas Research Institute.

EPRI has sponsored the most extensive publicly available performance and cost estimates of IGCC systems, including Texaco, Shell, KRW, British Gas Corporation/Lurgi, and Dow based systems. One contractor, Fluor Engineers, Inc., has prepared estimates for KRW (Fluor, 1985), Texaco (Fluor, 1983a; 1984; 1986), and Shell systems (Fluor, 1983b). Fluor Technology has prepared estimates of the performance and cost of alternative combined cycle systems integrated with Texaco gasification technology (Fluor Technology, 1986). Fluor Daniel has prepared a study of a Texaco system with Kraftwerk Union gas turbines (Fluor Daniel, 1988) and a study of a Dow-based IGCC system (Fluor Daniel, 1989). Bechtel Group, Inc. prepared an estimate of Texaco-based systems in a recent comparison of IGCC and conventional pulverized coal systems (Bechtel, 1988). Bechtel Group, Inc., in cooperation with others, also prepared a site-specific study of a BGC/L-based plant for Virginia Power (Bechtel et al., 1988). The Ralph M. Parsons Company has prepared studies of Texaco and BGC/L systems (Parsons, 1982; 1985). A recent set of studies focuses on site-specific applications of various IGCC technologies. In addition to the Bechtel study for Virginia Power, these include studies conducted with Baltimore Gas and Electric, Florida Power and Light, and Northeast Utilities Service Company concerning phased construction of Shell-based IGCC systems (BGE, 1989; FPL, 1989; NUSCo, 1988). These recent site-specific studies disaggregate costs based on a different definition of plant sections and plant section battery limits than the previous generic-site cost studies. Cost information is also available regarding the Cool Water demonstration plant (Cool Water, 1982; 1986; 1988).

All of the EPRI estimates present direct material costs for major process areas, generally including equipment and material installation costs, direct installation labor costs, sales tax, and indirect field costs and home office engineering, but may differ in the definition and battery limits of each process area. Process contingencies on a process area basis are estimated, and an overall project contingency is used. The total capital requirement also includes costs for royalties, spare parts inventory, organization and startup, working capital, allowance for funds during construction, and land. The amount of detail in the capital cost estimates varies from one study to another.

Figure 4 illustrates how an IGCC plant is divided into process areas in most of the EPRI capital cost estimates. This figure is based on the KRW-based IGCC system estimate prepared by Fluor (1985). Costs are reported only for ten major process areas, which are common to most of the studies for EPRI. Each of these process areas contains several specific pieces of equipment, which may vary from one system to another. For example, the KRW system uses a dry coal feed system, whereas the Texaco system uses a coal slurry feed system. Therefore, in many cases it is not possible to extract aggregated costs from one IGCC evaluation study and apply it to another system. Some reports contain more specifics; for

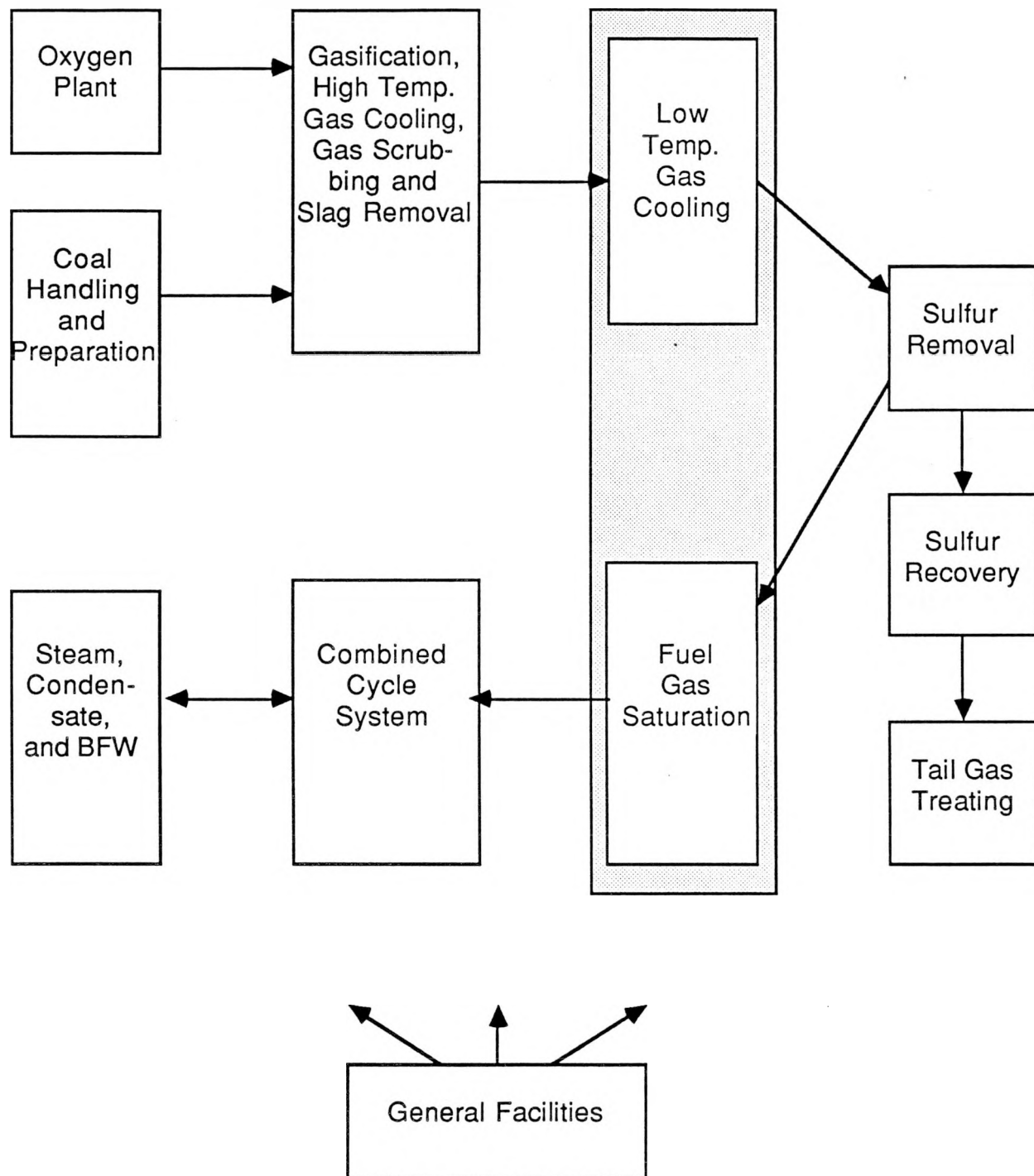


Figure 4. KRW-IGCC Process Cost Areas

example, the Bechtel report on a Texaco IGCC system includes a breakdown of general facilities costs into 20 line items.

Many of the cost estimates in the EPRI studies for specific equipment were developed in-house by EPRI's contractors or subcontractors. Manufacturers of some equipment were asked to submit quotes for a detailed commercial specification, while some quotes were obtained more informally by phone (e.g., Bechtel, 1988). General Electric provided Fluor with gas turbine combined cycle equipment costs. While Fluor considered about 350 separate items in their cost estimates of Texaco IGCC systems, much of the information is not included in the public report.

Several studies performed for the Gas Research Institute have been identified as useful for this research. These include evaluations of both KRW- and Lurgi-based systems. The studies of KRW-based systems include both coal-to-synthetic natural gas (SNG) facilities and IGCC plants, while the Lurgi-based studies are focused on coal-to-SNG facilities. Coal-to-SNG plants are similar to IGCC plants in the areas of coal feed, coal gasification, acid gas removal, and sulfur recovery. Kellogg Rust Synfuels, Inc. (KRSI) prepared the first detailed studies of coal-to-SNG systems for GRI, including both KRW and Lurgi gasification technology (Cover et al, 1985a; Cover et al, 1985b), and a comparison of how plant size affects the cost of both KRW and Lurgi-based systems (Cover et al, 1985c).

Fluor Technology has prepared all of the other studies of coal gasification systems for GRI. The studies of KRW-based coal-to-SNG plants include conversion of Western subbituminous coal to SNG (Smith, Hanny, and Smelser, 1986), conversion of Pittsburgh No. 8 coal to SNG (Smith and Smelser, 1987), and conversion of Pittsburgh No. 8 coal to SNG with reduced carbon conversion (Earley and Smelser, 1988a). Two studies, which focused on gas treating processes specific to coal-to-SNG systems, were based on the study of conversion of Western subbituminous coal to SNG, and contributed no additional information regarding gasification costs relevant to IGCC systems (Sandler and Smelser, 1987; Smelser et al, 1987). Three studies considered hot gas cleanup processes. These include performance and cost studies of: (1) a single train demonstration plant for either SNG production, electricity production, or both (Smelser, 1986a); (2) a commercial scale coal-to-SNG plant (Smith and Smelser, 1987); and (3) commercial scale coal-to-SNG and IGCC plants (Earley and Smelser, 1988b). The hot gas cleanup processes considered in these three studies include in-bed desulfurization in the gasifier, hot gas particulate removal using either sintered metal or ceramic candle filters, and hot gas desulfurization using the fixed bed zinc ferrite process.

The GRI-sponsored studies of Lurgi-based systems are based on conversion of lignite to SNG (Smelser, 1986b; 1986c). A summary of the GRI coal gasification technical analyses was recently published, which includes a brief summary of each of these studies (Smelser and Earley, 1988). All of the GRI studies are based on oxygen-blown systems. The coal-to-SNG plants are generally larger than IGCC plant sizes assumed in most studies, based on coal feed rate.

The capital cost model sections identified for each of the three selected IGCC systems are given in Table 5. These cost model sections are based on a comparison of the design basis available for each design either in the publicly available studies or the ASPEN input files for the performance simulations of each system. In particular, the performance model plant section most incompatible with available cost data is the moving bed zinc ferrite system assumed in the air-blown Lurgi hot gas cleanup system flowsheet. As indicated earlier, detailed performance and cost data are not publicly available at this time to support cost model development.

While the readily available data forms a basis for developing a cost model, it is desirable to obtain more detail on the most costly and/or most risky components of an IGCC system. Of the major subsystems in the KRW-based IGCC cost estimates prepared by Fluor, the most costly subsystems are: (1) the combined cycle power generation system, (2) the gasifier, and (3) the oxidant feed system. These subsystems are typically the most expensive ones for any oxygen-blown IGCC system. Technical risk may also reside in the sulfur removal and recovery system, especially for advanced concepts such as in-situ or zinc ferrite systems. Furthermore, alternative zinc ferrite system designs, such as the moving bed system, may merit cost model development when a detailed design basis and cost data become available. These systems therefore represent priorities for data collection and model development.

The early stage of a data collection effort involves determining the appropriate level of detail for the cost model, and then identifying the key performance parameters that significantly influence costs, consistent with the chosen level of detail. *Because the readily available cost estimates are disaggregated only to major plant sections, initial efforts should be based on developing a model suited to the data availability constraints.* The cost models developed in this research consist of capital cost correlations for approximately ten major process areas, based on currently available performance and cost evaluations. Parameters of these models include key process flow rates whose values will be estimated from the ASPEN performance simulations. Variable and fixed costs are estimated based on fuel consumption, makeup material requirements, catalyst and chemicals, other consumables, byproduct credits,

Table 5. IGCC Capital Cost Model Sections

Area No.	KRW, Cold Gas Cleanup	KRW, Hot Gas Cleanup	Lurgi, Hot Gas Cleanup
10	Oxygen Plant	Air Boost Compression	Air Boost Compression
20	Coal Handling	Coal Handling	Coal Handling
25	--	Limestone Handling	--
30	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasification, Coal Pressurization, Ash Depressurization
31	--	--	Coke Handling, Fines Agglomeration, Ash Removal
32	--	--	High Temperature Cyclones
35	--	Sulfation	--
40	Low Temperature Gas Cooling, Fuel Gas Saturation	--	--
50	Selexol Sulfur Removal	Zinc Ferrite Desulfurization	Zinc Ferrite Desulfurization
60	Claus Sulfur Recovery	--	Sulfuric Acid Plant
70	Tail Gas Treating	--	--
80	Steam, Condensate, Boiler Feed Water	Steam, Condensate, Boiler Feed Water	Steam, Condensate, Boiler Feed Water
85	Process Condensate Treatment	--	--
90	Combined Cycle	Combined Cycle	Combined Cycle
100	General Facilities	General Facilities	General Facilities

maintenance materials, maintenance labor, and operating labor costs. The total revenue requirement for the plant is estimated using the commonly accepted EPRI costing method (EPRI, 1986).

1.6 Integration of Performance and Cost Models

The cost models must be compatible with the ASPEN IGCC 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. Because the coal handling and oxidant feed sections are not modeled in the IGCC performance models, factors such as operating requirements are not modeled. However, these affect the annual costs of the system. Therefore, 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. Table 6 shows a comparison of parameters available in the ASPEN performance model of the KRW IGCC system with cold gas cleanup and parameter requirements for the capital cost model. Tables 7 and 8 provide comparable information for the KRW and Lurgi IGCC systems with hot gas cleanup.

ASPEN contains features that are potentially useful for estimating the costs of some process technologies. The ASPEN costing system consists of two main components: the cost estimation system (CES) and the economic evaluation system (EES). The former is used to estimate installed equipment costs and to track utility use within the plant. The latter is used to estimate capital investment, operating costs, and measures of profitability. The discussion in this section draws heavily on reports by Scientific Design Company (Schwint, 1985; 1986).

1.6.1 The ASPEN Cost Estimation System

The CES consists of a number of cost models, many of which correspond closely with unit operation blocks (e.g., pumps) and a number of others which are more general (e.g., vessels) which can be sized based either on a stream or a specific unit operation block (e.g., FLASH2). Each cost block contains a correlation for purchased equipment cost which is a function of one or more size parameters. The user may override the correlation with one of the form:

$$C = B \left(\frac{A_{\text{cap}}}{B_{\text{cap}}} \right)^n \quad (1)$$

Table 6. Major Plant Cost Sections, ASPEN Flowsheet Performance Sections, and Key Performance Parameters Affecting Cost for KRW IGCC With Cold Gas Cleanup

Area No.	Cost Section	ASPEN Flowsheet Section	Key Parameters
10	Oxygen Plant	N/A	Oxygen Flow Rate to Gasifier
20	Coal Handling	N/A	Coal Feed to Gasifier
30	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasifier Solids Separation	Coal Feed to Gasifier Syngas Flow Rate Gasifier Temperature, and Pressure, Coal Properties, Ash and Particulate Flow
40	Low Temperature Gas Cooling, Fuel Gas Saturation	Gas Processing Steam Cycle	Fuel Gas Flow Rate, Gas Temperature and Pressure
50	Selexol Sulfur Removal	Gas Processing	Fuel Gas Flow Rate, Hydrogen Sulfide Removal Efficiency
60	Claus Sulfur Recovery	Claus Plant	Claus Sulfur Production Rate
70	Tail Gas Treating	Beavon-Stretford	Beavon-Stretford Sulfur Production Rate
80	Steam, Condensate, Boiler Feed Water	Steam Cycle	Raw Water Input, Polished Water Flow Rate
85	Process Condensate Treatment	Solids Separation Effluent Water Primary Treatment	Scrubber Blowdown
90	Combined Cycle	Gas Turbine, Steam Cycle	Gas Turbine Model, Steam Flow Rate to Steam Turbine, Steam Turbine Electrical Output
100	General Facilities	N/A	Based on Other Costs

Table 7. Major Plant Cost Sections, ASPEN Flowsheet Performance Sections, and Key Performance Parameters Affecting Cost for KRW IGCC with Hot Gas Cleanup

Area No.	Cost Section	ASPEN Flowsheet Section	Key Parameters
10	Boost Air Compression	Gas Turbine	Extracted Air Flow Rate
20	Coal Handling	N/A	Coal Feed to Gasifier
25	Limestone Handling	N/A	Limestone Feed to Gasifier
30	Gasification, High Temperature Gas Cooling, Particulate and Ash Removal, Coal Pressurization	Gasifier	Coal Feed to Gasifier
35	Sulfation	Gasifier	Ash and Spent Limestone Flow Rate from Gasifier
50	Zinc Ferrite Desulfurization	Zinc Ferrite	Sulfur Flow Rate in Syngas, Syngas Volume Flow Rate, Syngas Pressure
80	Steam, Condensate, Boiler Feed Water	Steam Cycle	Raw Water Input Polished Water Flow Rate
90	Combined Cycle	Gas Turbine Steam Cycle	Gas Turbine Make and Model Steam Flow Rate to Steam Turbine, Steam Turbine Electrical Output
100	General Facilities	N/A	Based on Other Costs

Table 8. Major Plant Cost Sections, ASPEN Flowsheet Performance Sections, and Key Performance Parameters Affecting Cost for Lurgi IGCC With Hot Gas Cleanup

Area No.	Cost Section	ASPEN Flowsheet Section	Key Parameters
10	Boost Air Compression	Gas Turbine	Extracted Air Flow Rate
20	Coal Handling	N/A	Coal Feed to Gasifier
30	Gasification, Coal Pressurization, Ash Depressurization	Gasifier	Coal Feed to Gasifier
31	Coke Handling, Fines Agglomeration, Ash Removal	Gasifier	Coal Feed to Gasifier
32	High Temperature Cyclones	N/A	Syngas Flow Rate, Syngas Pressure
50	Zinc Ferrite Desulfurization	Zinc Ferrite	Sulfur Flow Rate in Syngas, Syngas Volume Flow Rate, Syngas Pressure
60	Sulfuric Acid Plant	Zinc Ferrite	Off-Gas Flow Rate, Sulfur Dioxide Concentration, Off-Gas Temperature
80	Steam, Condensate, Boiler Feed Water	Steam Cycle	Raw Water Input Polished Water Flow Rate
90	Combined Cycle	Gas Turbine Steam Cycle	Gas Turbine Make and Model Steam Flow Rate to Steam Turbine, Steam Turbine Electrical Output
100	General Facilities	N/A	Based on Other Costs

where:

C	=	Estimated cost at actual capacity
B	=	Base cost at base capacity
B_{cap}	=	Base capacity
A_{cap}	=	Actual capacity
n	=	Scaling exponent

and where capacity is a sizing parameter specified in the ASPEN User's Manual (MIT, 1987). The user-correlation form is not, in general, the form used by the ASPEN cost models; for example, the pump cost correlation is based on both the volumetric flow and the pressure head while the user correlation must be based only on the volumetric flow. Furthermore, a number of the cost models include several correlations to represent the breakdown of costs for complex systems (e.g., motor-driven pumps). While the user correlation may be useful for rough capital cost estimates, and while Equation (1) is often used by engineers to scale equipment costs, it lacks generality in terms of functional form. Furthermore, the user has no discretion over the input variable for the correlation.

Some of the major equipment in an IGCC system does not correspond directly to any of the existing ASPEN cost blocks. For example, while in principle a gas turbine could be costed by summing costs for a compressor, combustor vessel, turbine, and other components, in practice this would lead to unreliable results. A single cost block which models the cost of a gas turbine system is therefore more appropriate. Similarly, a heat recovery steam generator could be modeled as a collection of heat exchangers, but is more appropriately modeled as an integrated system. Although a gasifier could be modeled using one of the vessel cost models, the specialized nature of gasifiers implies that these cost correlations will not produce meaningful results. Therefore, the ASPEN CES is not an appropriate means for estimating IGCC capital costs.

Use of the ASPEN CES also results in an uneven level of detail (e.g., small pumps and large reactor vessels), and constrains the nature of the capital cost correlations that can be used. The lack of appropriate process area models (e.g., the gas turbine, HRSG, gasifier, and oxygen plant) indicates that either significant development work within the CES would be needed, or that an alternative approach should be developed.

While in principle existing ASPEN cost models could be modified for application to IGCC systems, such an effort is well beyond the scope of this research. As an alternative, Fortran blocks will be developed as needed to incorporate features essential to the capital cost model. This option is discussed below.

1.6.2 The ASPEN Economic Evaluation System

The Economic Evaluation System (EES) is used to determine operating costs, including the costs of imported utilities and raw materials. The EES can also be used to estimate the capital investment required to produce utilities on-site. In the case of an IGCC system, most or all of the utilities, except water, would be produced by the plant itself, except during startup. Utilities include water, steam, and electricity. Raw materials include both inputs and byproducts (e.g., coal, lube oil, Selexol solvent, sulfur).

The EES also calculates capital investment. For this purpose, the plant can be divided into cost sections. Each plant section can contain any number of cost blocks. In principle, if only one cost block contains a correlation, and the other cost blocks are specified as having zero capital cost, then the plant section cost would be based solely on the single cost block. The zero cost blocks might be used only to track utility use. Such an approach would minimize the number of blocks that would have to be created or modified, and could result in costs based on process areas and not specific equipment. However, limitations regarding the functional form and input variable selection for cost correlations would remain, and computation time would likely be excessive.

ASPEN uses an extensive cost indexing system to convert costs from one year to another for a variety of cost items (e.g., plant, equipment, labor, commodity, raw material, and fuel). Material and labor installation cost factors can be specified at either the block, section, or plant level to estimate installed costs based on purchased costs. There is provision to estimate costs for a variety of specific general facility items, including storage tanks, service buildings, site development, and land, plus keywords to specify indirect costs and additional depreciable or non-depreciable costs. There are also process and project contingency factor keywords. It appears that these last two can only be specified at the plant level. While many of these features are useful in calculating general facility costs and installation costs, the applicability of the EES depends on whether an IGCC cost model can be integrated within ASPEN.

Finally, ASPEN can be used to calculate the profitability of a plant. As a matter of convenience, it would be desirable to develop a cost summary similar in nature to the performance summary developed for the METC IGCC flowsheets. Such a summary would include annualized costs based on the EPRI costing method.

1.6.3 Cost Model Implementation via Fortran Blocks

There are three general approaches to developing a cost model of IGCC systems using the ASPEN performance models. They are:

- 1) Modify ASPEN to fully integrate cost correlations and cost models in a generalizable fashion;
- 2) Add user cost models and Fortran blocks to augment the existing CES and EES to facilitate calculation of special equipment costs and summary report generation; or
- 3) Do not use CES or EES. Instead, develop new Fortran block subroutines to access key flowsheet parameters, calculate costs based on correlations developed specifically for IGCC systems, incorporate uncertainty factors and error terms, calculate annualized costs according to EPRI guidelines, and write a summary report.

The first two approaches would utilize, to varying degrees, the current capabilities of the CES and EES. The first option, however, is beyond the scope of the current research to develop a cost model for IGCC systems. The second option is also less desirable since it would lead to potentially awkward or inefficient solutions.

The approach adopted here is to develop IGCC cost models independently of ASPEN, and to then use Fortran blocks to sample the key ASPEN performance variables needed to calculate the capital and annual costs. This approach requires the minimum resources to implement the cost models, and allows more effort to be devoted to the development and analysis of cost models. The cost models will be implemented as separate subroutines, to be linked with the ASPEN flowsheets via Fortran blocks.

1.7 Role of Regression Analysis

Regression analysis is used to help understand the interrelationships among a given set of variables. The use of regression analysis here is oriented toward developing useful and reasonable relationships between process area costs and key performance parameters. The emphasis is not on the use of extensive formal statistical tests but rather on the practical application of regression analysis for cost model development. Thus, some statistical tests, along with engineering judgments and the availability of data, are used to guide the selection of parameters, the representation of relationships in the regression models, and validation of the models. The "goodness" of the regression models are indicated with common summary statistics, graphical comparison of the model predictions with the actual data, and evaluation of the appropriateness of the model relationships with *a priori* engineering expectations.

This section will briefly discuss issues related to developing and interpreting the regression models. The issues related to developing the regression models include developing a data set for analysis, selecting parameters for inclusion in the model, and

validating the model. Specific issues related to the development and use of the models in this study are then discussed.

1.7.1 Overview of Multivariate Linear Least Squares Regression Analysis

The discussion in this section draws on Ang and Tang (1975), Chatterjee and Price (1977), DeGroot (1986), Dillon and Goldstein (1984), Edwards (1976), Montgomery and Peck (1982), and Weisberg (1985). An overview of key concepts is presented; details of multivariate regression can be obtained elsewhere in many texts such as the ones cited here.

In general, regression analysis involves describing the mean and variance of a random variable, Y , as a function of the value of another variable, X , or a set of variables $\mathbf{X}=(X_1, X_2, X_3, \dots, X_k)$. The variables in the vector $\mathbf{X}$ may take on specific values $\mathbf{x}=\mathbf{x}_i(x_{1,i}, x_{2,i}, x_{3,i}, \dots, x_{k,i})$. For each value $\mathbf{x}_i$ in an actual data set, there is a corresponding value y_i . We use the notation $E(Y|\mathbf{X}=\mathbf{x}_i)$ to indicate the mean, or expected value, of Y associated with a specific vector of values $\mathbf{x}_i$ of the variables $\mathbf{X}$. The notation $\text{Var}(Y|\mathbf{X})$ represents the conditional variance of Y on $\mathbf{X}$. If we expect that the value of Y can be estimated from a weighted linear combination of the k variables in $\mathbf{X}$, and if the conditional variance of Y is independent of the specific values $\mathbf{x}_i$ of $\mathbf{X}$, then:

$$E(Y|\mathbf{X}=\mathbf{x}) = \alpha + \beta_1 X_1 + \beta_2 X_2 + \beta_3 X_3 + \dots + \beta_k X_k \quad (2)$$

$$\text{Var}(Y|\mathbf{X}) = \sigma^2 = \text{constant} \quad (3)$$

The parameters in the linear equation are estimated, based on a limited number, n , of observed pairs of $(\mathbf{x}_i, y_i)$, using multi-variable linear regression with constant variance. The linear regression model is written as:

$$E(Y'|\mathbf{X}=\mathbf{x}_i) = a + b_1 X_1 + b_2 X_2 + b_3 X_3 + \dots + b_k X_k \quad (4)$$

or as:

$$Y' = a + b_1 X_1 + b_2 X_2 + b_3 X_3 + \dots + b_k X_k + \epsilon \quad (5)$$

The linear fit is usually obtained by selecting the values of a and b_j to minimize the sum of the square of the errors between $E(Y'|\mathbf{X}=\mathbf{x}_i)$ from Equation (4) and the values of Y from actual data, y_i . Equation (5) differs from Equation (4) in that the model is used to predict the conditional random values of Y' , rather than the conditional expected value of Y' . Equation (5) includes an error term, ϵ , that represents the variance in Y that is unexplained by the model. Thus, for a specific data point $\mathbf{x}_i$, there is a corresponding data value y_i , a conditional mean value $E(Y'|\mathbf{X}=\mathbf{x}_i)$, and a conditional random distribution for Y' . Using the method of least squares, as documented in any standard text, we obtain estimates for the coefficients of the

regression model. It is important to recall that the coefficients of the model, a and b_j , known as the partial regression coefficients, and the values of $E(Y'|X=x_j)$ or the parameters of the conditional distribution for Y' calculated using the model, are only estimates of the respective "true" population values of the parameters α and β_j and the "true" population of the values of Y associated with each value x_j .

Common statistical measures of the adequacy of the regression model in describing the data set (X,Y) include the standard error of the estimate, the coefficient of determination, the t-test for significance of each partial regression coefficient, and the F-test for the significance of the regression model and coefficient of determination. Confidence intervals, in addition to significance tests, can also be used. Proper application of these statistics requires the existence of certain properties in the data set (X,Y) and in the regression model. Several of these key assumptions are:

- random sample of n paired values (X,Y) (e.g., values of X are not preselected or screened)
- X and Y are multivariate normal
- for each value of x , there is an associated normal population of Y
- for each value of x , the variance of Y is constant
- no error in the measurement of X
- residual errors are not autocorrelated
- residual errors are normally distributed
- residual errors have constant variance

While these assumptions are often only approximately satisfied when developing regression models, the use of statistical evaluation methods based on these assumptions may provide some insight to guide the development of the model, even if a strict interpretation of the results is not correct. Therefore, blind application of significance tests to accept or reject parameters may not be appropriate. The most important consideration in selecting variables for use in a model, and for selecting the functional form of the model, is the analyst's knowledge of the substantive area under study and of each of the variables. The analyst will generally have expectations regarding the sign and magnitude of the coefficient for each variable, as well as which variables should be most significant in predicting the dependent variable.

The use of statistical tests is thus viewed here as an aid to, but not as a substitute for, the judgment of the analyst regarding the relationships among the variables. For example, it is common to test the significance of a model parameter by determining whether it is possible to reject a hypothesis that its coefficient is equal to zero. However, in many practical regression situations, it is known, based on theory or experience, that the coefficient must be greater than zero and, therefore, such a significance test is not particularly relevant. The potential inability

to reject the hypothesis that a coefficient is zero in a regression model may be more an artifact of a small number of data points than due to a lack of relationship between Y and the predictive variable of concern.

Statistical tests are useful in identifying the independent variables which are relatively more important in predicting Y than others for the available data. For example, one can examine a correlation matrix of X and Y to determine which variables X_i are most highly correlated with Y . These variables are logical candidates for inclusion in the regression model. However, if a potential predictive variable X_i is also highly correlated with another variable X_j , then the inclusion of both may not significantly improve the model and may lead to counter-intuitive results in terms of the sign or magnitude of the coefficient for one of the variables. In such cases, one of the variables would be excluded from the model. Statistical tests can be used to identify independent variables that have only a weak predictive power. These variables would also typically be excluded from the model. A few of the statistical measures used to evaluate regression models will be discussed here, with an indication of how they are used in this study.

The issues of statistical tests and model validation are closely linked. Statistical tests are used to determine the adequacy of the model in representing a known data set. To the extent that the model is used only to interpolate information from within the data set, checking the adequacy of the model is the same as model validation. A regression model can be used for prediction beyond the range of the original data set only if there is some basis in prior experience, industry practice, or physical theory for the relationships between variables. If the form of the regression model is not based on theoretical or expert judgment about the relationship between the dependent and independent variables, the model should not be used for extrapolation. The user is cautioned that the primary purpose of the models developed in this study is for interpolation within the range of data values used to develop the models. Furthermore, the user is cautioned that the models are intended for application with very specific systems. Throughout the report, the limitations on the ranges of predictive variables and discussion of the design basis for process areas will be presented.

In using multiple regression models, it is easy to inadvertently extrapolate beyond the original domain for X , because that domain is jointly defined by the pairing of the values of each independent variable used to generate the model. Therefore, range checks on each independent variable separately will not guarantee the avoidance of "hidden" extrapolation. However, because the regression models are developed with some engineering basis for the relationship between variables, hidden extrapolation may be acceptable, and individual range checks on the independent variables will be used as a practical convenience.

Standard Error

The standard error of the estimate is the standard deviation of the residual errors ε for Y' . The standard error is a measure of the variability in Y that is not captured by the model. If the functional form of the regression model is "correct", this variability can be attributed to factors that are not quantified in the database and therefore cannot be investigated quantitatively. If the functional form of the regression model is not appropriate, then some portion of the standard error may be associated with an incorrect choice of the model, rather than unexplainable variability in the data set. Therefore, it is often useful to compare alternative functional forms of the model in terms of the standard error.

The standard error is estimated based on the residual sum of squares and the degrees of freedom of the residuals. The residual sum of squares is the sum of the squares of the difference between the values of $E(Y'|X=x_i)$ estimated by the model in Equation (4) and the values y_i from the data. The degrees of freedom of the regression model are the number of variables, k . The degrees of freedom of the residuals are the number of data points less the number of partial regression coefficients, including the intercept term. Thus, the standard error is given by:

$$s = \sqrt{\frac{\sum_{i=1}^n [E(Y'|X=x_i) - y_i]^2}{n-k-1}} \quad (6)$$

This is an unbiased estimate of the standard deviation of the error. The error is assumed to be normally distributed with a mean of zero. In practice, this assumption may be difficult to verify, particularly for a small number of observed data points. Typical methods for evaluating the normality of the error include plotting the residuals against the fitted values $E(Y'|X=x_i)$, or plotting the errors on normal probability paper. A normality test may also be based on a one-sample Kolmogorov-Smirnov test (e.g., see DeGroot, 1984). In this test, the estimated cumulative probability distribution (cdf) for the errors is compared to a cdf based on the standard normal distribution. The maximum difference between the values of the sample and normal cdf's, adjusted for sample size, is the basis for estimating the test statistic. If the test statistic is larger than a specified value, based on the acceptable significance level for the test, then the hypothesis that the errors are normally distributed is rejected.

The estimate of the standard error is dependent on the actual data as well as the number of data points. As the quantity $(n-k-1)$ becomes small, the estimate of the standard error will tend to increase. The standard error can be used to place a confidence interval on

the values of Y' using Equation (4) or to generate conditional random values of Y' using Equation (5) and a probabilistic modeling capability. In the envisioned application of the regression models developed in this work, the standard error will be used as a basis for generating conditional random values of Y' .

Coefficient of Determination

The most commonly used measure of the adequacy with which a regression model fits the data is the coefficient of determination, R^2 , which is defined as:

$$R^2 = 1 - \frac{\sum_{i=1}^n (y_i - E(Y'|X=x_i))^2}{\sum_{i=1}^n (y_i - E(Y'))^2} \quad (7)$$

The numerator of the fractional term is the sum of the square of the residual errors between the actual data and the predicted conditional expected values of Y' from Equation (4). The denominator is the sum of the square of the differences between the actual data and the sample mean. The value of the coefficient of determination is interpreted as the proportion of the total variance in Y which is explained by the regression model, and it varies from 0 to 1, with values near 1 typically considered to represent "good" fits. The coefficient of determination is the square of the multiple correlation coefficient, R , between Y and the regression model. The multiple correlation coefficient is a measure of the degree of linear relationship between the dependent variable Y' and the linear combination of predictive variables.

The coefficient of determination is not a sufficient measure of the goodness of the model. At a minimum, evaluation of a regression model should include consideration of how reasonably the functional form and values of the coefficients represent the expected relationships between variables, the significance level of the coefficients and the regression model as a whole, and a graphical comparison of the model results with the actual data. The coefficient of determination may be highly influenced by extreme data points. If those data points are removed, the correlation coefficient may be drastically altered. The addition of a new data point may lead to a large change in the value of the coefficient of determination. Also, if the range of the predictive variables is reduced or increased, the correlation coefficient may change considerably.

Statistical Significance of the Model

It may be appropriate to consider a significance test for the correlation coefficient. A significance test based on the t-statistic can be used for this purpose to test the hypothesis that the correlation is not significantly different from zero. The hypothesis that a parameter is equal to zero is known as the null hypothesis. The likelihood that a parameter is significantly different from the null hypothesis is determined using a test statistic, such as the t-test. The value of the test statistic computed from the data is then compared to the value of the statistic estimated for the significance level of the test. It is common to use significance levels of 0.05 or 0.01 as the basis for comparison. If the probability of obtaining a value of the test statistic is less than the significance level (e.g., 5 percent or 1 percent), then the null hypothesis is rejected as being sufficiently improbable that it is regarded as false.

The null hypothesis for the correlation coefficient is a hypothesis that the correlation is zero. A correlation of zero implies that the regression model is not useful, and that the best predictor for the value of Y is the mean of Y. Instead of doing a significance test, it is also possible to use a transformation of the correlation coefficient for use in developing a confidence interval for the correlation (Edwards, 1976). However, statistical tests on the correlation coefficient are related to statistical tests on the coefficient of determination. Furthermore, a test of the null hypothesis for the coefficient of determination is implicitly a test of the null hypothesis that the regression coefficients for the predictive parameters are all zero (Edwards, 1976; Dillon and Goldstein, 1984). This hypothesis is commonly tested using the F test statistic. Thus, an F test implicitly is a test of the null hypothesis for the coefficient of determination as well as for the partial regression coefficients for the predictive parameters.

The F test involves first computing the F-ratio of the regression model, which is related to the coefficient of determination as follows:

$$F = \frac{\left(\frac{R^2}{k} \right)}{\left(\frac{1 - R^2}{n - k - 1} \right)} \quad (8)$$

As the coefficient of determination becomes large, the value of the F-ratio increases. The value of the F-ratio is then compared to the value of the F-distribution (published in many texts) for a selected significance level based on the degrees of freedom of the numerator (k) and denominator (n-k-1) of the F-ratio. Therefore, the F-test is influenced by both the number of data points and the number of predictive parameters included in the model. If the F-ratio is larger than the selected value of the F-distribution, then it is possible to reject the hypothesis

that all the regression coefficients are equal to zero. However, rejection of this hypothesis does not imply that all of the regression coefficients are significantly different from zero; it only implies that at least one coefficient is significantly different from zero. Furthermore, even if the regression model is statistically significant, it may not necessarily be the best model of the data or even a theoretically valid model of the data. In this study, the F-ratio is compared to a significance level of 0.001 as the basis for rejecting the null hypothesis. In cases where the significance level is higher than 0.001, the significance level of the F-ratio is reported.

To test the significance of individual regression coefficients, a commonly used technique is a t-test. For each regression coefficient, most computer regression packages will report the results of a t-test of the hypothesis that the individual regression coefficients are significantly different from zero. If a regression coefficient is not significantly different from zero, it can be deleted from the model with usually little effect on the residual error. In addition, the standard error for each coefficient is generally reported, which permits the evaluation of confidence intervals for the coefficients, using the t-distribution.

The regression models in this report have been developed using Statworks™, a statistics package for the Apple Macintosh™ computer. This package reports the statistics discussed here, and also facilitates use of the Kolmogorov-Smirnov test for testing the normality of the errors.

1.7.2 Application of Regression Analysis to Model Development

In general, the models developed here have high coefficients of determination and meet the F-test of significance at a significance level below 0.001. These results are not unexpected, because the development of the models is based on prior engineering knowledge of the primary relationships between performance, design, and cost. In this section, issues specifically related to the development of the regression models in this report are discussed. These issues relate to the number of observations available in each model data set, the use of transformation of variables to develop nonlinear models using linear regression, the selection of predictive parameters, the collection of data, and the reporting of results.

Number of Observations

The number of data points used to develop the regression model has an important effect on variable selection and interpretation of model results. As the number of data points becomes small, the number of independent variables that can be used may become constrained. It is often possible to obtain a model with a high coefficient of determination by

selecting a large number of independent parameters; however, such a model may contain counter-intuitive relationships, or relationships that violate principles of engineering. This often occurs when the range of a predictive variable is small, when other important predictive variables have not been included in the model, or when there is correlation or collinearity between predictive variables. It is often appropriate to include only a small set of independent parameters that are expected to be fundamentally important and robust as more data are gathered, rather to include all possible variables for which data are currently available. To select the most important parameters, one may begin by including all possible predictive variables in the model. Those variables with regression coefficients that fail the t-test for significance are then deleted to yield a new model with fewer predictive variables. The deletion or inclusion of a variable may be tempered by judgment regarding relationships that must be included in the model, assuming that the coefficients of the particular variable are of the correct sign and magnitude.

For small numbers of data, the estimates for the standard error, and the significance levels for the F-ratio, will tend to increase, because the degrees of freedom are reduced. Therefore, confidence intervals on the regression coefficients and the estimate for Y will usually be larger than when more data are available. As more data become available, the regression models can be redeveloped. While the specific values of the regression coefficients would likely change, they would be expected to remain within the confidence intervals, unless the new data are from a different sample population than the original data. In this case, the original regression model is not an appropriate representation of the new data. It is important, therefore, to ascertain if the basis for the new data is the same as for the older data (e.g., same design for process area equipment, same battery limits for the process area). In some cases, it may not be appropriate to add the new data without also including other predictive variables to capture the differences in the basis for the new and old data.

Transformation of Variables

While linear regression analysis has been used for all the regression model developed in this report, in many cases variable transformations have been used because the relationship between the dependent and predictive variables is non-linear. For example, the simplest cost model involves exponential scaling of a performance parameter to estimate cost, as presented in Equation (1). This functional form is standard in the chemical process industry, and cost capacity exponents for standard process plants are published in various sources (e.g., Peters and Timmerhaus, 1980; Ulrich, 1984; Humphreys and Wellman; 1987). The exponential scaling rule can be converted to linear form using the natural logarithm to transform the variables. A typical assumption for the functional form of the cost models is:

$$Y = aX_1^{b_1}X_2^{b_2}\dots X_k^{b_k} \quad (9)$$

This model represents the expected exponential scaling relationship between key process flowrates or design parameters and cost. In most cases, the exponent is expected to be less than one, representing the "economy of scale" of building larger units compared to smaller units. Typically, the exponent of one of the parameters will be much larger than for the other parameters. This result is expected, for example, when the flow rate of one material stream is expected to have a major influence in cost, while other parameters, such as temperature, may have only a secondary effect. The model in Equation (9) can be transformed to linear form using the natural logarithm:

$$\ln(Y) = \ln(a) + b_1\ln(X_1) + b_2\ln(X_2) + \dots + b_k\ln(X_k) \quad (10)$$

A linear regression is then developed based on the transformed variables. The transformation of variables affects the interpretation of distribution of the errors. If the errors for Equation (10) are normally distributed, which is the underlying assumption for the statistical tests discussed in the previous section, then the errors for Equation (9) will be lognormally distributed. The statistical tests are applied to the transformed model of Equation (10). These cases are noted in the text.

Recall that the probability density function (pdf) for the normal distribution is given by:

$$f(x) = \frac{1}{\sigma\sqrt{2\pi}} \exp\left(-\frac{[x - \mu]^2}{2\sigma^2}\right); \quad -\infty \leq x \leq \infty \quad (11)$$

where μ is the mean and σ is the standard deviation. If y is lognormally distributed, then $\ln(y)$ is normally distributed. The pdf for the lognormal distribution is given by:

$$f(y) = \frac{1}{\phi y \sqrt{2\pi}} \exp\left(-\frac{[\ln(y) - \xi]^2}{2\phi^2}\right); \quad 0 \leq y \leq \infty \quad (12)$$

The parameters of the lognormal distribution are ξ and ϕ . These parameters correspond directly to the mean and standard deviation of the normal distribution for $\ln(y)$. The mean and variance of the lognormal distribution are given by:

$$\mu_y = \exp\left(\xi + \frac{\phi^2}{2}\right) \quad (13)$$

$$\sigma_y^2 = \omega(\omega-1)\exp(2\xi) \quad (14)$$

where,

$$\omega = \exp(\phi^2)$$

Using these relationships, the parameters of the lognormal distribution of errors for the nonlinear regression models can be estimated from the parameters of the normal distribution for the errors of the linearized model. Therefore, the statistical model based on the functional form in Equation (9) is given by:

$$Y' = aX_1^{b_1}X_2^{b_2}\dots X_k^{b_k}\epsilon \quad (15)$$

where the error term is multiplicative and lognormal, not additive and normal as with the linear model in Equation (5). The mean of $\ln(\epsilon)$ is zero and the standard deviation is the standard error of the estimate for the linearized model. These parameters for $\ln(\epsilon)$ are used to estimate the mean and standard deviation for the lognormal distribution of ϵ using the relationships shown in Equations (12), (13), and (14). The median of the lognormal error term in Equation (15) will always be 1. The mean of the error term will typically be a value close to, but larger than, 1 and the standard deviation will typically be less than 1. The parameters that are reported for lognormal error terms in this report are the mean and standard deviation given by Equations (13) and (14).

Two-Step Regressions

In many cases, the relationship between cost and performance parameters is expected to be nonlinear, as described by Equation (15). However, the cost is also directly proportional to the number of trains of equipment for a given process area. To capture both the nonlinear relationships between performance and cost and the linear relationship between the number of trains and cost, a two-step approach to developing the regression models may be required. The primary reason for the two-step approach is because it is not possible to specify that the exponent of the number of trains must be equal to one when developing the nonlinear model. In the first step, a linearized regression of cost and performance parameters as just described is developed on the basis of a single train of equipment. In the second step, the predicted values from the nonlinear model for a single train are combined with information about the number of trains to predict the total cost of the process area. Thus, the final regression model from this process contains predictive variables for both performance and the number of total and operating trains.

The first step in the process involves estimating the coefficient and exponents of a model of the form of Equation (9) on the basis of a single train of equipment. The values of Y estimated in this fashion are then multiplied by the corresponding total number of trains to form a new predictive variable. This predictive variable is then used in a simple linear regression model. The first regression yields a model of the form:

$$Y^1 = a \left(\frac{X_1}{N_o} \right)^{b_1} \left(\frac{X_2}{N_o} \right)^{b_2} \dots \left(\frac{X_k}{N_o} \right)^{b_k} \quad (16)$$

Note that this is a general functional form; in some cases, the predictive performance parameters (such as temperature or pressure) do not depend on the number of trains, and therefore would not be divided by the number of operating trains. The estimated values of cost from Equation (16) represent the cost per operating train. However, we are ultimately interested in the total cost for the process area. Therefore, we calculate a new predictive variable which is the estimated cost for all operating and spare trains:

$$X^{(2)} = N_T Y^1 \quad (17)$$

We then use this new variable as the basis for a simple linear regression of the form:

$$Y' = a^{(2)} + b^{(2)} X^{(2)} + \varepsilon \quad (18)$$

Typically, the value of $b^{(2)}$ from this model is close to 1.0. The value of $a^{(2)}$ may occasionally be small enough (or statistically insignificant) to exclude from the model by estimating the regression without a constant. Note that the error term here is in the linear space. If the errors conform to a hypothesis of normality, then the error can be represented as normally distributed with a mean of zero. Based on Equations (16), (17), and (18), we can write the final regression model as:

$$Y' = a^{(2)} + b^{(2)} \left[N_T a \left(\frac{X_1}{N_o} \right)^{b_1} \left(\frac{X_2}{N_o} \right)^{b_2} \dots \left(\frac{X_k}{N_o} \right)^{b_k} \right] + \varepsilon \quad (19)$$

where the term within the square brackets is treated as a single variable in the simple linear regression. Thus, the first regression is essential a method for grouping a number of performance parameters into a single aggregate predictive term, while the second regression permits the addition of the linear relationship between cost and the number of trains of equipment. This approach permits the calculation of model statistics based on total, rather than per train, process area costs, which are the ultimate measures of interest.

Selection of Predictive Variables

Direct capital cost regression models for each IGCC plant section, and in some cases estimates of annual operating requirements, have been developed based on an analysis of approximately 30 detailed performance and cost studies of IGCC and coal-to-SNG (synthetic natural gas) systems. These models have been developed based on analysis of plant section direct costs and key plant section performance parameters. In each regression model, the parameters selected for inclusion in the model and the analytic relationships between model inputs and outputs were based on engineering judgments, statistical analysis, and data availability. These regression models relate the total direct cost (which includes delivered equipment cost, installation labor, and installation materials) to the statistically most significant performance parameters influencing cost. These parameters are typically mass flow rates, although in some cases parameters such as removal efficiency, pressure, or temperature were found to be statistically significant. In cases where parameters that are expected to be important were not found to be statistically significant, the variation in these parameters often is small for the available data samples (e.g., most KRW gasifier designs are at about 450 psig and 1850 °F), or the variation in these parameters is highly correlated with variations in the statistically most significant parameter (e.g., the syngas output from the KRW gasification section is highly correlated with the coal feed rate). Similarly, some parameters that are expected to be important in influencing cost may yield counter-intuitive results in the regression models (e.g., cost inversely proportional to mass flow rate). This, too, occurs when two parameters are highly correlated.

Collecting Data

Performance and cost data were collected into separate data bases for each plant section, based on similarity of plant section definitions. Only direct equipment costs were collected. Direct costs include equipment, material, and labor costs associated with installing plant equipment. Because indirect costs are treated differently in different studies (e.g., EPRI vs. GRI), these were not included in the cost data bases. All direct costs were adjusted to a common year using the Chemical Engineering plant cost index (January 1989 = 351.5). Because the studies varied in the amount of detail for each plant section, only a few performance parameters may be reported in common among studies for a given plant section. This limits the number of parameters that are candidates for regression analysis.

Reporting Results

For each plant section, the direct capital cost model is reported along with the error of the regression, the coefficient of determination, the number of data points used in developing the regression, and the range of values over which the regression was developed. The error

term is typically expressed as a normal distribution with a mean of zero and a standard deviation estimated from the difference between the direct costs available in the literature and the direct costs estimated from the regression model. In cases where a non-linear variable transformation was used, the error is reported as a lognormal distribution. The error term provides a measure of the variance of the direct cost estimate. In principle, the variance would be zero if the model accounted for all the parameters that influence costs and if the model were of an appropriate functional form. However, because the models are simplified and include only one or a few parameters, not all of the variation in cost is captured. The variance represents differences in plant location, design, or performance parameters that are not included in the cost model.

1.8 Representing Uncertainty in the Cost Models

The primary shortcoming of most engineering process models is that they do not facilitate the representation of uncertainty in key performance or cost parameters. A key feature of this research is the use of uncertainty analysis techniques as an integral part of model development and analysis.

New technologies are characterized by their early phase of development, for which limited test data and no commercial design or operating experience are available. Uncertainties in system performance at the commercial scale lead to uncertainties in capital and operating costs, which are the ultimate measures of interest for comparative cost-effectiveness analysis. Furthermore, even if process performance were known with certainty, uncertainties regarding the costs of equipment and reagents would remain. To explicitly characterize these uncertainties, and to evaluate the overall uncertainty in process costs, a probabilistic engineering modeling framework is needed for systematic analysis of uncertainty in evaluating alternative designs.

Decisions regarding process research and development priorities should be based on an understanding of the uncertainties in performance and cost estimates and the factors that contribute to them. While sensitivity analysis can be performed to examine the effect of variation in an uncertain input parameter on process performance and cost, typically only one or two parameters are varied at a time in a model containing perhaps hundreds of parameters. Thus, important interactions may be overlooked, and the number of cases that can be conveniently analyzed is extremely limited. A probabilistic capability permits an evaluation of interactions among many uncertain parameters, each of which is represented by a probability distribution. Probabilistic modeling thus provides estimates of the probability of different outcomes when many factors are varying simultaneously. Furthermore, analysis of model results permits identification of the input uncertainties that contribute most to output

uncertainty. Such a tool enhances the capability to assess technology performance, judge the economic viability of alternative designs, and target new research in areas that are likely to be most productive.

There are three general areas of uncertainty that should be explicitly reflected in engineering models. These are uncertainties in: (1) process performance parameters (e.g., flow rates), (2) process area capital costs, and (3) raw material and other system operating costs. For example, in calculating the cost of a gasifier, there may be uncertainty (because of the lack of commercial experience with the design) in the coal flow rate required to achieve a given electrical output. This leads to uncertainty in the size (hence, cost) of the gasifier for a particular system. However, for a given gasifier size and type, there is also a probability that the equipment cost could be higher or lower than the nominal estimate (e.g., due to expected improvements in equipment design and cost, or to potential problems with fouling and corrosion, requiring more expensive materials, design modifications, or additional maintenance). The same type of uncertainties may apply to operating and maintenance cost factors. The uncertainties associated with advanced systems or subsystems will typically be much larger than for conventional technology. A probabilistic engineering modeling framework is required to evaluate the overall uncertainty in process cost as a result of performance and cost uncertainties in specific process areas to determine the overall technical and cost risks and to identify research priorities.

1.8.1 Characterizing Uncertainties in Process Plants

For the cost models, it is convenient to calculate equipment cost using a cost correlation based on a size parameter, and then to use an uncertainty factor to explicitly characterize the uncertainty in the cost estimate aside from uncertainty in the size parameter. The general equation is:

$$C = f(x) \cdot u \quad (20)$$

where:

C	= cost of equipment
x	= size parameter(s) for the equipment
f(x)	= cost correlation (e.g., from regression analysis)
uf	= uncertainty factor for the cost estimate

For example, in calculating the cost of a gasifier, there may be uncertainties in coal flow rate, leading to uncertainties in the size of the gasifier. However, for a given gasifier size and type, there is an additional probability that the costs could be higher or lower than the nominal cost estimate. The uncertainty factor can represent judgments about possible improvements expected in equipment design and cost, or potential problems with fouling and corrosion

(which would be revealed by further research) requiring more expensive materials, design modifications, or more maintenance. Uncertainties in unit costs (e.g., labor hours, coal price) or financial parameters (e.g., discount rate) can be specified directly using the new ASPEN stochastic block developed concurrently with this work, which is discussed in Section 1.8.2.

A number of cost methods are commonly used in estimating the cost of electricity for utility plants. The most widely used of these is the Electric Power Research Institute (EPRI) approach (EPRI, 1986). The EPRI methodology will be used and modified as appropriate for probabilistic analysis, especially with regard to estimating contingency factors.

Generally, all cost methods involve estimating one or more contingency factors. A contingency is used to represent additional costs that are likely or expected to occur, but that are not included explicitly in the cost estimate. Typically, this is one of the largest items in a cost analysis.

It is important to distinguish between "project" contingency and "process" contingency. The project contingency is intended to cover the costs that would result from a more detailed design of a definitive project at a specific site. EPRI recommends that the project contingency be applied as a percentage of the total of process capital, engineering and home office fees, and process contingencies. EPRI suggests different ranges for the project contingency based on the type of cost estimate (e.g., simplified, preliminary, detailed, finalized).

The process contingency is intended to quantify the uncertainty in the technical performance and commercial scale cost of a technology, and should be reduced as a technology proceeds from bench scale to full commercial use. EPRI recommends ranges of project contingency based on the state of technology development (e.g., new concept, bench-scale data, pilot plant, full scale demonstration, commercial) for each major plant section. However, both the project and process contingency values appear to be merely "rule-of-thumb" recommendations.

GRI requires an estimate of "process development allowances" (PDA) for all major plant sections (GRI, 1983). The PDA is intended to account for increases in cost as the design definition of a new technology is increased, revealing additional equipment required for a commercial-scale plant, and as a technology proceeds from early stages of development through commercialization. The PDA is an average of assessments, on a percentage of direct cost basis, of expected cost increases based on the state of the technology, the availability of experimental data, assumptions in the performance and cost estimate that have not been tested, and expected difficulty of control and operation. For each of these categories, the GRI

Guidelines suggest a number of areas to consider when developing the assessments. The PDA is similar to the EPRI definition of process contingency.

The Rand Corporation conducted a survey of 18 companies in the chemical and petroleum industries to determine the actual methods used to develop contingency factors (Milanese, 1987). The study indicates that contingency factors are badly underestimated, which may be leading to bad decisions about certain projects. Rand recommends greater and more formalized use of experience and the inclusion of costs associated with risks and innovation into the cost estimates.

A probabilistic modeling approach supplants the traditional contingency factor approach by incorporating expert knowledge about uncertainties and quantifying the uncertainty in performance and cost associated with risky or innovative technologies. Given uncertainties in process performance (e.g., flow rates, temperatures, pressures) and in costs (e.g., equipment, reagents, numbers of operating and spare trains), the resulting distribution in the capital cost estimate can be simulated through a probabilistic sampling technique (e.g., Latin hypercube sampling).

Because the probabilistic modeling approach is based on a disaggregated consideration of the factors that influence the technical and cost risks of a project, a more accurate representation of costs can be expected compared to deterministic estimates with "rule-of-thumb" contingency factors. A probabilistic modeling result can be linked to a deterministic cost estimate by appropriate selection of the contingency factor. However, a contingency factor is not required in a probabilistic estimate if all the uncertainties that contribute to a contingency allocation are treated probabilistically in the model. Typical examples of uncertainties that contribute to process contingency are stoichiometric ratios, stream flow rates, temperatures, regeneration efficiencies, the number of operating and spare trains, and unit equipment costs, all of which can be assigned either discrete or continuous probability distributions.

In this study, the suggested values of process contingency factors and process development allowances are summarized to provide an indication of the uncertainty in capital costs. Also, uncertainties in annual cost parameters are summarized, where these data are reported.

1.8.2 Probabilistic Modeling Capability in ASPEN

As part of DOE-supported research at Carnegie-Mellon University, a probabilistic modeling capability has been added to the publicly available version of ASPEN (Diwekar, Rubin, and Frey, 1989). An initial step in this effort was the identification of suitable software

for sampling probability distributions and performing output analysis. A Fortran program developed by Iman and Shortencarier (1984) using Latin hypercube sampling (LHS) was identified as the best publicly available program for assigning probability distributions to model parameters and generating samples from those distributions. A Fortran program developed by Iman, Shortencarier, and Johnson (1985) was identified for analysis of model output. This program uses partial correlation coefficients and standardized regression coefficients for measuring linear correlations and partial rank correlation coefficients and standardized rank regression coefficients for measuring nonlinear correlations.

In random Monte Carlo simulation, a random number generator is used to take samples from the inverse cumulative probability distribution of a variate. In LHS methods, a probability distribution is first divided into ranges of equal probability, and one sample is taken from each equal probability range in random order. In random LHS, one sample is taken from random within each interval. LHS methods guarantee that values from the entire range of the distribution will be sampled proportional to the probability density of the distribution. Because the distributions are sampled over the entire range of probable values in LHS, the number of samples required to adequately represent a distribution is less for LHS than for random Monte Carlo sampling.

A partial correlation coefficient (PCC) analysis is used to identify the degree to which correlations between an output and an input variate may be linear. Standardized regression coefficients (SRC) can be used to measure the relative contribution of the uncertainty in the input variables on the uncertainty of the output variables. This analysis involves standardization of all distributions and a multi-variate regression of an output variate based on the input variates. The regression coefficients for each input variate then indicate the relative importance of that variate as a factor determining the output. A PCC measures the unshared contribution of each input variate on the output, whereas SRC measure the shared contribution of the input to the output. These measures are related, but do not always lead to the same conclusions. PCC and SRC analysis is limited to cases where the relationship between input and output variables is linear; however, by basing the regression analysis on the ranks of the samples for each variable, rather than on the values of the samples, the PCC and SRC techniques can be extended to non-linear cases. These techniques are known as partial rank correlation coefficients (PRCC) and standardized rank regression coefficients (SRRC) (Iman, Shortencarier, and Johnson, 1985).

The LHS sampling and output analysis programs have been implemented into ASPEN through a new unit operation block which is documented in a new user's manual (Diwekar, 1989) and technical reference manual (Diwekar and Rubin, 1989). The stochastic block assigns user-specified distributions to the key input parameters selected by the user, uses the

LHS program to generate samples from the distributions, and passes the sampled values of each uncertain parameter to the flowsheet. After a flowsheet simulation is run, the output variables of interest are collected. The simulation is then repeated for a new set of samples selected from the probabilistic input distributions. A new Fortran block is used to control the cycling of the stochastic block, and another Fortran block is used to access and assign samples to model parameters.

The probabilistic modeling capability in ASPEN has both Monte Carlo and random LHS options. There is also a capability to specify rank order correlations in input variables.

1.9 Scope and Organization of This Report

The 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. For example, it is desirable to increase the level of detail of the cost model for the gasification section of each technology, to reflect additional parameters not now incorporated. Additional detail would be required for specific process equipment units, but such information is not available publicly at this time. Thus, initial estimates of uncertainty are based on the standard error of the regression models and the contingency factors assumed in the performance and cost studies used for model development.

This report includes separate chapters describing the models developed for the direct capital cost of each type of IGCC system. Chapter 5 then describes how the total capital cost of each system is calculated. Chapter 6 describes annual operating cost calculations for each system. Chapter 7 describes the total annualized cost model used to calculate the cost of electricity. Chapter 8 presents a summary of the models and describes implementation of the cost models with the ASPEN performance models, sample results, model validation, priorities for further cost model development, and recommendations for further work. The computer code for the newly implemented cost models and sample outputs from each model are given in the appendices.

2.0 CAPITAL COST OF A KRW-BASED SYSTEM WITH COLD GAS CLEANUP

Figure 5 shows an oxygen-blown KRW IGCC power plant with cold gas cleanup. The system shown in the figure is based on the design presented in a study for EPRI (Fluor, 1985) and the configuration of the ASPEN flowsheet of a KRW-based IGCC system with cold gas cleanup (Stone, 1985). In the gasification section, coal is partially oxidized in a reaction with oxygen and steam to produce a gas containing CO and H₂. Oxygen for the gasification reaction is provided by an air separation plant, and steam is provided from the plant steam cycle. The hot fuel gas leaving the gasifier is cooled by generating saturated steam in high temperature heat exchangers as part of the IGCC power plant steam cycle. Particulates in the fuel gas are removed by cyclones, which recycle fines to the gasifier, and by a particulate scrubber. The sulfur in the fuel gas, primarily H₂S, is removed in the acid gas removal section using the Selexol process, and the off-gas from this section, containing concentrated H₂S, is sent to a Claus plant for sulfur recovery. Prior to entering the acid gas removal section, the coal syngas must be cooled to approximately 100 °F. The clean fuel gas is reheated (not shown in the figure) and then combusted in gas turbines, which generate electric power. The hot exhaust gas from the gas turbines is used to superheat saturated steam from the gasification section and to reheat steam for the steam turbine, which also generates electricity. A portion of the electrical output from the generators must be used to power equipment in the plant, most notably the air separation plant. The remaining electricity is exported for sale. Overall, the plant consumes coal, air, and water, and produces ash, sulfur, flue gas, wastewater, and electricity. Although many individual components of IGCC systems are commercially proven in other applications, there is only limited experience with some specific components.

The IGCC plant is divided into 13 sections for the purpose of estimating total direct capital costs. These sections include the following:

1. Coal handling (Area 20)
2. Oxidant feed (Area 10)
3. Gasification (Area 30)
4. Gas cooling (Area 40)
5. Acid gas removal (Area 50)
6. Sulfur recovery (Area 60)
7. Tail gas treating (Area 70)
8. Boiler feedwater system (Area 80)
9. Process condensate treatment (Area 85)
10. Gas turbine (Area 91)
11. Heat recovery steam generators (HRSG) (Area 92)
12. Steam turbine (Area 93)
13. General facilities (Area 100)

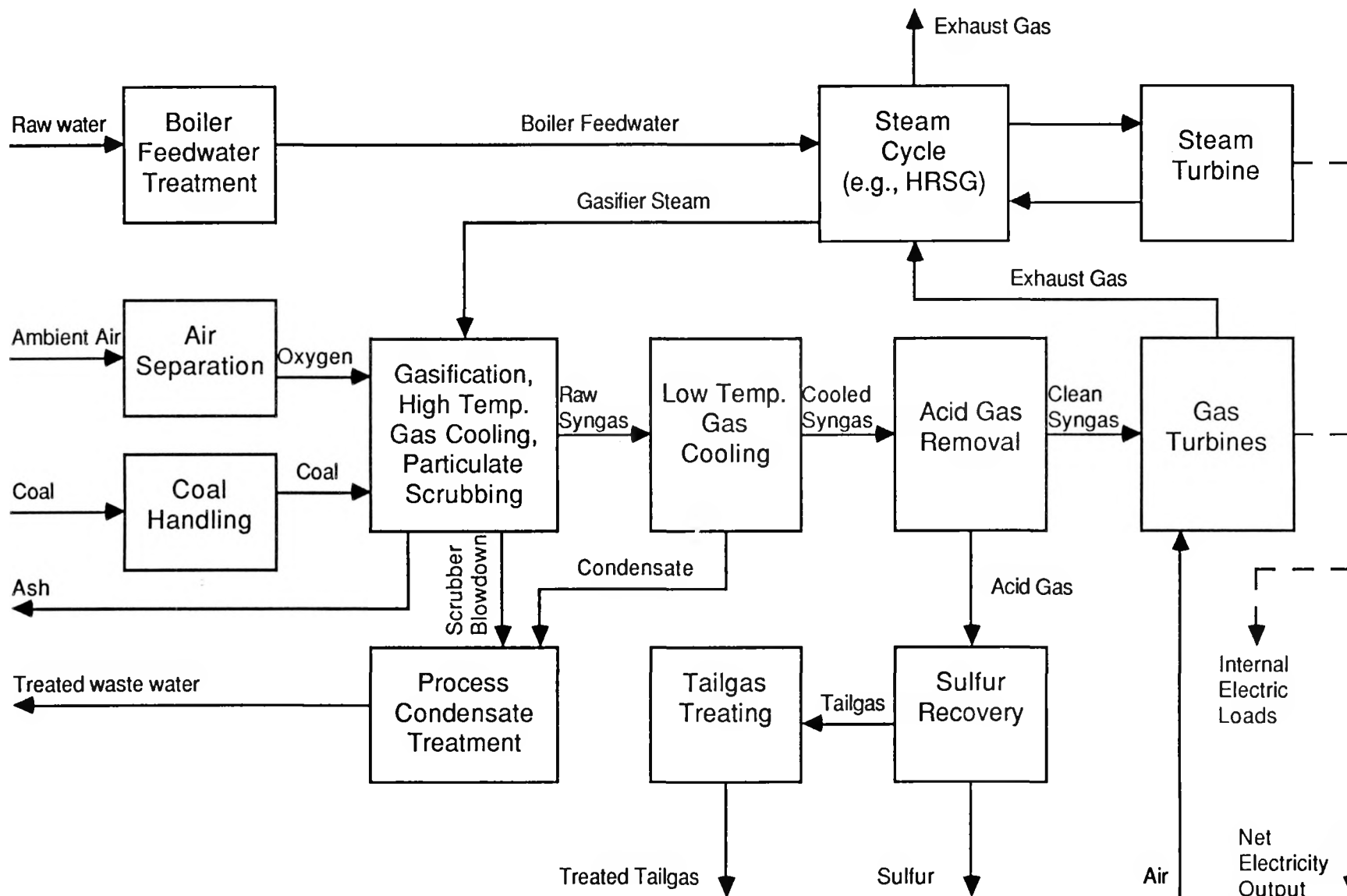


Figure 5. Simple Schematic of Oxygen-Blown KRW IGCC System with Cold Gas Cleanup

The direct cost correlations for each plant section are described in the following sections of this report. While some of the process area models may be applicable to a variety of IGCC or coal-to-SNG systems, the models are intended for the specific purpose of estimating the direct cost of the KRW-based IGCC system shown in Figure 5. The reader is cautioned that the models should not be used to estimate costs for systems with a design basis that differs from the assumptions discussed here. Nor, except when noted, should the models be extrapolated. The most salient features of the design basis for each process area are discussed in this report. However, the purpose here is not to recapitulate each detail of the process area design basis, but rather to document the development of the cost models. Therefore, the reader may wish to read this report in conjunction with some of the performance and cost studies cited here to obtain more detail about specific process areas. Perhaps the single most useful reference is the EPRI-sponsored study by Fluor (Fluor, 1985).

2.1 Coal Handling Section

The coal handling section for a KRW-based IGCC system must deliver dry pulverized coal to the coal surge bin in the gasification section. The design basis for the coal handling system assumed here is specific to KRW-based systems; therefore, the coal handling cost model should not be applied to other types of gasification systems. However, the KRW-based coal handling is generic to both IGCC and coal-to-SNG systems; therefore, data from conceptual design studies of both types of systems were used to develop the direct capital cost model. These studies include: Bechtel and WE, 1983; Fluor, 1985; Cover et al, 1985a; Smith et al, 1986; Smith and Smelser, 1987; and Earley and Smelser, 1988a. The first two of these studies are for IGCC systems, while the others are for coal-to-SNG systems. An additional study by Earley and Smelser (1988b) develops costs for a KRW-based IGCC system. This study uses the same coal handling design and cost estimate as Earley and Smelser (1988a).

The coal types represented in the data set include Illinois No. 6, Pittsburgh No. 8, Wyodak subbituminous, Texas lignite, and North Dakota lignite. The design basis for the coal handling section is similar across the KRW-based design studies. 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. There is typically one train of equipment for coal receiving, and multiple trains for coal crushing and drying. However, detailed information on the size or number of trains of process area equipment is

limited, particularly for the coal-to-SNG studies. In addition, some studies report only a total direct cost for coal receiving, crushing, and drying.

The coal throughput determines the sizing of most of the process area equipment. Therefore, this is the primary predictive parameter for direct cost. While other parameters, such as coal moisture content, also may affect cost, they are only secondary in importance. Only 15 data points for direct cost and coal mass throughput are available from the design studies of KRW-based systems. There are fewer than 15 data points for potential secondary predictive parameters, except for coal moisture content. However, there is a high correlation between coal moisture content and mass flow rate for this particular set of data. Because of these limitations in the available data, and because it is desirable to develop the regression model based on all of the available cost sample points, a single variable regression equation based on coal feed rate to the gasifier was developed. Both linear and nonlinear functional forms were investigated. A linear model yielded marginally better summary statistics (e.g., coefficient of determination, F-ratio, t-statistics for individual parameters) than a nonlinear exponential scaling model. About half of the data points are drawn from one study in which the cost of coal handling appears to scale linearly with coal mass flow rate. Therefore, a linear model for the direct cost of the coal handling process area as a function of the coal mass flow delivered to the gasifier is assumed:

$$DC_{CH} = 3,300 + 4.09 m_{cf,G,i} \quad (R^2 = 0.859; n = 15) \quad (21)$$

where,

$$4,700 \leq m_{cf,G,i} \leq 23,000 \text{ tons/day}$$

The range of values for the coal mass flow indicates the limits within which the regression model should be applied. The model should not be used to predict coal handling direct cost for coal mass flow rates significantly outside this range. The standard error of the regression model is \$10 million. Based on only these 15 data points, it is not possible to reject the hypothesis that the error is normally distributed at a significance level of 0.05 using a one-sample Kolmogorov-Smirnov test. The coal handling section cost data and regression model are shown graphically in Figure 6. From the graph, it is apparent that the errors may not be normally distributed. Therefore, summary statistics for the regression model should be interpreted with care. However, the regression model does capture the key relationship between coal mass flow rate and coal handling system direct capital cost.

Figure 6. Direct Cost for Coal Handling Section

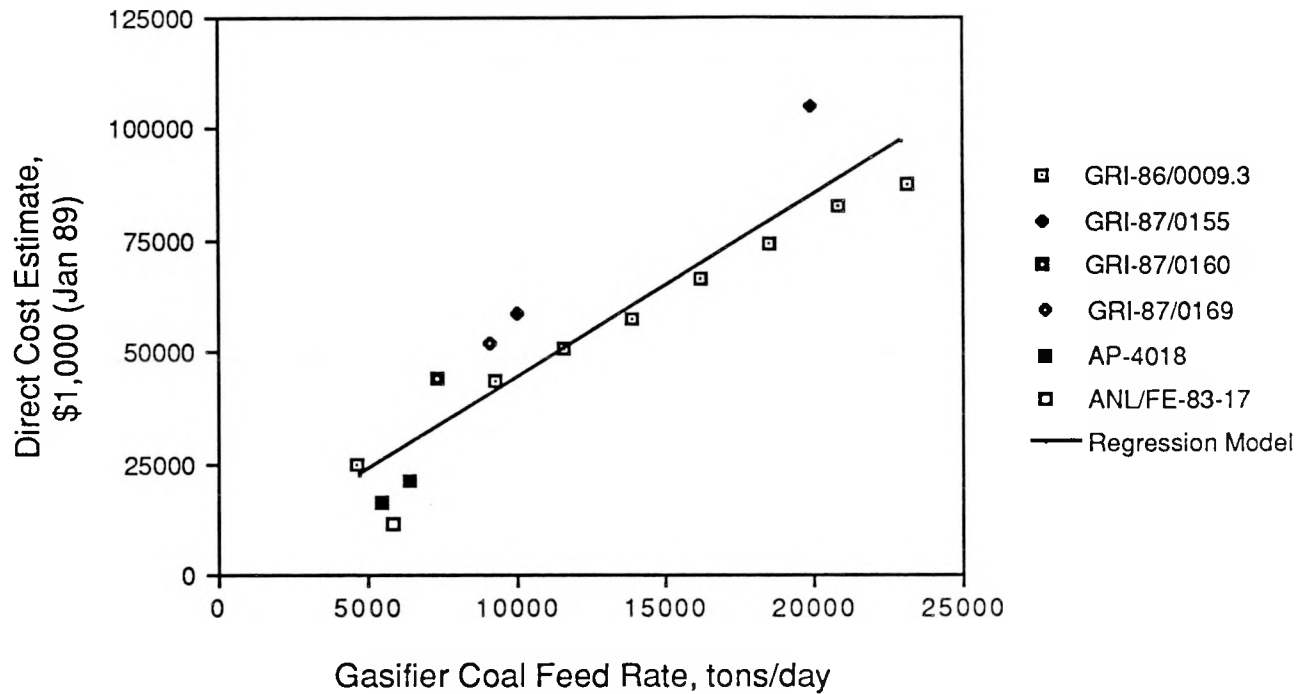
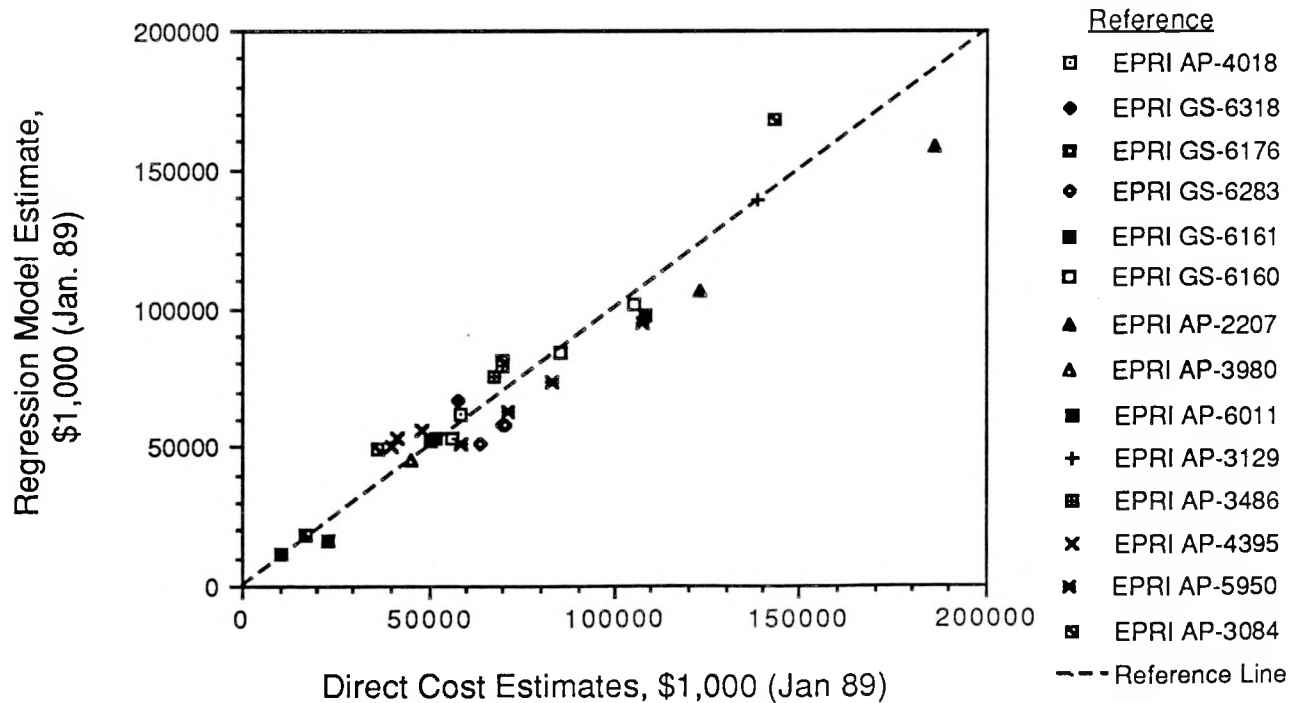


Figure 7. Predicted vs. Actual Direct Costs for Oxidant Feed Section



2.2 Oxidant Feed

The oxidant feed section is applicable to all oxygen-blown gasification systems. Performance and cost data for 31 oxygen plants were taken from 14 studies of oxygen-blown IGCC systems, all prepared for EPRI. These plants all include electric motor-driven compressors. Data from coal-to-SNG systems were not included because many of these use steam-driven, rather than motor-driven, compressors. Electric motor-driven systems offer advantages in terms of plant operation, although steam-driven systems may be more energy efficient. These plants produced between 625 and 11,350 lbmole/hr of oxygen per train. A typical plant consists of two parallel operating trains with no spare trains. Each train includes an air compression system, air-separation unit and an oxygen compression system. For more detail on the oxygen plant design, see Fluor (1985). The oxygen plants produce an oxidant feed to the gasifier containing typically 95 to 98 percent oxygen on a volume basis. It is possible to recover argon as a saleable byproduct from high purity oxygen plants (99.5 percent oxygen or greater) (BOC Cryoplants, 1987); however, the available data are not for the oxygen purity levels and plant designs required to do this. The oxygen plants represented in the database are considered commercially available.

The direct cost of oxygen plants is expected to depend mostly on the oxygen feed rate to the gasifiers, because the size and cost of compressors and the air separation systems are proportional to this flow rate. The oxygen purity of the oxidant feed stream is expected to affect the cost of the air separation system. As oxygen purity increases, it is expected that the cost of the oxygen plant will increase because the size of equipment in the air separation plant (e.g., high pressure column) increases (BOC Cryoplants, 1987). The ambient temperature determines the volume flow rate of air entering the inlet air compressor; as ambient temperature increases, the volume flow rate increases for a given mass flow, thereby requiring an increased compressor size.

A number of regression models were considered in which alternative combinations of predictive parameters and functional forms were assumed. These regressions were based on nonlinear variable transformations using the natural logarithm, as discussed in Section 1.7.2. A single-variate regression of cost and oxygen flow rate, using an exponential scaling formulation (see Equation (1) for an example), not unexpectedly yielded excellent results ($R^2 = 0.90$). The scaling exponent in this case was 0.9. The addition of terms for ambient temperature and oxidant purity yielded a marginal improvement in the summary statistics for the model. From an engineering viewpoint, the inclusion of these additional predictive terms significantly improves the utility of the model, allowing costs to be sensitive to both primary and

secondary factors. Therefore, the following multi-variate regression is assumed for the oxidant feed process area direct capital cost:

$$DC_{OF} = 14.35 \frac{N_{T,OF} T_a^{0.067}}{(1 - \eta_{Ox})^{0.073}} \left(\frac{M_{O,G,i}}{N_{O,OF}} \right)^{0.852} \quad (R^2 = 0.936; n = 31) \quad (22)$$

where,

$$20 \leq T_a \leq 95;$$

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

$$0.95 \leq \eta_{Ox} \leq 0.98.$$

The robustness of the exponential scaling relationship between oxygen flow rate and direct capital cost is indicated by the similarity of the exponent for oxygen flow rate in the single and multi-variable regression models. In the single variable model previously described, the exponent was 0.9, while for the multivariate model above it is 0.85. The limits for each parameter indicated above represent the ranges for which the regression model is valid. While to obtain accurate results these ranges should not be violated, it is not a severe violation to exceed the range for the oxygen flow rate per train, particularly on the high side, because the model reasonably captures the expected relationship between oxygen flow rate and cost. An alternative to extrapolating the model for oxygen flow rate per train, however, is to alter the number of trains so that the flow rate per train is within the limits given above. The ambient temperature and oxygen purity parameters should not be extrapolated.

Regression of a linearized model (see Equation (10)) including a term for the number of trains was possible for this process area because the estimated exponent for the total number of trains was approximately 1.0. This result is expected because the cost of the oxidant feed plant should be a linear function of the number of trains. A one-sample Kolmogorov-Smirnov test indicated that the residual error of the linearized model is reasonably similar to a normal distribution. In the nonlinear form reported here, the error is therefore lognormally distributed, as discussed in Section 1.7.2. The mean of the lognormal error is 1.012 and the standard deviation is 0.155. The 90 percent probability range for this distribution is 0.78 to 1.29, implying a 90 percent confidence limit from 78 to 129 percent of the predicted values from the model.

A graphical representation of the model is shown in Figure 7 on page 54. Figure 7 shows a comparison of the direct costs estimated by the regression model versus the costs available in the engineering studies. The dashed line in the figure represents the ideal case where the model would predict the study costs with no error. In the cases where a point lies above the reference line, the model over-predicts costs, and in the cases where a point lies below the reference line, the model under-predicts costs. The largest error in prediction occurs for two cases at the upper end of the cost range. These two cases represent designs with several trains (four in one case, five in the other) and costs based on older studies. If these two cases are excluded from the estimate of the model error, the standard deviation of the residual errors is reduced from \$10.8 million to \$8.7 million.

2.3 Gasification, Ash Removal, High Temperature Gas Cooling, and Particulate Removal

For a KRW-based system, the gasification section includes equipment associated with coal pressurization, gasification, char recycle, syngas recycle, ash removal, high temperature gas cooling, and particulate removal from the syngas.

2.3.1 Technology Overview

The KRW gasifier is a pressurized fluidized bed, agglomerating dry ash design. The gasifier vessel is a refractory lined carbon steel pressure vessel. The gasifier has a number of different "zones," including low velocity ash cooling and removal, moderate velocity ash separation and gasification, and high velocity combustion and devolatilization. In addition, a freeboard disengaging zone is used to reduce the amount of char entrained in the outgoing syngas. The diameter of the gasifier is largest in the freeboard zone, in order to reduce syngas velocity and entrainment of char. Char leaving the gasifier is removed from the syngas by cyclones and recycled to the gasifier to improve overall carbon conversion efficiency (Smith et al, 1986).

No commercial scale KRW gasification system has yet been built. Kellogg Rust Synfuels, Inc. (KRSI) has operated a 24 ton coal per day process development unit (PDU) test facility in Waltz Mill, PA since 1975. By contrast, typical performance and cost studies of commercial scale gasifiers assume an average of about 1,000 tons per day of coal feed, which implies a scale up of a factor of 40. Furthermore, the test facility is operated at a maximum pressure of 230 psig, compared to the typically assumed value of 450 psig in design studies. The PDU has higher heat losses, higher fines elutriation, higher recycle gas flow rates, higher ash annulus gas velocities, and a lower bed height than most design studies. Furthermore, the carbon conversion rates and oxygen-to-coal and steam-to-coal ratios assumed in most design studies may not be justifiable based on PDU experience, because the PDU has

operated at significantly higher steam-to-oxygen ratios than commonly assumed (Shinnar, Avidan, and Weng, 1988).

The basic processes involved in gasification are chemical, thermal, and hydrodynamic. Many of the chemical processes do not depend on scale because they take place at a particle level. However, hydrodynamic processes are generally scale-dependent, and influence the thermal history of particles and gases in the gasifier. The main areas of design that have the least commercial experience are the combustion jet zone and the ash separation zone. Analytic models of the jetting and ash separation zone have been developed by KRW based on a variety of tests at different scales, ranging from four inch to ten foot diameter for cold flow facilities and four inch to 24 inch diameter for hot flow facilities. KRSI claims an excellent correlation between the analytic models and the observed test results (Smith et al, 1986).

Perhaps the most significant scale-up uncertainty is in the jet combustion zone. It is important that the jet surface area and the solid recirculation rate near the jet be sufficient to allow for dissipation of the heat from combustion, otherwise agglomeration, clinkering and sintering of bed material will occur. Commercial designs may require the use of multiple jets, rather than a single jet, for better distribution of heat. However, this alternative may introduce problems if the jet velocities are not uniform. Also, multiple jets may interact to form stagnant regions between jets. KRSI recommends more extensive testing using semi-circular and circular models of multiple jets before designing a commercial reactor with multiple jets.

Gasifier performance for a specific coal is predicted by M.W. Kellogg based on analytic models and empirically-derived data. Experimental data are required to determine reaction rates and the influence of contained mineral matter on coal reactivity. The caking properties of the coal are also important, as is the ash fusion temperature. M.W. Kellogg has devised a number of bench-scale tests that are used to determine the empirical data needed for the analytic models. The combination of bench scale testing and mathematical modeling is reported to yield a good predictive capability for gasifier performance. However, the predictive capability is limited by uncertainties in free board temperature, bed density, bed carbon content, and bed height (Floyd and Agrawal, 1989).

Six performance and cost studies of KRW-based systems were used to develop the gasification section cost correlation. These include two studies of IGCC systems (Bechtel and WE, 1983; Fluor, 1985) and four studies of coal-to-SNG systems (Cover et al, 1985a; Smith et al, 1986; Smith and Smelser, 1987; Earley and Smelser, 1988a). An additional earlier study (Bostwick et al, 1981) was reviewed for possible inclusion in the cost model database, but was

excluded because the cost estimate was unusually low compared to the other data and because information on the performance of the system was limited.

In all six studies, the gasification section definitions are similar. The coal-to-SNG systems are the same as the IGCC systems in the areas of coal pressurization, gasification, ash removal, and particulate scrubbing. In fact, it appears that the physical dimensions of the gasifier are similar across all studies. Typical dimensions are an overall height of about 100 to 115 ft, a maximum outer diameter of about 14 feet, and a minimum outer diameter of about 5.5 ft. All systems use a coal surge bin, coal pressurization lockhopper, coal feed lockhopper, and rotary feed valve to deliver pressurized coal to the gasifier. All systems also use an ash receiving lockhopper and an ash depressurization lock hopper for ash removal. All systems use pneumatic transport of coal from the rotary feed valve to the gasifier, and pneumatic cooling and separation of ash. All systems have a recycle gas compressor and motor. Particulate recovery appears to be the same for the coal-to-SNG as for the IGCC cases, although relatively little detail is provided on this part of the plant section. A typical particulate recovery system includes a venturi scrubber and a gas scrubber. Dirty wash water is filtered in a filtration system, and the filter cake is sent to disposal. Blowdown from the wash water and sour condensate from knock-out drums are sent to the process condensate treatment section of the plant.

There appear to be some differences in the area of high temperature gas cooling and char recycle. For example, the number of high temperature heat exchangers appears to vary from one to two in the coal-to-SNG systems. With the exception of the most recent KRW-based IGCC study (Fluor, 1985), all the systems produced high pressure superheated steam from high temperature syngas cooling. The IGCC systems include a fuel gas reheater that is not used in the coal-to-SNG systems. In general, the coal-to-SNG systems appear to use only one cyclone for char recycle, compared to two cyclones in the IGCC cases.

A data base of performance and cost information for a total of 15 different plant sizes from the six references was developed. Of these 15 cases, seven are a variation in size from one case (Cover et al, 1985a) and one is a variation in size for another case (Smith et al, 1986); detailed performance data are not available for these scaled cases. For all 15 cases, the data base includes the number of spare and operating gasifier trains, number of particulate scrubbing units, gasifier operating temperature, gasifier operating pressure, coal feed rate to the gasifier, oxidant feed rate to the gasifier, and total direct capital cost. For seven of the cases, more extensive data are available. These data include steam feed rate to the gasifier, boiler feed water to the gasification section, raw water make up required for particulate scrubbing, outlet syngas from the gasification section, ash disposal, high pressure

steam production from gas cooling, sour water output to process condensate treatment, and particulate cake to disposal. The 15 cases in the database include Illinois No. 6, Pittsburgh No. 8, Texas lignite, North Dakota lignite, and Wyodak subbituminous coal as feedstocks, with gasifier coal feed moisture contents ranging from about 1 to 23 percent.

2.3.2 Factors Affecting Cost

The cost of the gasification section is expected to depend on factors such as the coal feed rate, oxygen feed rate, steam feed rate, syngas flow rate, high pressure steam flow rate from the heat recovery system, ash disposal flow rate, raw water makeup for the particulate scrubbing unit, sour water flow rate to process condensate treatment, operating temperatures and operating pressures. While data on all of these parameters is available for some of the cost studies, data are not available for all the cost studies, such as for sour water flow rates. In some cases, such as gasifier operating temperature and pressure, there is little variation from one study to another. In other cases, such as with coal feed rate and syngas flow rate, there is a high correlation between parameters, so that inclusion of both in a regression equation does not significantly improve the "goodness-of-fit" and may yield unexpected results (e.g., cost inversely proportional to syngas flow rate).

2.3.3 Model Form

Prior to investigating the development of regression models, an attempt was made to develop a detailed cost model for the gasification section based on estimating the equipment cost of each major equipment item within the process area. This approach requires detailed information regarding the design basis of each equipment item and the performance parameters that determine equipment size. For example, the cost of a gasifier vessel would be estimated using the shell weight of the gasifier vessel, which is calculated based on the dimensions of the gasifier. The dimensions of the gasifier would be determined from detailed information regarding the ash separation zone, the coal combustion jet zone, the fluidized bed reaction zone, and the transport disengaging zone. The shell weight is multiplied by a unit material cost, and the costs of refractory lining, manholes, and installation would then be developed. Because the gasifier is a vertical pressure vessel, the installation costs must account for the difficulty of erecting the large vessel at the site. However, this approach proved unsatisfactory for several reasons. The required detailed design information for the KRW gasifier is not reported in the cost studies. In addition, detailed design information is not reported for other components of the gasification process area. Finally, because the direct costs are reported only for the process area as a whole, there would be no direct way to validate the cost of each individual equipment item. Therefore, a regression model approach was used. The advantage of the regression model is that, if it is not extrapolated, it will yield cost estimates consistent with the published studies.

Several alternative regression models were investigated. These models included single variable regressions on parameters such as coal feed rate (on as-received, dried, or moisture- and ash-free bases) or syngas flow rate, and multivariate regressions based on various combinations of two or more predictive parameters such as coal feed rate, syngas flow rate, oxidant feed rate, ash removal rate, and number of trains. From these analyses, a regression model of gasification section costs was developed (using "Cricket Statworks" on a Macintosh computer) based on three performance parameters, the number of operating trains, and the number of total trains. This model was developed using the two-step approach described in Section 1.7.2. The performance parameters selected are coal feed rate (moisture- and ash-free basis), oxidant feed rate, and ash removal rate. This combination of parameters yielded reasonable results for the scaling of gasification section costs to mass flow rates. Of these three performance parameters, the model is least sensitive to the ash removal rate, as is expected. The coal feed rate directly affects the size of the coal pressurization and gasifier systems, and the combination of coal feed rate and oxidant feed rate influences the syngas flow rate, which in turn is a primary sizing parameter for high temperature gas cooling and the particulate scrubbing system. The regression model is given as:

$$DC_G = 27,700 + 0.0184 N_{T,G} \left(\frac{m_{cm,G,i}}{N_{O,G}} \right)^{0.867} \left(\frac{M_{Ox,G,i}}{N_{O,G}} \right)^{0.987} \left(\frac{m_{ash,G,o}}{N_{O,G}} \right)^{0.083} \quad (23)$$

$$(R^2 = 0.948; n = 15)$$

where,

$$800 \leq \left(\frac{m_{cm,G,i}}{N_{O,G}} \right) \leq 1,250 \text{ tons/day};$$

$$1,450 \leq \left(\frac{M_{Ox,G,i}}{N_{O,G}} \right) \leq 2,000 \text{ lbmole/hr};$$

$$50 \leq \left(\frac{m_{ash}}{N_{O,G}} \right) \leq 200 \text{ tons/day}.$$

The model should not be extrapolated for values of the predictive parameters outside the ranges given above. These ranges indicate the allowable size for each gasification train. The standard error of the estimate is \$20.5 million. A graph of the cost data and the regression model is shown in Figure 8. The model should only be used to estimate the cost of oxygen-blown KRW gasification systems operating at 465 psia and between 1,575 and 1,875 °F.

Figure 8. Predicted vs. Actual Direct Costs for Gasification Section

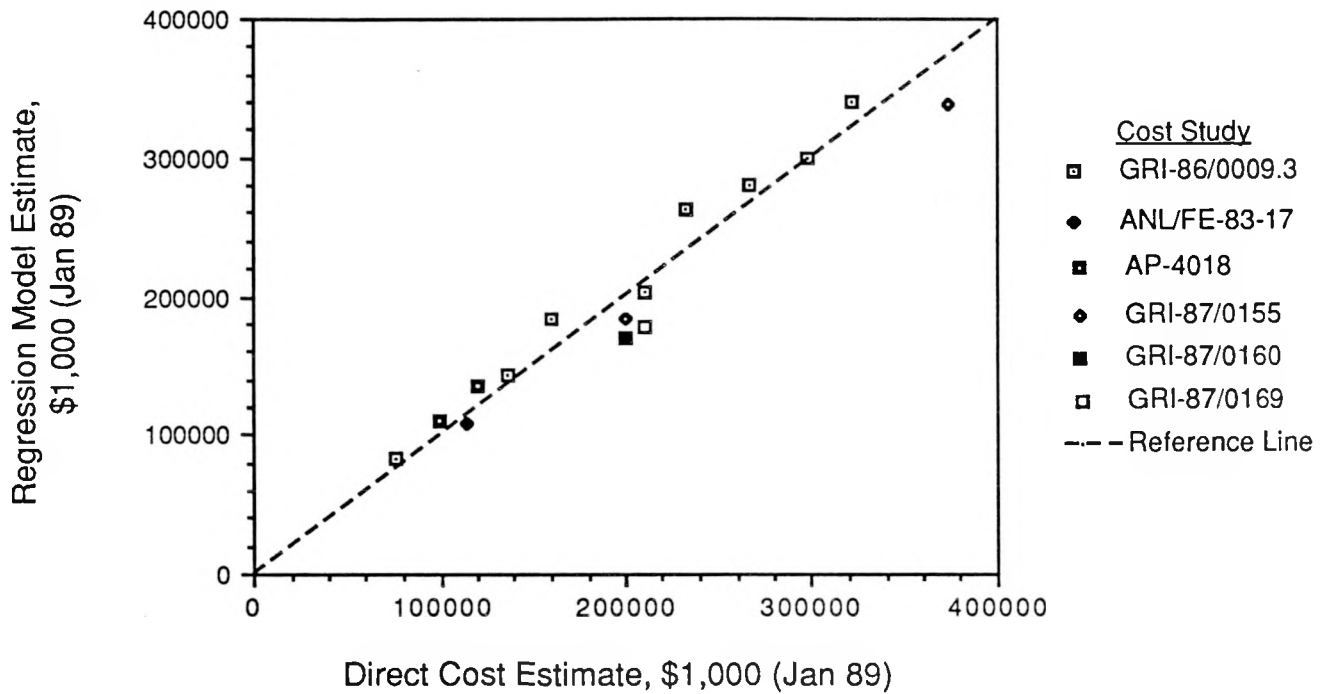
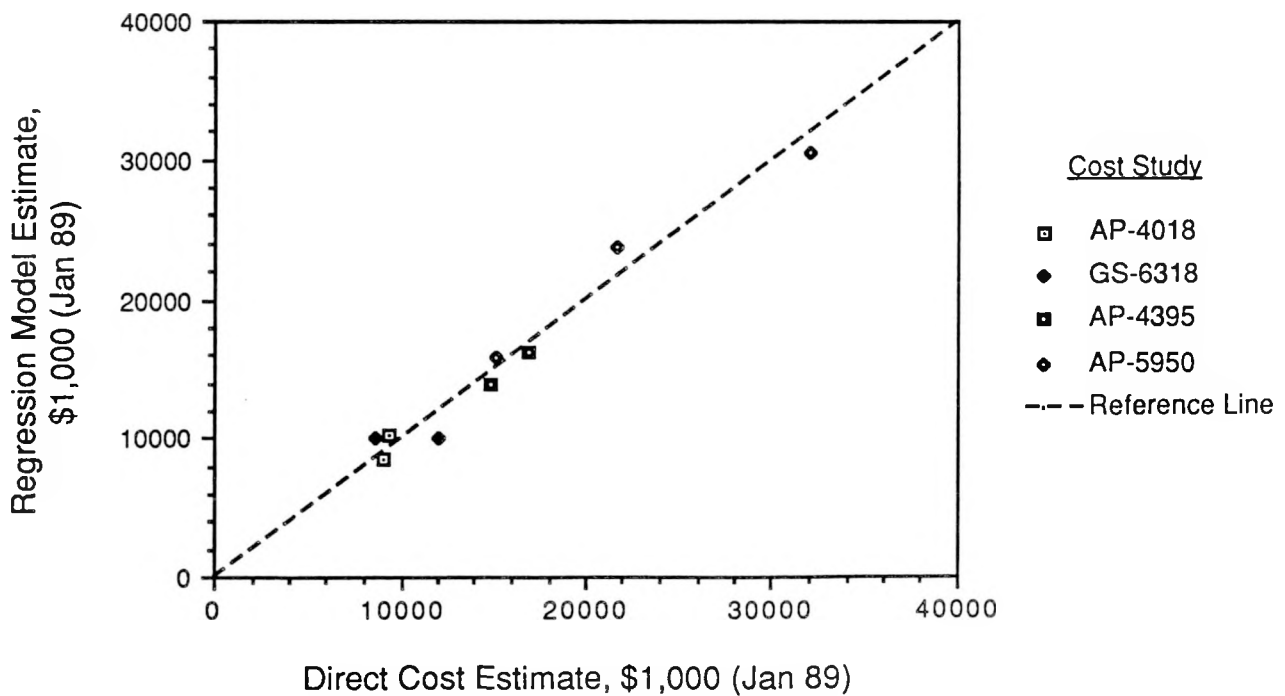


Figure 9. Predicted vs. Actual Direct Costs for Gas Cooling Section



The number of trains used in the gasification section can be estimated from the following correlations:

$$N_{T,G} = \text{INT} \left(0.556 + 1.049 \times 10^{-3} m_{cm,G,i} \right) \quad (R^2 = 0.944; n = 15) \quad (24)$$

$$N_{O,G} = \text{INT} \left(0.557 + 9.10 \times 10^{-4} m_{cm,G,i} \right) \quad (R^2 = 0.936; n = 15) \quad (25)$$

where INT means the nearest integer value. The number of spare trains can be estimated from the difference between the number of total and number of operating trains.

2.4 Low Temperature Gas Cooling and Fuel Gas Saturation

In IGCC systems featuring "cold gas cleanup," the syngas is cooled to about 100 °F before entering the acid gas removal plant section. Additionally, in many IGCC designs, moisture is added to the fuel gas in a fuel gas saturator to reduce NO_x formation during syngas combustion in the gas turbine. The low temperature gas cooling section consists primarily of a series of shell and tube heat exchangers. The fuel gas saturator is a vertical column with sieve trays in which fuel gas is contacted countercurrently with hot water flowing downward. Data for this particular plant section design was available from four studies for nine different sizes. In all four studies, the number of trains was two, with no spare. These studies were based on KRW, Texaco, and Dow IGCC systems (Fluor, 1985a; Fluor, 1986; Bechtel, 1988; Fluor Daniel, 1989). Although all "cold gas" IGCC systems have a fuel gas cooling process area, not all IGCC system designs are based on fuel gas moisturization. Alternatively, many are based on direct steam injection in the gas turbine.

The performance parameters that were common to all nine cases are the outlet syngas flow rate, the syngas moisture content, and the temperature and pressure of the outlet syngas. Of these, the syngas moisture content was not a statistical significant parameter in the regression models, nor is it expected to be the most important determinant of cost. The syngas flow rate generally affects the size of all the equipment in the plant section and for this reason is a major determinant of process area capital cost. The syngas outlet temperature affects the sizing of heat exchangers within the system. The syngas outlet pressure affects the selection of pressure vessel wall thickness. A regression model obtained from a two-step process (see Section 1.7.2) based on syngas outlet temperature, pressure, and flow rate, and the number of operating and total trains, is given as:

$$DC_{LT} = 8.78 \times 10^{-21} N_{T,LT} T_{syn,LT,o}^{2.38} P_{syn,LT,o}^{6.06} \left(\frac{M_{syn,LT,o}}{N_{O,LT}} \right)^{0.79} \quad (R^2 = 0.966; n = 9) \quad (26)$$

where,

$$85 \leq T_{LT,o} \leq 120 \text{ } ^\circ\text{F};$$

$$385 \leq P_{LT,o} \leq 435 \text{ psia; and}$$

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

The model should not be extrapolated for temperature or pressure. Extrapolation of the model for syngas flow rate per train is possible, because the scaling relationship between syngas flow rate and cost is quite reasonable. However, such extrapolation is not recommended, and should be confined to values not significantly different (e.g., within 50 percent) from the given range. The preferred alternative to extrapolation is to alter the number of trains so that the syngas mass flow rate per train is within the limits given above.

The standard error of the estimate is \$1.5 million. The error has a mean of zero and is approximately normally distributed, based on a Kolmogorov-Smirnov test. A comparison of the regression model and cost study direct cost estimates is shown in Figure 9 on page 62. The figure indicates that there is excellent agreement between the cost estimated by the regression model and reported in the design studies over a range of costs which vary by a factor of three.

2.5 Sulfur Removal

A number of different sulfur removal and recovery systems have been studied in IGCC and coal-to-SNG plant designs. The most common configuration is the Selexol process for sulfur removal from the raw syngas, a two-stage Claus plant for recovery of elemental sulfur, and the Shell Claus off-gas treating (SCOT) process for treatment of the tailgas from the Claus plant. However, a number of alternative designs have also been considered. These include integration of the Selexol and SCOT processes in the LONGSCOT design, as well as the use of alternative processes including the Dow GAS/SPEC MDEA and Selectox processes. The design basis assumed here is a Selexol unit for sulfur removal, a two-stage Claus plant for sulfur recovery, and either a SCOT or a Beavon-Stretford unit for Claus plant tail gas treatment. In this section, the development of a cost model for the Selexol process is discussed.

The proprietary Selexol process selectively removes hydrogen sulfide from the raw syngas. Typically, about 95 percent of the hydrogen sulfide is removed through counter-current contact of the syngas with Selexol solvent. The Selexol process also removes approximately 15 percent of the carbon dioxide in the flue gas. The composition of the acid gas stream which is sent from the Selexol unit to a sulfur recovery plant is typically over 50 percent carbon dioxide (Bechtel, 1983a; Bechtel, 1988; Cover et al, 1985a, 1985b; Fluor, 1983a, 1983b, 1984, 1985; Parsons, 1982). The studies cited here include both IGCC and coal-to-SNG systems based on a variety of gasifiers, including KRW, Texaco, and Shell designs. From these studies, 28 individual data points were developed. Thus, the database for the Selexol cost model represents a variety of coal gas compositions.

From the available performance and cost information for the Selexol process applied to gasification systems, a database containing total direct cost, syngas inlet flow rate, syngas composition (e.g., carbon dioxide, hydrogen sulfide, carbonyl sulfide, water vapor), removal efficiency of syngas components, acid gas flow rate and composition, and syngas temperature and pressure was developed. The inlet crude syngas temperatures for these data ranged from 95 to 120 °F and the inlet pressures ranged from 315 to 557 psia.

The inlet syngas is contacted countercurrently in a packed bed with Selexol solvent. For a more detailed discussion of this process area, the reader is referred to any of the design studies used as a basis for cost model development, and in particular Fluor (1985). The reactions occurring in the absorber reduce the temperature of the syngas. The treated syngas flows through a knock-out drum to remove solvent mist and is then heated in a heat exchanger by the incoming fuel gas. The cost of the Selexol section includes the acid gas absorber, syngas knock-out drum, syngas heat exchanger, flash drum, lean solvent cooler, mechanical refrigeration unit, lean/rich solvent heat exchanger, solvent regenerator, regenerator air-cooled overhead condenser, acid gas knock-out drum, regenerator reboiler, and pumps and expanders associated with the Selexol process.

The absorption of hydrogen sulfide by the solvent is influenced by the liquid to gas molar ratio in the absorption tower, the partial pressure of the hydrogen sulfide in the syngas, the contact temperature, the number of absorption stages or trays in the tower, and the amount of residual hydrogen sulfide left in the regenerated solvent (EPA, 1983). The absorption tower must be sized based on the syngas volume flow rate and the number of trays required for contacting solvent with the syngas. The solvent circulation rate depends on both the syngas molar flow rate and the desired removal efficiency for hydrogen sulfide. As the removal efficiency is increased, the solvent circulation rate must be increased (EPA, 1983). The solvent circulation rate affects the cost of most of the process equipment in the Selexol

process. However, data for the circulation rate are not reported in the design studies. Therefore, to a first order approximation, the cost of the Selexol process is assumed to depend on the syngas flow rate for the syngas temperature and pressure range of the database. The hydrogen sulfide removal efficiency is expected to have a secondary effect on cost, because it also influences the solvent circulation rate. Other parameters such as syngas temperature or the concentration of hydrogen sulfide in the syngas may also have secondary effects on the process area cost.

Several alternative regression model formulations were attempted based on syngas flow rate, temperature, pressure, hydrogen sulfide concentration, and the removal efficiency for hydrogen sulfide. The cost of the Selexol process was found to depend primarily on the syngas flow rate entering the acid gas absorber. The cost is also influenced to a much smaller degree by the hydrogen sulfide removal efficiency. Other parameters had less significant or statistically insignificant effects in explaining the cost of the system. Therefore, these additional parameters were excluded from the model. Thus, the cost correlation for the Selexol process is:

$$DC_S = \frac{0.420 N_{T,S}}{(1 - \eta_{HS})^{0.059}} \left(\frac{M_{syn,S,i}}{N_{O,S}} \right)^{0.980} \quad (R^2 = 0.909; n = 28) \quad (27)$$

where,

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

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

The range for the syngas molar flow rate per train indicates the size range for a single train. Because the scaling exponent for the syngas flow rate term is within the range typically expected for chemical process plants, extrapolation above this range may yield satisfactory results. However, the range for syngas molar flow per train is actually quite large, implying that extrapolation is unlikely in practice. Moreover, the preferred alternative to extrapolation is to adjust the number of trains so that the molar flow rate per train is within the given range. The range for the hydrogen sulfide removal efficiency should not be extrapolated. Based on a two-step regression as described in Section 1.7.2, the standard error of the estimate is \$5.1 million. A graph comparing the regression model estimates of direct cost with the costs reported in the literature is given in Figure 10.

Figure 10. Predicted vs. Actual Costs for Selexol Acid Gas Removal

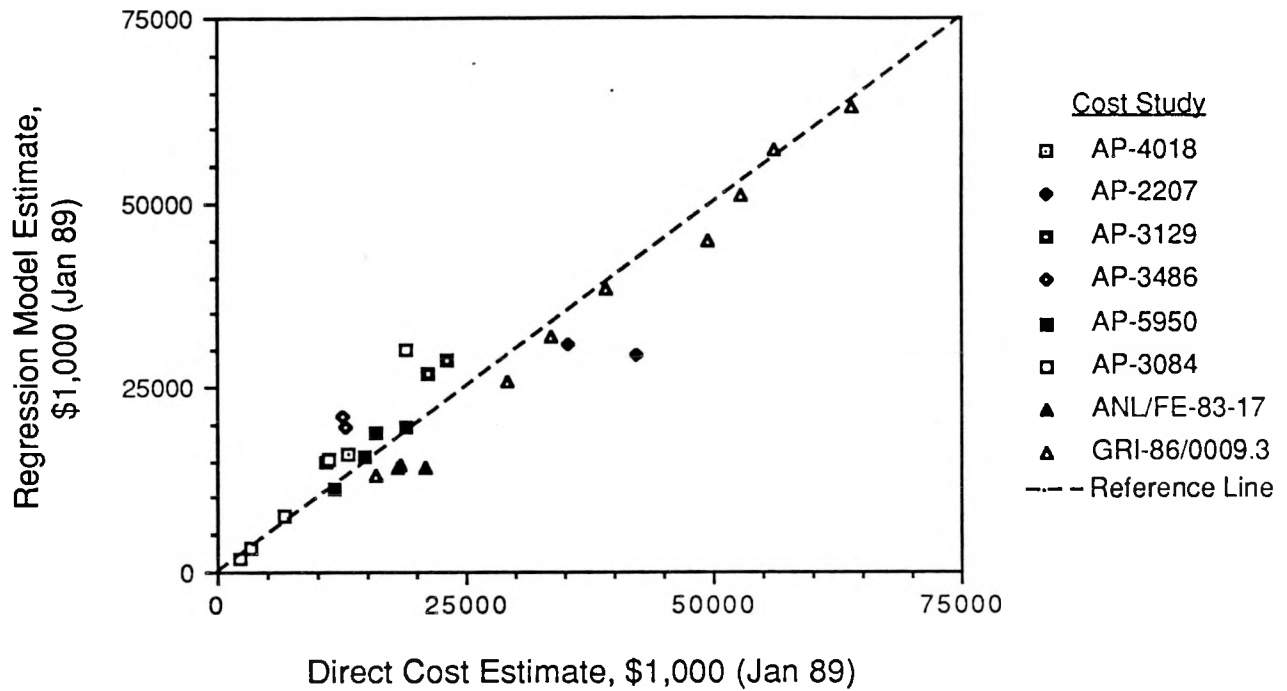
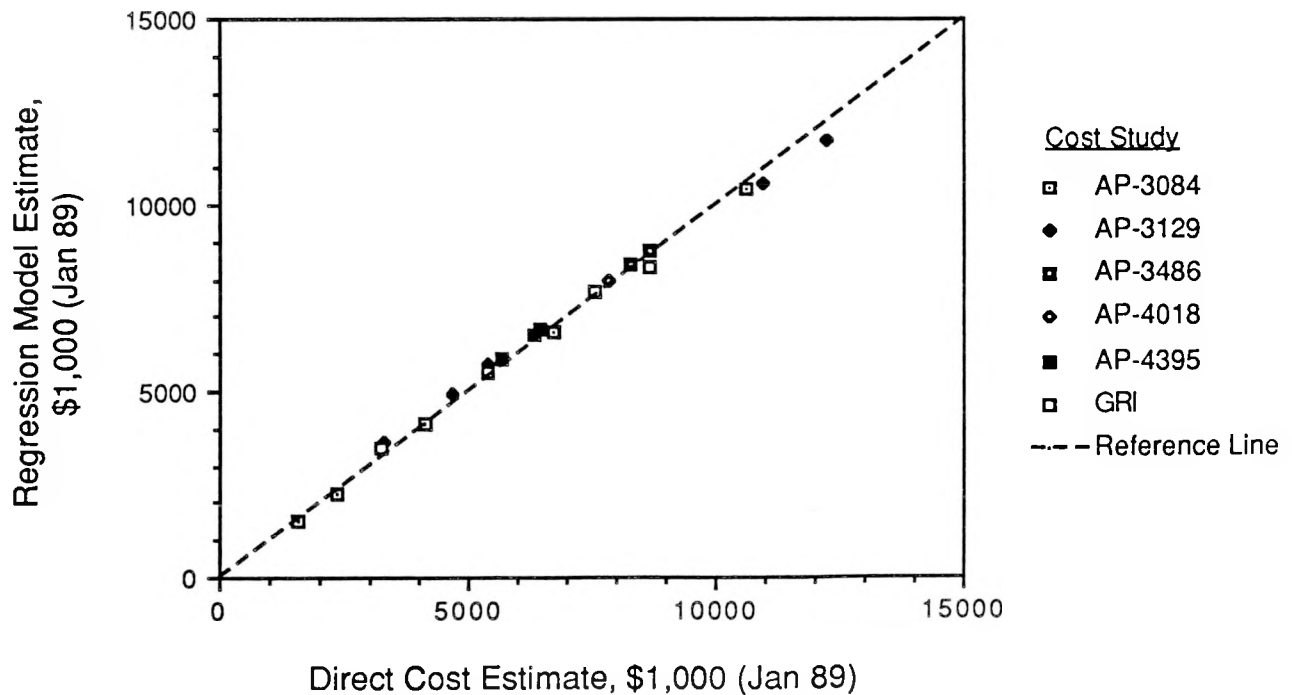


Figure 11. Predicted vs. Actual Costs for 2-Stage Claus Plants



2.6 Sulfur Recovery

In most IGCC cost studies, sulfur recovery is assumed to be achieved using a Claus plant to produce elemental sulfur. This section presents an overview of the design features of a Claus plant in the IGCC process environment. For additional detail see Fluor (1985) or any of the other detailed design studies of IGCC or coal-to-SNG systems used to develop this process area cost model.

The inlet stream to the Claus plant is the acid gas from the sulfur removal section. In this study, only data for Claus plants that process the acid gas from a Selexol unit are considered. The acid gas typically contains primarily carbon dioxide and hydrogen sulfide. In order to produce elemental sulfur, a 2:1 ratio of hydrogen sulfide and sulfur dioxide is required. Therefore, a portion of the incoming acid gas is combustion in a two-stage sulfur furnace. The furnace temperature is high enough in the first stage (typically 2,500 °F) to destroy any ammonia in the acid gas. Intermediate pressure steam (e.g., 350 psia) is generated from the waste heat produced in the sulfur furnace, cooling the feed gas to the Claus converters to about 600 °F. Further cooling to 350 °F occurs in a sulfur condenser, generating low pressure steam (e.g., 55 psia). Sulfur flows to a gravity sump, and is kept molten by condensing low pressure steam that flows through coils in the bottom of the sump.

Some of the furnace gas is used to heat the feed gas from the first condenser to approximately 450 °F prior to entering the sulfur converter, where hydrogen sulfide and sulfur dioxide react in the presence of a catalyst (e.g., Kaiser S-501) to produce elemental sulfur and water. This reaction is exothermic, and the outlet temperature of the gas is approximately 630 °F. The conversion rate is limited by thermal equilibrium. Gaseous sulfur is recovered in a second condenser. The cooling may be accomplished by heating water for fuel gas saturation. The feed gas then is mixed with remaining combustion gases and then enters the second converter. A third condenser, in which water for fuel gas saturation may be heated, is used for final sulfur recovery. The effluent gas from the Claus plant then passes through a coalescer and then on to tail gas treatment.

A direct cost correlation was developed for two-stage Claus plants based on data from a number of gasification plant studies. A number of data points are not included in this correlation because they represent either three-stage Claus plants or two-stage Claus plants with tail gas incineration and no tail gas treatment, with the incinerator costs included in the direct cost. The cost of a Claus plant is known to scale primarily with the recovered sulfur mass flow rate capacity using the standard exponential scaling model of Equation (1) with an

exponent of approximately 0.6 (e.g., EPA, 1983b). It appears that this scaling rule may have been the basis for developing the cost estimates of Claus plants used in the design studies, because an excellent goodness-of-fit was found for a single variable regression based on sulfur recovered. The scaling exponent that was obtained in the single variate analysis was 0.668. The regression model was further developed to represent the number of operating and spare trains for each data point in the database. The resulting cost equation is:

$$DC_c = 6.28 N_{T,C} \left(\frac{m_{S,C,O}}{N_{O,C}} \right)^{0.668} \quad (R^2 = 0.994; n = 21) \quad (28)$$

where,

$$695 \leq \left(\frac{m_{S,C,O}}{N_{O,C}} \right) \leq 18,100 \text{ lbmole/hr}$$

The standard error of the estimate is \$235,000, and this is probably primarily attributable to differences in site location. The regression model is shown graphically in Figure 11 on page 67.

As indicated above, the capacity of a single train varies by a factor of over 20. Typically, one or two operating trains and one spare train are used, each with equal capacity. Because there was a prior expectation that the cost of the Claus plant should be modeled using an exponential scaling relationship based on recovered sulfur capacity, with a coefficient near 0.6, this model can be extrapolated at the high end of the range. However, as with all other models, it is recommended that the number of trains be selected so that extrapolation is not required.

2.7 Tail Gas Treating

Cost models for two tail gas treatment technologies are developed in this section. The Shell Claus Off-gas Treating (SCOT) process is commonly assumed in the studies performed for EPRI. The Beavon-Stretford process is assumed in the ASPEN flowsheet for the oxygen-blown KRW-based IGCC system with cold gas cleanup.

2.7.1 SCOT Process

The Shell Claus Off-gas Treating (SCOT) process removes hydrogen sulfide from atmospheric pressure gases, such as the tail gas from a Claus plant, in a manner similar to the Selexol process, and the concentrated acid gas stream from the SCOT process is then sent to the Claus plant for sulfur recovery. The process is proprietary, and limited performance data

regarding this process are available in the detailed cost studies. The process is available commercially.

The tail gas from the Claus plant typically contains a number of sulfur species, including hydrogen sulfide, sulfur dioxide, carbonyl sulfide, and elemental sulfur. However, the SCOT solvent is not suitable for removing sulfur species other than hydrogen sulfide. Therefore, the tail gas is reacted in a hydrogenation reactor in the presence of a cobalt-molybdenum catalyst to convert all sulfur species to hydrogen sulfide. Hydrogen and carbon monoxide are required for these reactions, and are supplied by mixing flash gas from the acid removal unit with the Claus plant tail gas. The flash gas is partially combusted in a reducing gas generator prior to entering the hydrogenation reactor, which raises the temperature of the inlet gases to the appropriate reaction temperature. The outlet gas from the hydrogenation reactor is cooled in a waste heat boiler, where intermediate pressure steam (e.g., 100 psia) is generated. The gas is further cooled by direct contact with water in a quench water cooling tower.

The cooled gas containing hydrogen sulfide then enters an absorber vessel where almost all of the hydrogen sulfide and a portion of the carbon dioxide is absorbed by countercurrent contact with an amine solution. The treated tail gas goes to the gas turbine air intake, and contains approximately 300 ppm of total sulfur.

Like the Selexol process, the hydrogen sulfide-rich amine solution is pumped through a lean/rich solvent heat exchanger, and then enters the regenerator vessel, where acid gases are stripped from the solution by indirect application of heat. This heat is supplied by condensing steam in the regenerator reboiler. The lean solvent leaves the regenerator and is pumped through the lean/rich solvent heat exchanger to the absorber vessel. The acid gas is cooled by cooling water in an overhead condenser, and flows through a knock-out drum for removal of any amine solution prior to recycle to the Claus plant sulfur furnace.

The cost of the SCOT process is expected to vary directly with performance parameters such as the tail gas flow rate, flash gas flow rates, and amine solution flow rates. However, these data are not available from the cost studies because of the proprietary nature of the process. The SCOT process is used in conjunction with the Claus process to achieve approximately 99.7 to 99.9 percent recovery of the sulfur removed from the syngas. The Claus plant removal efficiencies vary within a relatively narrow range of about 91 to 95 percent. This implies that the total sulfur recovered to the sulfur loading facilities could serve as a proxy for the unavailable detailed performance information about the SCOT process.

In analyzing the relationship between the cost of the SCOT process and the total amount of sulfur sent to the loading facilities, two groupings of data were identified. These include a set of ten cost estimates based on four studies performed for EPRI in 1983 to 1986, referred to here as "Case 1", and a set of ten cost estimates based on five studies performed for EPRI or GRI in 1987 and 1988, referred to as "Case 2". The Case 2 cost estimates are significantly higher than the Case 1. Furthermore, the Case 1 cost estimates correlate almost exactly with the sulfur flow rate to the loading facilities, indicating that perhaps a simple scaling equation was used in developing these costs. The Case 2 estimates are more scattered, indicating perhaps that more detailed cost estimates, based on several performance parameters, were used or that the studies were performed with different assumptions or data. The latter may be a likely cause of the variation in costs, because three separate contractors prepared the cost estimates.

The cost correlations for Case 1 and Case 2 are summarized below, and shown graphically in Figure 12:

$$\text{Case 1: } DC_{SC} = 2.64 m_{s,S,o}^{0.75} \quad (R^2 = 0.998; n = 10) \quad (29)$$

where,

$$4,900 \leq m_{s,S,o} \leq 30,800 \text{ lb/hr}$$

and,

$$\text{Case 2: } DC_{SC} = 1,700 + 0.29 m_{s,S,o} \quad (R^2 = 0.865; n = 10) \quad (30)$$

where,

$$3,700 \leq m_{s,S,o} \leq 16,600 \text{ lb/hr.}$$

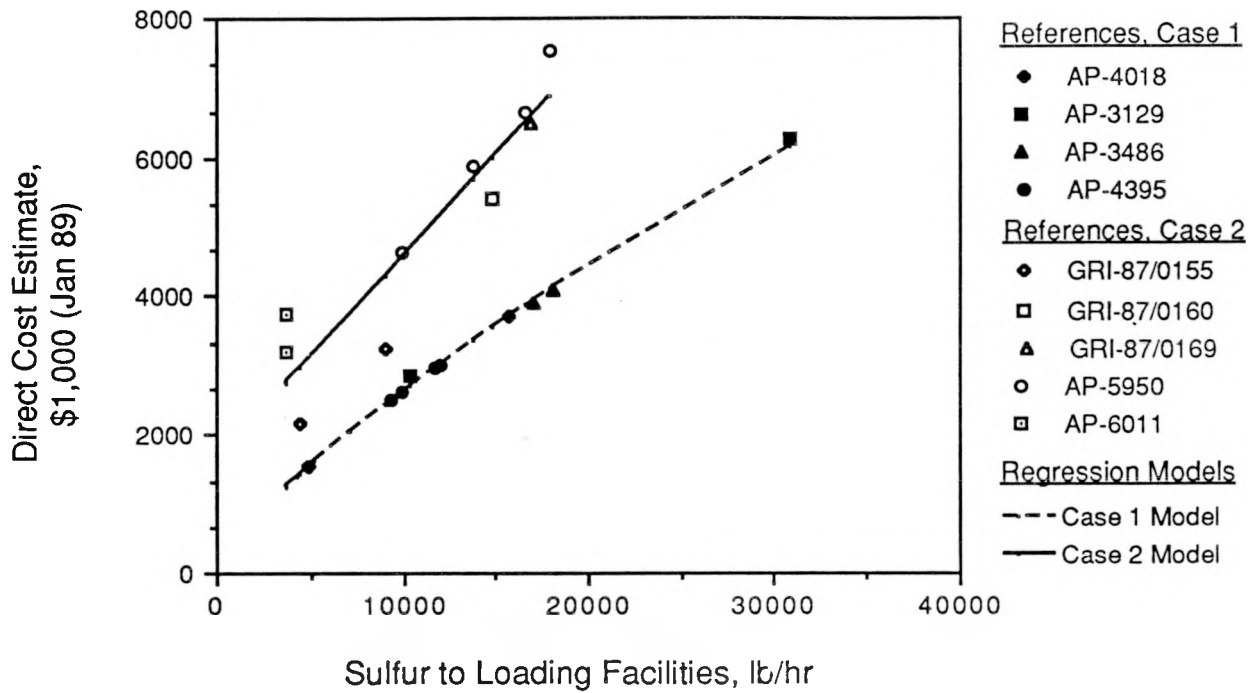
In Case 1, the standard error is a low \$60,000, while in Case 2 the standard error is \$700,000. The number of trains assumed in the cost studies is one for all cases except for one study (AP-3129).

The Case 2 correlation is recommended for use in the cost models because it represents the most recent cost studies. Because this model is based on a proxy variable, rather than on performance variables that directly affect the sizing of SCOT process equipment, this model should not be extrapolated.

2.7.2 Beavon-Stretford Process

In this section, an overview of the performance and design of the Beavon-Stretford process is presented as background information for the development of a regression cost model. See Fluor (1983a) or Fluor (1983b) for a more detailed discussion of this process.

Figure 12. Direct Cost of the SCOT Process



The Beavon-Stretford process is a modification of the Stretford process, which is designed to remove hydrogen sulfide from atmospheric pressure gas streams and convert it to elemental sulfur. However, the Stretford process is not appropriate for handling effluent gases containing sulfur dioxide, carbonyl sulfide, or elemental sulfur. Therefore, a Beavon unit is used to catalytically reduce or hydrolize these species to hydrogen sulfide in the presence of a cobalt molybdate catalyst.

Because hydrogen is required for the reactions occurring in the Beavon unit, flash gas from the acid gas removal section is used as a feed stream. The flash gas is partially combusted in a reducing gas generator, mixed with the Claus plant tail gas, and the total gas stream then enters the Beavon hydrogenation reactor. The hot gas from the reactor is cooled in a waste heat boiler where intermediate pressure (e.g., 100 psia) steam is generated. The gas stream is further cooled in the desuperheater section of a thermally integrated desuperheater/absorber vessel. The cooling of the gas stream is accomplished by heat transfer with cooling water, which is recirculated through an air-cooled heat exchanger. The gas stream then enters the absorber portion of the vessel, where over 99 percent of the hydrogen sulfide is removed by contact with a Stretford solution containing sodium carbonate. The treated gas is vented to the atmosphere.

The Stretford solution flows to a soaker/oxidizer, where anthraquinone disulfonic acid (ADA) is used to oxidize the reduced vanadate in the Stretford solution. The ADA is regenerated by air sparging, which also provides a medium for sulfur flotation. The sulfur overflows into a froth tank, and the underflow from the oxidizer/soaker is pumped to a Stretford solution cooling tower and then to a filtrate tank.

The sulfur from the froth tank is pumped to a primary centrifuge, where the wet sulfur cake product is reslurried and sent to a second centrifuge, after which the sulfur is again reslurried. The slurry is then pumped through an ejector mixer, where the sulfur is melted and separated in a separator vessel. The sulfur goes to a sump.

The process is considered commercially available. The capital cost of a Beavon-Stretford unit is expected to vary with the volume flow rate of the input gas streams and with the mass flow rate of the sulfur produced. Data from two EPRI-sponsored studies were used to develop a regression cost model (Fluor, 1983a; 1983b). An additional two studies were reviewed for inclusion in the database, but information regarding key process parameters (e.g., recovered sulfur flow rate) was not reported. The two EPRI studies report limited performance and cost data for nine different Beavon-Stretford unit sizes. For example, there is incomplete information about inlet gas streams flow rates. Because of the limited availability

of performance data, a regression analysis based only on the sulfur produced by the Beavon Stretford process was developed. However, this regression yielded an excellent fit to the data:

$$DC_{BS} = 57.5 + 66.2 N_{T,BS} \left(\frac{m_{s,BS,o}}{N_{O,BS}} \right)^{0.645} \quad (R^2 = 0.998; n = 7) \quad (31)$$

where,

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

The high coefficient of determination indicated for this model implies either that an exponential cost model is an excellent predictor of the costs of Beavon-Stretford units, or that the costs developed in the EPRI studies were based on a simple scaling model as an approximation. Therefore, it is not immediately clear if this model merely represents an accepted industry practice for developing preliminary cost estimates, or if it accurately reflects the cost of Beavon-Stretford units. The standard error of the estimate is \$260,000, and the errors have an approximately normal distribution with a mean of zero. Typically, two operating and one spare train are assumed. Figure 13 on page 72 shows a comparison of the regression model estimate and the cost study estimates. Although the regression model is an excellent fit to the data, it is recommended that the number of trains be adjusted so that the recovered sulfur flow rate per train does not exceed the limits given above. As a default, the number of operating and total trains for this process area is assumed to be the same as for the Claus plant process area.

2.8 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. Data on the cost of the boiler feedwater section and the flow rates of the raw water and polished water streams is available from five studies for 14 plant sizes. These studies include Texaco-based, Shell-based, and KRW-based IGCC systems (Fluor, 1983a; 1983b; 1984; 1985; 1986). Because all of these studies were developed by the same contractor using a consistent approach, they provide an excellent basis for developing a cost model. The boiler feedwater section is generic to the steam cycle.

The cost of the boiler feedwater section is expected to depend 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 was found to yield good results:

$$DC_{BF} = 0.145 m_{rw}^{0.307} m_{pw}^{0.435} \quad (R^2 = 0.991; n = 14) \quad (32)$$

where,

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

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

For this model, a nonlinear variable transformation was used. The error of the linearized model is approximated by a normal distribution. Therefore, the error of the nonlinear model shown above is represented by a lognormal distribution. The median of the errors is 1.0, with a mean of 1.002 and a standard deviation of 0.063. The 90 percent probability range for the error is approximately 0.9 to 1.1, implying a 90 percent confidence band of 90 to 110 percent of the nominal cost estimate.

Typically only one train of equipment is used in this section, and all the equipment is commercially available. A comparison of the regression model cost estimates and the direct cost estimates from the detailed cost studies is shown in Figure 14. This model should not be extrapolated beyond the range of the predictive variables as indicated above. However, because the cost of the boiler feed water section is a very small portion of the total direct cost for a typical IGCC plant, the effect of any errors introduced by modest extrapolations may be acceptable for some purposes.

2.9 Process Condensate Treatment

The process condensate treatment section is used to treat blowdown from the particulate scrubber and process condensate from gas cooling (Fluor, 1983b; 1985). These streams contain ammonia, carbon dioxide, and hydrogen sulfide, and the scrubber blowdown also has a high chlorides content. The blowdown and condensate stream are treated in separate strippers. The overhead vapor streams from both strippers are cooled in air-cooled heat exchangers and then they flow through knock-out drums prior to feed to the Claus plant sulfur furnace. The stripped bottoms product from the blowdown water stripper is cooled by

Figure 14. Predicted vs. Actual Direct Costs for the Boiler Feedwater Section

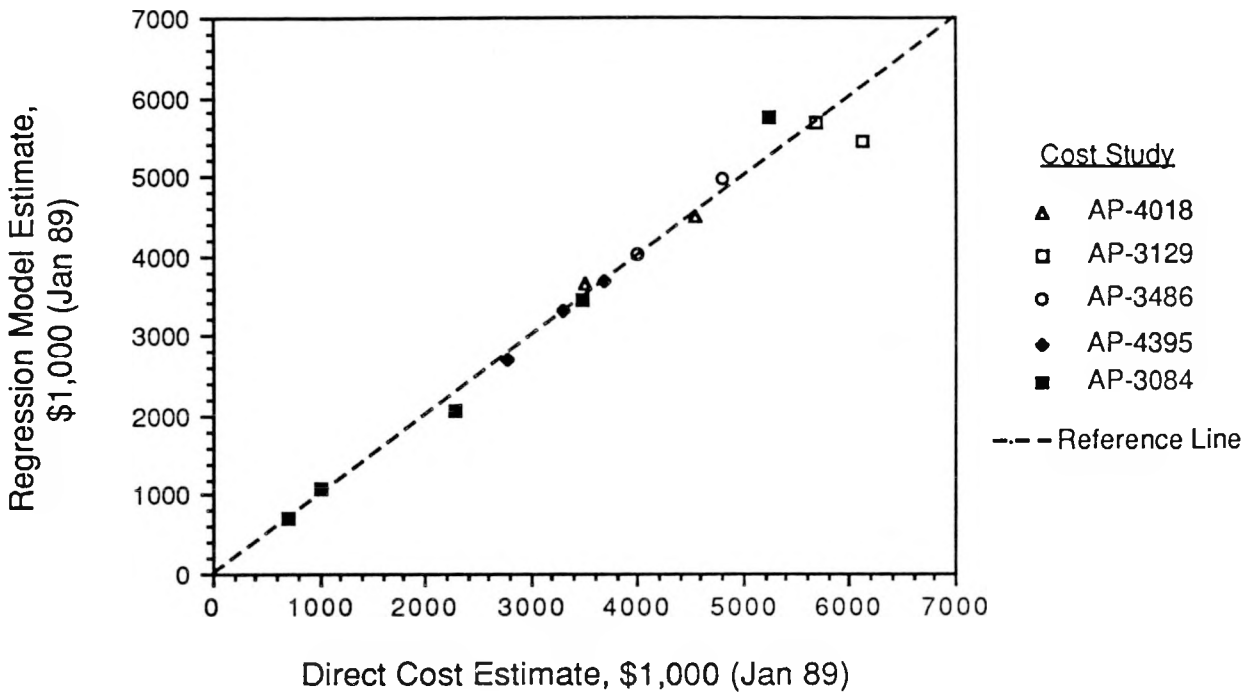
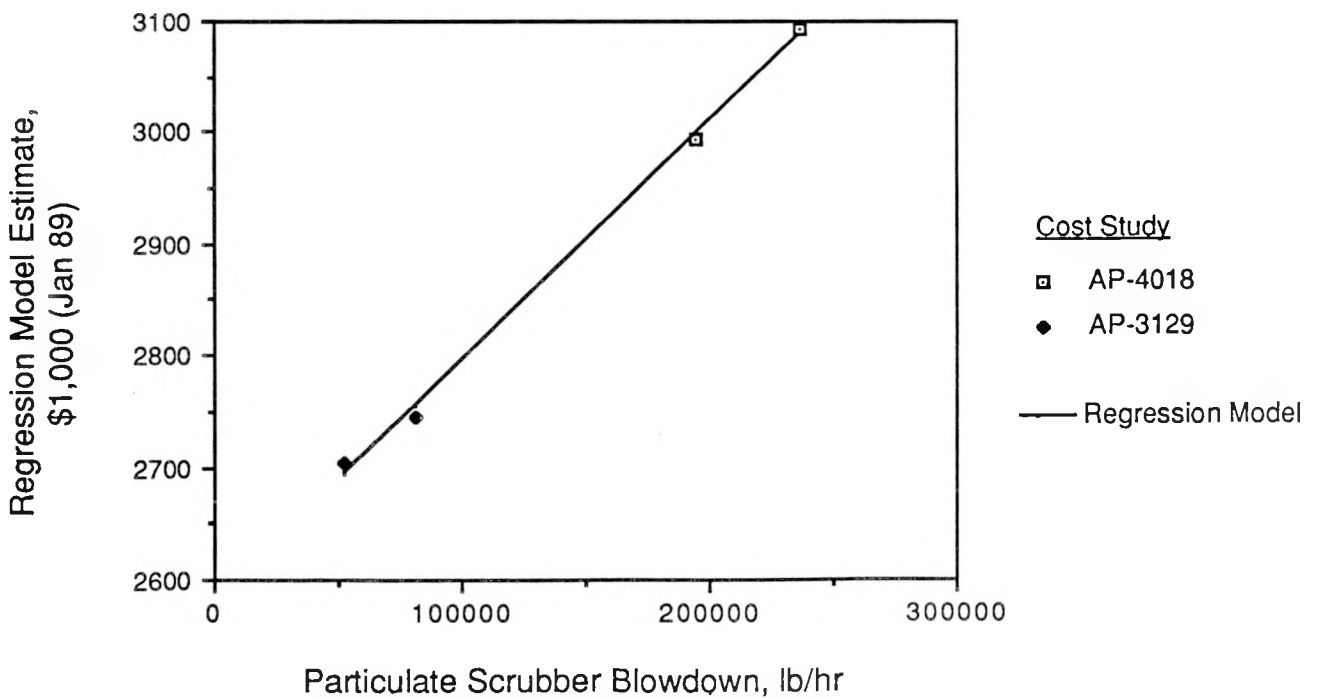


Figure 15. Direct Cost for the Process Condensate Treatment Section



the incoming process condensate water and then sent to a water treatment plant for biological treatment prior to flow to the cooling tower. The bottoms from the process condensate water stripper is sent as make up to the gas scrubbing unit.

Because the treated process condensate is used as make-up to the gas scrubbing unit, and because blowdown from the gas scrubbing unit is the larger of the flow streams entering the process condensate treatment section, it is expected that process condensate treatment direct cost will depend primarily on the scrubber blowdown flow rate. Because only two cost studies were identified with similar designs and sufficient detail for regression analysis, a single variate regression analysis was used:

$$DC_{PC} = 2583 + 0.00213 m_{sbd} \quad (R^2 = 0.998; n = 4) \quad (33)$$

where,

$$52,000 \leq m_{sbd} \leq 237,000 \text{ lb/hr.}$$

The standard error of the estimate is \$11,000. Typically only one operating train is used, with no spare. The equipment is commercially available. However, the designs depend on the chemical composition of the coal feedstock. Detail regarding the design of these systems is not available from the cost studies. The regression model is shown graphically in Figure 15 on page 76.

2.10 Gas Turbine

The most commonly assumed gas turbine in IGCC performance and cost studies is the General Electric (GE) model MS7001F, also referred to as the "Frame 7F". This gas turbine is designed for a turbine inlet temperature of 2,300 °F and has a power output of about 125 to 150 MW. By contrast, typical gas turbines have firing temperatures in the range of 2,020 to 2,150 °F. The thermal efficiency of gas turbines increases as the firing temperature increases. The higher firing temperature is the result of advances in turbine blade manufacturing. The Frame 7F turbine blades are manufactured using a process known as "directional solidification," which has been used on smaller aircraft engine turbine blades. Because of improvements in molding technology, the process can now be applied to the larger turbine blades of the Frame 7F. Further advances in manufacturing techniques may lead to the capability to cast turbine blades as single crystals with no grain boundaries, permitting an additional 50 to 150 °F increase in firing temperature (Smock, 1989).

The first Frame 7F has completed factory tests at General Electric and has been delivered to a Virginia Power site in Chesterfield, VA as part of a combined-cycle power plant. General Electric has rated this machine at 150 MW with a heat rate of 9,880 Btu/kWh.

There are a number of design factors that affect the cost of a gas turbine in an IGCC process environment. For example, the firing of medium-BTU coal gas, as opposed to high-BTU natural gas, requires modification of the fuel nozzles and gas manifold in the gas turbine (BGE, 1989). Some additional concerns associated with firing coal gas are discussed by Cincotta (1984). The presence of contaminants in the syngas may affect gas turbine maintenance and long term performance. Liquid droplets may cause uneven combustion or may burn in the turbine first-stage nozzles, causing damage. Solids can deposit on fuel nozzles or cause erosion in the hot gas path of the gas turbine (e.g., combustor, turbine). Alkali materials that deposit on hot gas path parts cause corrosion. It is expected that, at fuel gas temperatures less than 1,000 °F, that alkali material is essentially condensed on any particulate matter in the raw syngas, and that the alkali removal efficiency is approximately the same as the particle removal efficiency. For sufficiently high particle removal efficiencies, erosion is not expected to be a problem. Corrosion is not expected to be any worse than for distillate oil firing. Deposition of particles is expected to be within the allowance of reasonable maintenance schedules. The design for an advanced high firing temperature gas turbine employs advanced air film cooling which could be affected by the ash content of combustion products.

Another design issue is the gas turbine fuel inlet temperature. A study by Fluor (Earley and Smelser, 1988) assumes that hot desulfurized syngas from an advanced hot gas cleanup process is fed directly to the gas turbine at 1,200 °F. The Fluor study indicates that General Electric expects that a fuel system capable of a 1,200 °F fuel inlet temperature could be developed by 1994. The maximum fuel temperature test to date has been at 1,000 °F. An earlier study with hot gas cleanup included a hot gas cooler to reduce the gas temperature to 1,000 °F (Corman, 1986). For the KRW system with cold gas cleanup, the coal gas temperature is within the limits of current technology. However, the gas turbine costs developed here should not be used in conjunction with IGCC systems featuring hot gas cleanup without some adjustments to account for the uncertainty in using a higher fuel inlet temperature.

Unfortunately, there is currently a lack of reported data from which to develop a detailed gas turbine cost model that is explicitly sensitive to the type of factors discussed above. In preliminary cost estimates, the typical approach to accounting for these uncertainties in performance, or for the possibility of increased capital cost due to design modifications, is through process contingency factors. The approach taken here is to use the

available cost data for the GE Frame 7F to develop a cost estimate for a single gas turbine. In the use of this cost estimate for actual case studies in a later task, judgments about the uncertainty in cost, and about the likelihood of cost increases for applications with coal gases, will be encoded using either process contingency factors or the new probabilistic modeling capability developed for ASPEN, as discussed in Section 1.7.

Although cost estimates of the GE Frame 7F are available in a number of IGCC cost studies, recent cost estimates are significantly higher than older estimates. However, the more recent estimates are expected to be more reliable, because the Frame 7F was at or near commercialization at the time of the recent studies. In four recent site-specific IGCC studies performed for EPRI (BGE, 1989; Fluor Daniel, 1988, 1989; FPL, 1989), 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 (Jan 89). This cost excludes equipment associated with combined cycle systems, which are discussed in the following two sections. In two other studies (JCP&L, 1989; NUSCo, 1988), 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. The higher estimate of \$32.0 million per unit is consistent with the expectation that the cost of the gas turbine modified to fire medium-BTU coal gas will be higher than for the standard natural gas-fired unit. This high estimate will be used in the cost model:

$$DC_{GT} = 32,000 N_{T,GT} \quad (34)$$

A competitor to the GE Frame 7F is under development by Mitsubishi Heavy Industries and Westinghouse Electric. The prototype model 501F is expected to achieve a rating of 148.8 MW and a turbine inlet temperature of 2,300 °F. This model is expected to be commercially available in 1992 (GTW, 1989). No cost data are currently available for this model; however, competition between the Frame 7F and the 501F could result in similar prices for both machines.

A Kraftwerk Union (KWU) gas turbine, model 84.2, was analyzed in an EPRI study (Fluor Daniel, 1988). This is a commercially available, moderate firing temperature machine that is rated at approximately 100 MW with a cost of about \$24.2 million per unit. The combustor features a low-NO_x design, and does not require water injection when operated on natural gas.

2.11 Heat Recovery Steam Generator

The heat recovery steam generator (HRSG) is a set of heat exchangers in which heat is removed from the gas turbine exhaust gas to generate steam. Typically, steam is generated at two or three different pressures, and associated with the HRSG is one steam drum for each steam pressure level. High pressure superheated steam is generated for use in the steam turbine, and typically the exhaust from the steam turbine first stage is reheated. The input streams to the HRSG section include the gas turbine exhaust 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, including the superheater, reheater, high pressure steam drum, high pressure evaporator, and the economizers.

Most studies of IGCC systems aggregate the cost of the HRSG units with the cost of the gas turbine and the steam turbine. Only four studies were identified in which the cost of the HRSG units were reported as a separate line item. A study of Texaco and British Gas/Lurgi IGCC systems includes performance and cost estimates for several sizes of HRSGs used in combination with reheat steam turbines (Parsons, 1982). These HRSG units include two steam pressure levels, and are used in conjunction with a conventional gas turbine. The high pressure steam varies from 650 psia to 1520 psia for these HRSGs. The exhaust gas flow rate and temperature indicate that the gas turbine is a GE Frame 7E or equivalent. A study by Bechtel and WE (1983c) for a KRW-based system included an HRSG design with three pressure levels using a large 130 MW gas turbine with a high exhaust gas temperature. A study of Texaco-based IGCC systems included performance and cost estimates for reheat steam turbines and HRSGs with two pressure levels (Fluor Technology, 1986). A recent study of Dow-based IGCC systems includes performance and cost estimates for two-pressure level reheat HRSGs applied in conjunction with large advanced gas turbines (Fluor Daniel, 1989).

A detailed approach to estimating the cost of HRSGs is reported by Foster-Pegg (1986). This approach requires detailed performance and design information for each heat exchanger in the HRSG. The necessary design values were not reported in the performance and cost studies, nor was sufficient detail about performance available to develop such a model. Furthermore, the level of detail in the Foster-Pegg model is not justifiable for the applications envisioned for this model for several reasons. The technical and cost growth risks of IGCC systems reside primarily in process areas such as gasification, gas cleanup, and advanced gas turbine designs. The HRSG is a conventional, commercially available component. Therefore, the priorities for cost model development should be with the more

innovative systems. Secondly, in comparative studies of IGCC systems, the cost of HRSGs will be similar, and will not be a factor in distinguishing one system from another. Instead, differences in the gasification process area, gas cleanup, and byproduct recovery, as examples, are expected to be important in distinguishing alternative systems. Third, the purpose of this model is not to develop detailed, final estimates of site-specific costs for a particular project, but to develop preliminary cost estimates for the purpose of research planning. Therefore, there is not a need for a highly detailed cost model for this particular process area.

The cost of the HRSG is expected to depend on factors such as the high pressure steam flow rate to the steam turbine, the pressure of the steam, the gas turbine exhaust gas volume flow rate, the number of steam drums, and, to a lesser extent, the boiler feed water or saturated steam flowrates in each of the heat exchangers in the HRSG. A variety of regression models were investigated to represent these potential predictive parameters. However, because only 10 data points are included in the database, only a limited number of predictive parameters can be reasonably included in the model, based on statistical considerations. Furthermore, some parameters that are expected to be important in determining HRSG cost, such as the gas turbine exhaust flow rate, are not statistically important for this data set. When the gas turbine exhaust flow rate, high pressure inlet steam flow rate to the steam turbine, and the steam pressure are included in a regression model, the exponent for exhaust flow rate is small and is not statistically significant. The exhaust gas flow rate is not an influential predictive parameter because the cost studies are based primarily on either GE Frame 7E or 7F gas turbines; therefore, there was not a large range of variation for the exhaust gas flow rate. A simple regression model based only on the high pressure steam flow rate to the steam turbine yielded a high coefficient of determination. A multivariate regression based on the high pressure steam flow to the steam turbine and the pressure of the steam yielded satisfactory results:

$$DC_{HR} = -5364 + 7.24 \times 10^{-3} N_{T,HR} P_{hps,HR,o}^{1.526} \left(\frac{m_{hps,HR,o}}{N_{O,HR}} \right)^{0.242} \quad (R^2 = 0.966; n = 10) \quad (35)$$

where,

$$650 \leq P_{hps,HR,o} \leq 1,545 \text{ psia; and}$$

$$66,000 \leq \frac{m_{hps,HR,o}}{N_{O,HR}} \leq 640,000 \text{ lb/hr.}$$

The standard error of the estimate is \$6.0 million. In all the cost studies one HRSG is assumed for each gas turbine, with no spares. A comparison of the regression model and the cost estimates is shown graphically in Figure 16. The regression model should not be extrapolated beyond the ranges for the predictive parameters indicated above.

2.12 Steam Turbine

A typical steam turbine for an IGCC plant consists of high-pressure, intermediate-pressure, and low-pressure turbine stages, a generator, 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 from the intermediate pressure stage goes to the low pressure stage.

The cost of a steam turbine is expected to depend on the mass flow rate of steam through the system, the pressures in each stage, and the generator output, among other factors. Nine cost estimates for the steam turbine were available from four studies. A single-variate regression based on the generator output was found to yield reasonable results:

$$DC_{ST} = 158.7 MW_{ST} \quad (R^2 = 0.958; n = 9) \quad (36)$$

where,

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

The standard error of the estimate is \$5.5 million. Only one steam turbine is used in most IGCC designs. A graphical representation of the regression model is shown in Figure 17.

2.13 General Facilities

General facilities include:

Cooling water systems

- Process heat rejection
- Condensation of exhaust steam
- Cooling of mechanical equipment

Plant and instrument air

- Motor driven compressors, dryer packages

Potable and utility water

- Motor driven pumps, in-line chlorination
- Storage tanks, supply drum

Figure 16. Predicted vs. Actual Direct Cost for Heat Recovery Steam Generators.

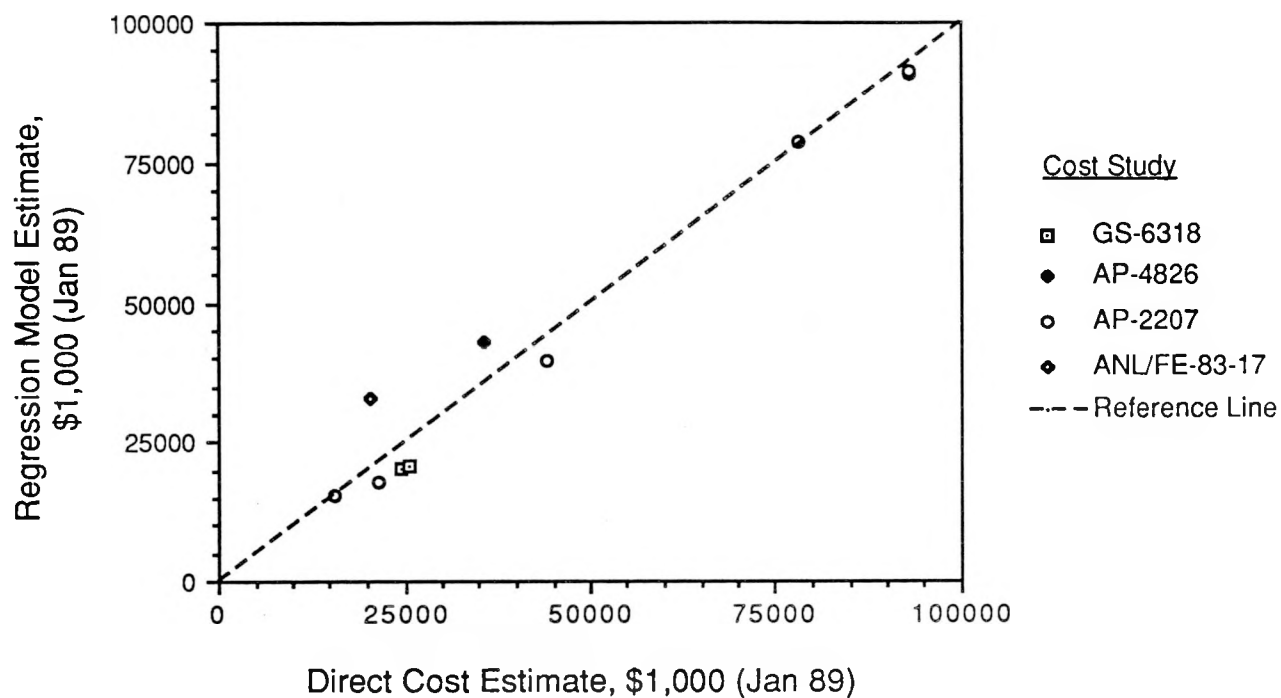
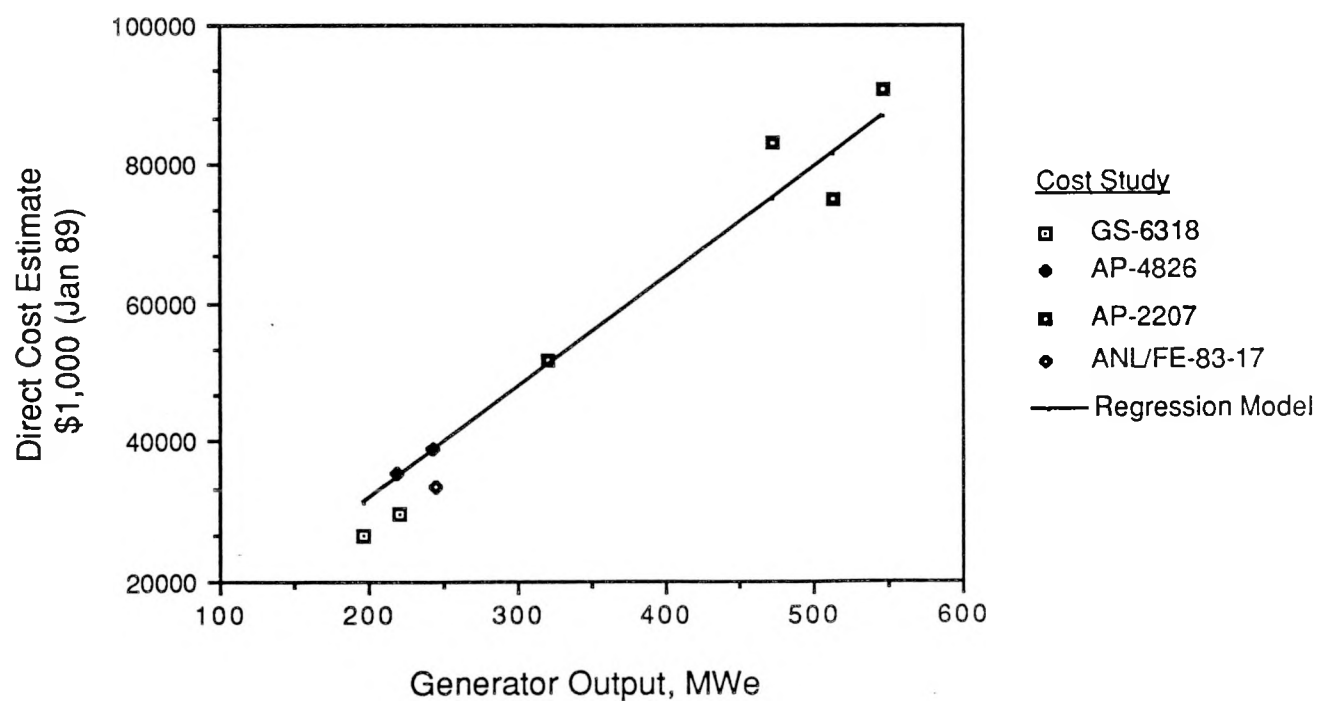


Figure 17. Direct Cost for the Steam Turbine-Generator Section



Fuel system	Fuel oil for startup, 3 30,000 bbl tanks, two motor driven pumps and suction heaters
Nitrogen system	Liquid N ₂ storage sphere, air fin heaters
Effluent water treating	Storm and oily water Utility wastewater Cooling tower blowdown
Flare system	Elevated flare stacks, pilot flame LPG tank, pumps, and vaporizer, condensate separator,
Fire water system	Fire water loop, motor driven jockey pump, hydrants, monitors, fire water pumps, fire water storage tank
Interconnecting piping	Major process and utility piping between battery limits of plant units.
Buildings	Substations Control houses Operators shelters Administration Laboratory Cafeteria Change house and guard house Fire house First aid Maintenance Warehouse
Railroad facilities, roads, and lighting	Unloading coal, loading liquid sulfur 15,000 lineal ft of track, 3 switches, 1 bumper 6 RR crossings. Heated liquid sulfur storage tanks. Security and plant service roads, fencing, parking, street lighting.
Computer control system	Control centers, consoles, panels, control and measurement devices, computer, interfaces, graphic displays, logging and reporting, and data highway system

Electrical system

captive transformers
high voltage electric power distribution from
switchyard to high voltage substations
high voltage substations, step-down
transformers

These equipment items are typically represented in preliminary, study-grade cost estimates as a percentage of the direct costs. Most studies assume that general facilities are approximately 15 percent of direct costs. The cost of general facilities can be estimated as:

$$GF = f_{GF} \sum_{i=1}^{12} DC_i \quad (37)$$

where,

$$f_{GF} = 0.15.$$

3.0 CAPITAL COST OF A KRW-BASED SYSTEM WITH HOT GAS CLEANUP

A schematic of an air-blown KRW-based IGCC system with hot gas cleanup is shown in Figure 18. The schematic represents process elements based on design and cost studies prepared for GRI (Smelser, 1986a; Earley and Smelser, 1988b) and the configuration assumed in the ASPEN simulation model (Craig, 1988). The primary features of a hot gas cleanup design, compared to the IGCC system with cold gas cleanup presented in Figure 5, are: (1) elimination of an oxygen plant; (2) in-situ desulfurization with limestone or dolomite; (3) external (e.g., not in the gasifier) desulfurization using a high temperature removal process; (4) reduced requirement for syngas cooling prior to desulfurization; (5) elimination of sulfur recovery and tail gas treating; and (6) addition of a circulating fluidized bed boiler for sulfation of spent limestone (to produce an environmentally acceptable waste) and conversion of carbon remaining in the ash. In the GRI studies, recycle syngas is cooled for the purposes of pneumatic transport of coal and ash, and for cooling of the raw syngas through mixing. The sour condensate from the raw gas cooler knock-out drum must be treated prior to discharge. However, the ASPEN flowsheet contains no provision for estimating process condensate streams which require treatment. Instead, the design basis for the ASPEN flowsheet includes the use of water quench, rather than heat exchange, for high temperature syngas cooling. Therefore, there is no knock-out drum for process condensate removal.

Many of the direct cost models developed in Section 2.0 are applicable to an air-blown KRW-based system with hot gas cleanup. These cost models include:

1. Coal Handling (Area 20);
2. Boiler Feedwater System (Area 80);
3. Gas Turbine (Area 91);
4. Heat Recovery Steam Generators (Area 92);
5. Steam Turbine (Area 93);
6. General Facilities (Area 100).

The reader is referred to the discussions of these systems in Section 2.0. In this section, the direct capital cost models required in addition to those above are presented. These cost models include:

1. Limestone Handling (Area 25);
2. Oxidant Feed (Area 10);
3. Gasification (Area 30);
4. Zinc Ferrite Desulfurization (Area 50);
5. Sulfation (Area 35).

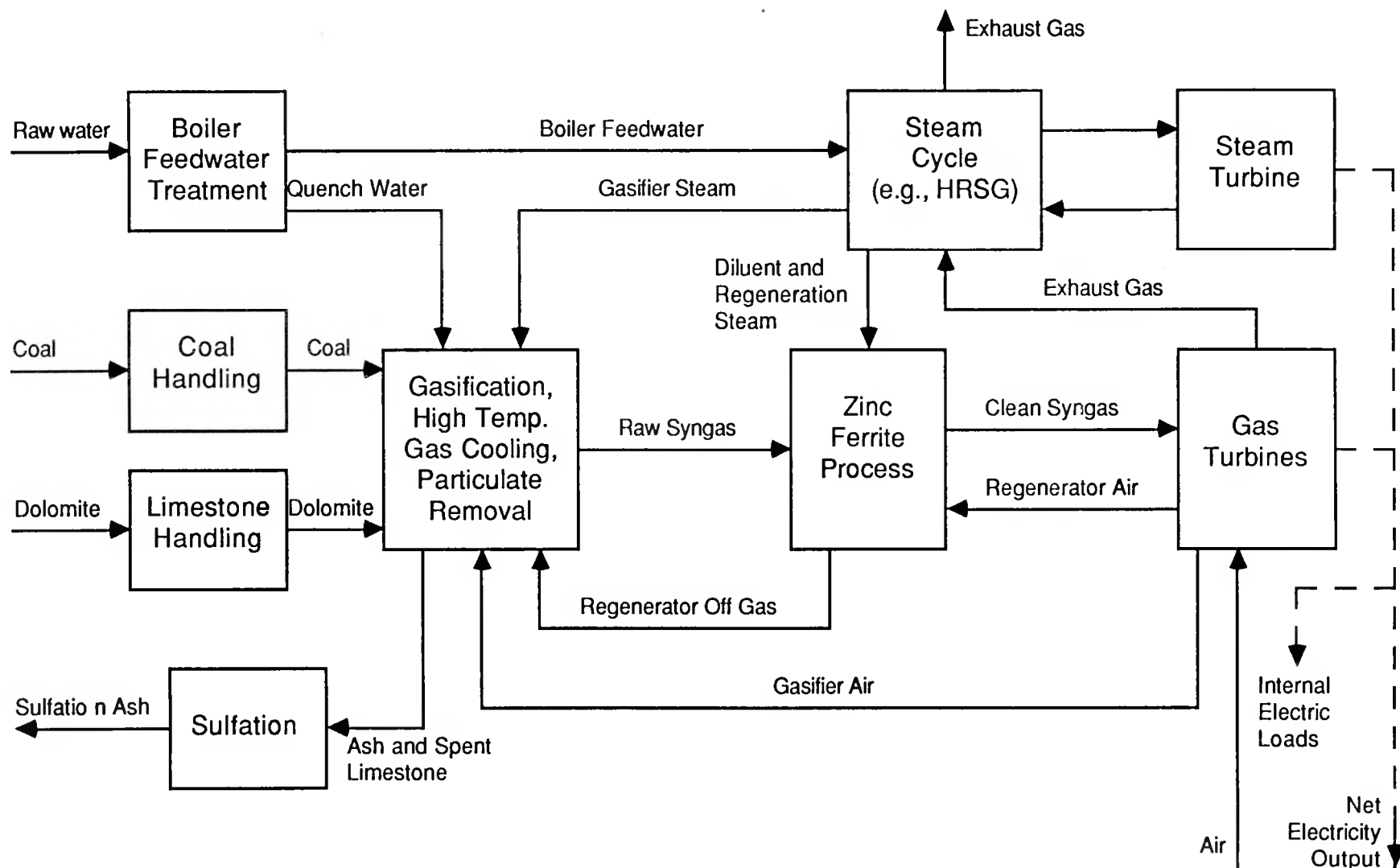


Figure 18. Simple Schematic of Air-Blown KRW IGCC System with Hot Gas Cleanup

The additional direct cost models required for a hot gas cleanup system are presented in the following subsections. The cost correlations presented in this chapter are generally more tentative, with a higher degree of uncertainty, than the cost curves presented in the previous chapter. This is due partly to the limited amount of data available for this advanced technology system, and to the risks of cost growth associated with new process technology.

In this analysis, it is assumed that there is no need for cooling of the fuel gas exiting the desulfurization unit prior to entering the gas turbine. Furthermore, no allowances are made for either steam injection or fuel gas moisturization, because these are not included in the ASPEN flowsheet simulation. Instead, in the ASPEN simulation, raw gas cooling takes place by adding water to the hot syngas, thereby moisturizing the syngas and reducing NOx emissions in the gas turbine. One design study proposes the use of catalytic NOx reduction in the gas turbine exhaust gas, presumably using selective catalytic reduction (SCR) (Earley and Smelser, 1988b). This option is not considered in this study.

The discussion of each process area will provide an overview of the design basis for each section. Earley and Smelser (1988b) provide additional detail on the limestone handling, gasification, zinc ferrite desulfurization, and sulfation process areas. Corman (1986) provides a discussion of air extraction from the gas turbine for oxidant feed to air-blown gasifiers.

3.1 Limestone Handling

Limestone or dolomite are used as reagents for in-bed removal of sulfur compounds in the gasifier reactor. Limestone handling facilities are required to receive limestone, store it, and deliver it to the coal surge hoppers of the gasifier trains. A typical limestone handling design for a commercial scale system includes: an unloading hopper to receive minus 2 inch limestone from a unit train; a conveyor to a storage silo with seven day capacity; a conveyor from the storage silo to an impact crusher to produce minus 12 mesh limestone; two pulverized limestone storage conveyors with a combined capacity of 24 hours; pneumatic transport to the coal/limestone surge bin in the gasification section; and a dust suppression system. Three cost estimates are available for limestone handling systems associated with hot gas cleanup IGCC systems (Smelser, 1986a; Smith and Smelser, 1987; Earley and Smelser, 1988b). From these data a simple linear regression model of direct cost as a function of limestone feed rate was developed:

$$DC_L = 1,160 + 0.026 m_{L,G,i} \quad (R^2 = 0.907; n = 3) \quad (38)$$

The standard error of the estimate for this model is \$655,000. A graph of the model is shown in Figure 19. Because of the small number of data points, the statistical significance of the model is questionable. The significance level for the F-ratio of this model is approximately 0.2, compared to far less than 0.001 for the models developed for the KRW-based IGCC system with cold gas cleanup. Thus, it is not possible to reject the hypothesis that the regression model is statistically insignificant (i.e. that the regression coefficient of the predictive term is zero) based on commonly used significance levels of 0.01 or 0.05. However, it is clear from an engineering standpoint that the cost of the limestone handling system does increase as the limestone throughput capacity increases. Therefore, the inability to reject the hypothesis that the model is statistically insignificant is the result of a small number of data points, rather than due to a lack of relationship between cost and limestone flow rate. The low F-ratio for this model implies a wide confidence interval for the regression coefficient of the limestone flow rate.

Although the coefficients of the regression model are not known with certainty, the cost of the limestone handling system is small compared to other parts of the system. Therefore, uncertainty about the estimated costs for the limestone handling system will be a very small contributor to uncertainty in the total cost of the IGCC system. Furthermore, the technology in the limestone handling system is commercially available, which implies that the limestone handling process area will not be a source of cost growth or technical failure of an IGCC system.

3.2 Oxidant Feed

For air-blown gasification systems, gasifier air is extracted from the gas turbine compressor and is further boosted in pressure using a boost compressor. For this analysis, it will be assumed that the only change compared to an oxygen blown system is the removal of the oxygen plant and addition of an air boost compressor and air precooler for each gas turbine.

Air extraction reduces the amount of air that enters the gas turbine combustor. However, in air-blown gasification systems, the fuel gas contains a substantial amount of nitrogen. Therefore, coal gases from air-blown systems have a substantially lower heating value than coal gases from oxygen-blown systems. The reduced mass flow rate of air to the gas turbine combustor due to air extraction is thus offset (more or less) by the increased mass flow rate of the low-BTU coal gas. Furthermore, in order to maintain the design mass flow rate of exhaust gas through the turbine, the air mass flow rate into the compressor can be restricted through the use of variable position inlet guide vanes (IGV) (NUSCo, 1988). For example, the

Figure 19. Direct Cost of Limestone Handling Section

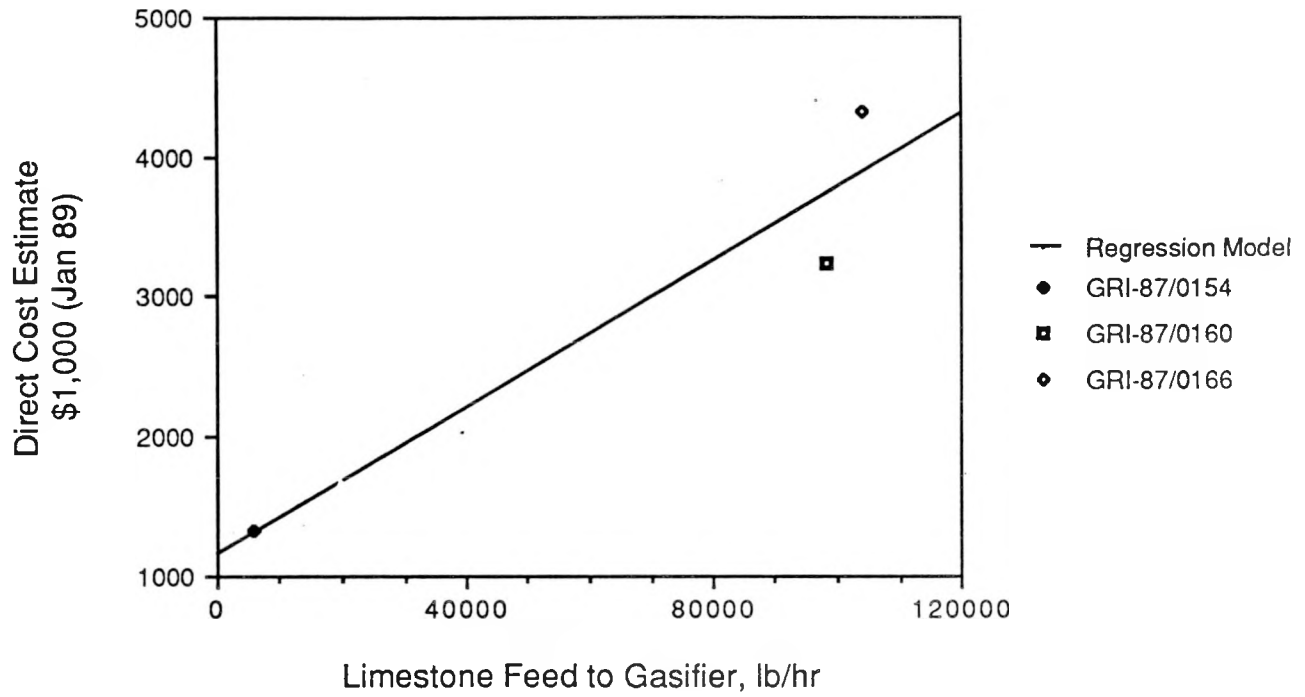
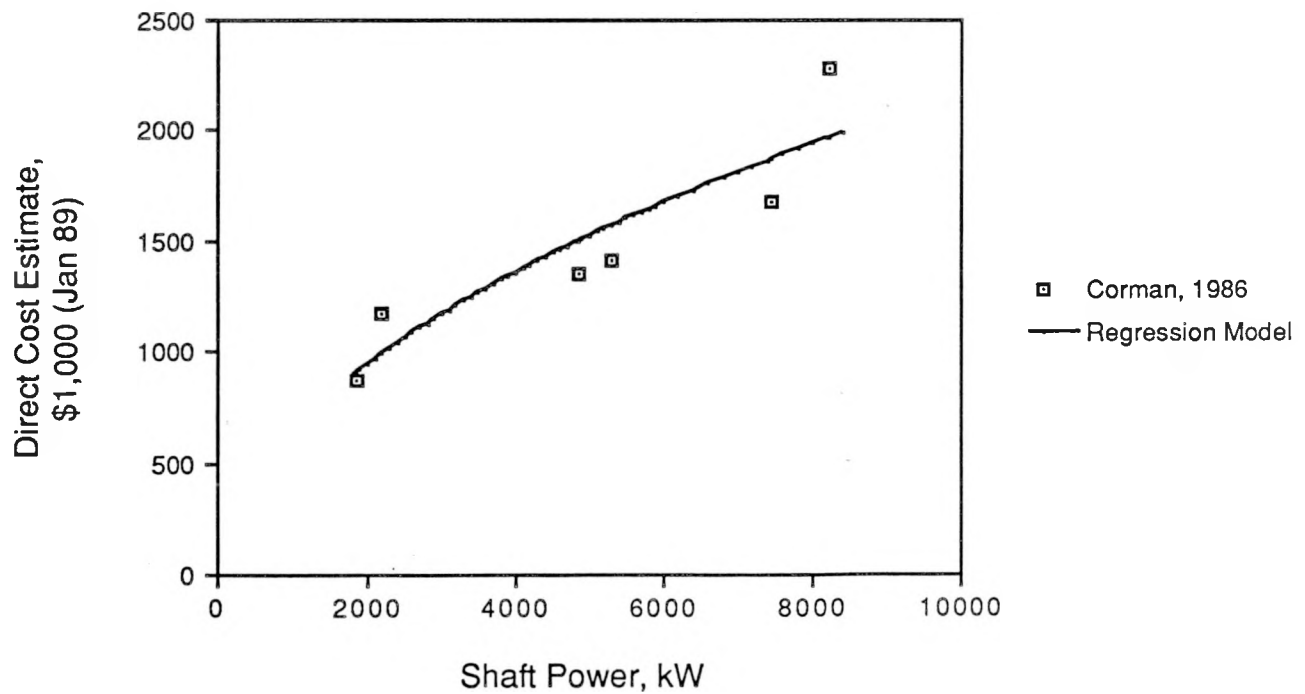


Figure 20. Direct Cost of Air Boost Compressor



exhaust gas flow rate for the GE Frame 7F gas turbine is approximately 920 lb/sec when operating at an ambient temperature of 59 °F. A study of a Shell-based phased IGCC plant indicates a compressor inlet air flow of 888 lb/sec when firing natural gas, which is reduced to 814 lb/sec when co-firing medium-BTU coal gas and distillate oil at the same ambient temperature (VEPCo, 1989). The use of IGVs allows flexibility in gas turbine operation for a variety of fuels. Therefore, it is not expected that the gas turbine will require substantial redesign due to air extraction.¹

In the METC ASPEN simulation of an air-blown KRW-based IGCC system, the boost air compressor is assumed to be a single stage device with an isentropic efficiency of 88 percent (Craig, 1988). The outlet pressure from the GE Frame 7F compressor is approximately 198 psia, assuming a pressure ratio of 13.5. A typical pressure ratio for the air boost compressor is approximately 2 or 3, which can be achieved with a single stage compressor. The ASPEN model calculates the shaft work required for the compressor. This information can be used to estimate the direct cost of the compressor and the electric motor drive for the compressor.

The ASPEN simulation also assumes precooling of the air extracted from the gas turbine prior to boost compression. The cooling occurs in two heat exchangers, one to heat low pressure feed water and the other to heat high pressure feedwater. Based on the heat duty, inlet and outlet temperatures, and an assumed universal heat transfer coefficient, the required heat transfer area can be estimated, and this information can be used to estimate the direct cost of the precoolers.

Several sources of compressor cost data were analyzed. These include a conceptual design study for air-blown Lurgi-based systems with hot gas cleanup (Corman, 1986), the ASPEN Cost Estimating Manual (Schwint, 1986), Chemical Engineering magazine (Hall, Matley, and McNaughton, 1982) and chemical engineering texts (Peters and Timmerhaus, 1980; Ulrich, 1984). Based on this review, the compressor shaft work input was selected as the sizing parameter. The Corman study includes budgetary cost estimates for the boost air compressor and the electrical power requirements for these units. From this information, a simple cost correlation was developed based on the compressor shaft work requirement:

$$DC_{CM} = -98 + 26 N_{T,CM} \left(\frac{W_{s,CM}}{N_{O,CM}} \right)^{0.485} \quad (R^2 = 0.819; n = 6) \quad (39)$$

¹ Modifications may be required to handle the lower heating value fuel gas generated by air-blown systems, as compared to oxygen-blown systems.

where,

$$1,900 \leq \left(\frac{W_{s,CM}}{N_{O,CM}} \right) \leq 8,200 \text{ kW.}$$

The standard error of the estimate is \$230,000. The model is shown graphically in Figure 20 on page 90. The F test of significance for this model indicates that it is possible to reject the hypothesis that the regression coefficient of the predictive term is equal to zero at a 0.01 significance level. Thus, we may conclude that the model is statistically significant. From previous engineering experience, it is possible that this model may under-predict the cost of large compressors. For example, the scaling exponent for compressors from Ulrich (1984) is approximately 0.94, while Peters and Timmerhaus (1980) report approximately 0.7, and Hall et al (1982) shows approximately 0.94. However, unlike these other data sources, the estimate based on data from Corman (1986) is specific for this application. Furthermore, the cost of the boost air compressor is a relatively small component of the total IGCC system direct cost, so the effects of any errors in cost are not expected to be significant. The model should not be extrapolated.

The cost of the boost compressor precoolers was developed based on a cost curve for fixed tube sheet shell and tube heat exchangers in Ulrich (1984), adjusted for compatibility with cost estimates of precoolers in Corman (1986). The precooler is used to reduce the temperature of the air extracted from the gas turbine compressor prior to entering the boost compressor. The extraction air is cooled by heating boiler feed water. The heat exchanger cost depends on the heat transfer area, which in turn depends on the heat flow, universal heat transfer coefficient, and temperature difference as given by:

$$A = \frac{q}{U \Delta T_{LM}} \quad (40)$$

For heat transfer between air and water, a heat transfer coefficient of 40 Btu/(ft²-hr-°F) was assumed based on Ulrich (1984).

The cost curve for fixed sheet shell and tube heat exchangers in Ulrich was approximated by a cubic polynomial. The heat transfer area for the six different size precoolers used in the Corman study was estimated using the assumed heat transfer coefficient and the reported values of air flow, water flow, and air temperatures. The data reported in the cost study was used to determine the heat flux in the heat exchanger and the log mean temperature difference. The estimated equipment purchase cost from the cost curve in Ulrich was then adjusted to reproduce, as closely as possible, the cost estimates in the

Corman study. This was done using a regression analysis of the purchased equipment cost estimated from the Ulrich curve and the reported direct cost estimate for each of the six precoolers in the cost study. The resulting cost curve is shown in Figure 21, which is a graph of the total direct cost of a single heat exchanger versus the heat transfer area. A comparison of the direct cost estimate and the reported cost estimates for the six precooler cases is given in Figure 22. The cost correlation is:

$$DC_{PC} = (24 + 0.067A - 5.5 \times 10^{-6}A^2 + 2.8 \times 10^{-10}A^3)N_{T,PC} \quad (R^2 = 0.972; n = 6) \quad (41)$$

where,

A = Heat transfer area per precooler, ft^2

The standard error of the estimate is \$30,000. The estimate of goodness of fit is based on the data shown in Figure 22. The cost of the precoolers is on the order of \$0.5 million for a 500 MW system. This is not a significant percentage of the total plant capital cost.

The direct cost for the oxidant feed section for an air-blown system is the sum of the boost compressor and precooler costs, and is given by:

$$DC_{OF} = DC_{CM} + DC_{PC} \quad (42)$$

3.3 Gasification

In the currently available design studies for KRW-based systems with hot gas cleanup (Smelser, 1986a; Smith and Smelser, 1987; Earley and Smelser, 1988b), the gasifier is oxygen-blown with in-bed desulfurization using limestone. Although the amount of syngas cooling is reduced compared to cold gas cleanup designs, high temperature syngas cooling cannot be eliminated. The gasifier syngas exit temperature in the detailed studies is 1,850 °F, which must be cooled below 1,200 °F prior to entering the hot gas desulfurization system.

The equipment assumed for the gasification section in the hot gas design studies includes the following:

- coal and limestone pressurization (surge bin, pressurization lockhopper, feed lockhopper)
- gasifier
- solids (ash and spent sorbent) removal
- raw syngas recycling and cyclone
- primary heat exchanger
- particulate filter (sintered metal candle filter or ceramic candle filter)
- gas recycle cooler

Figure 21. Direct Cost of Precooler Heat Exchanger

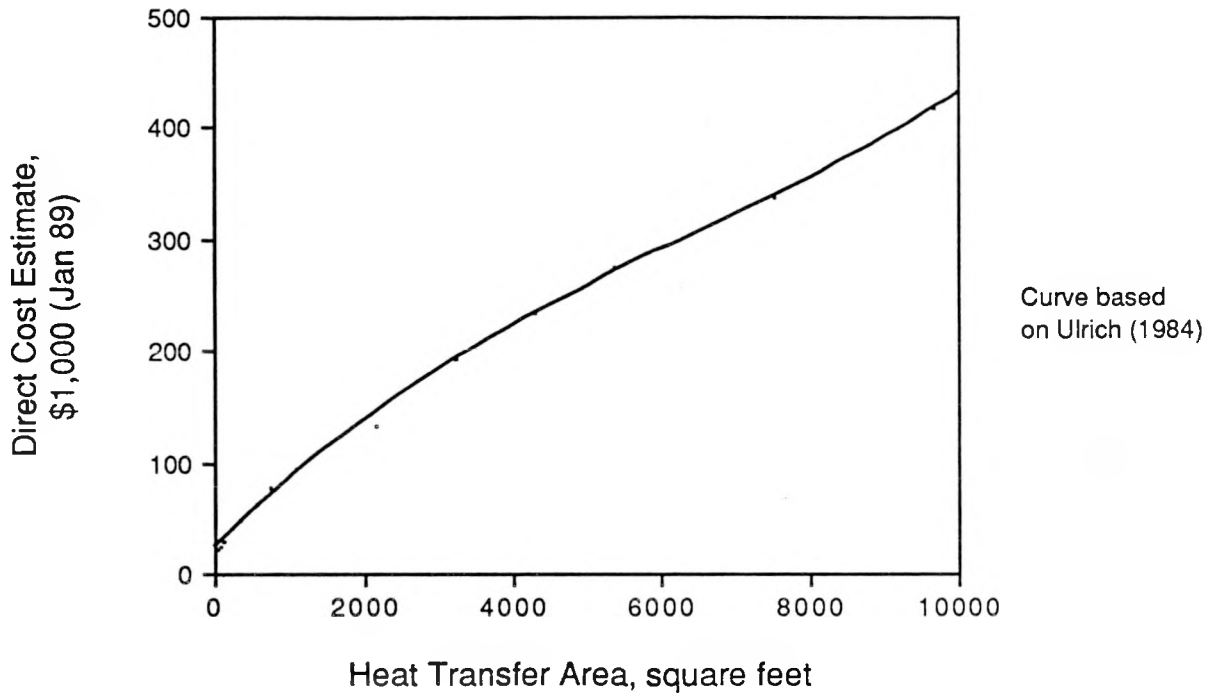
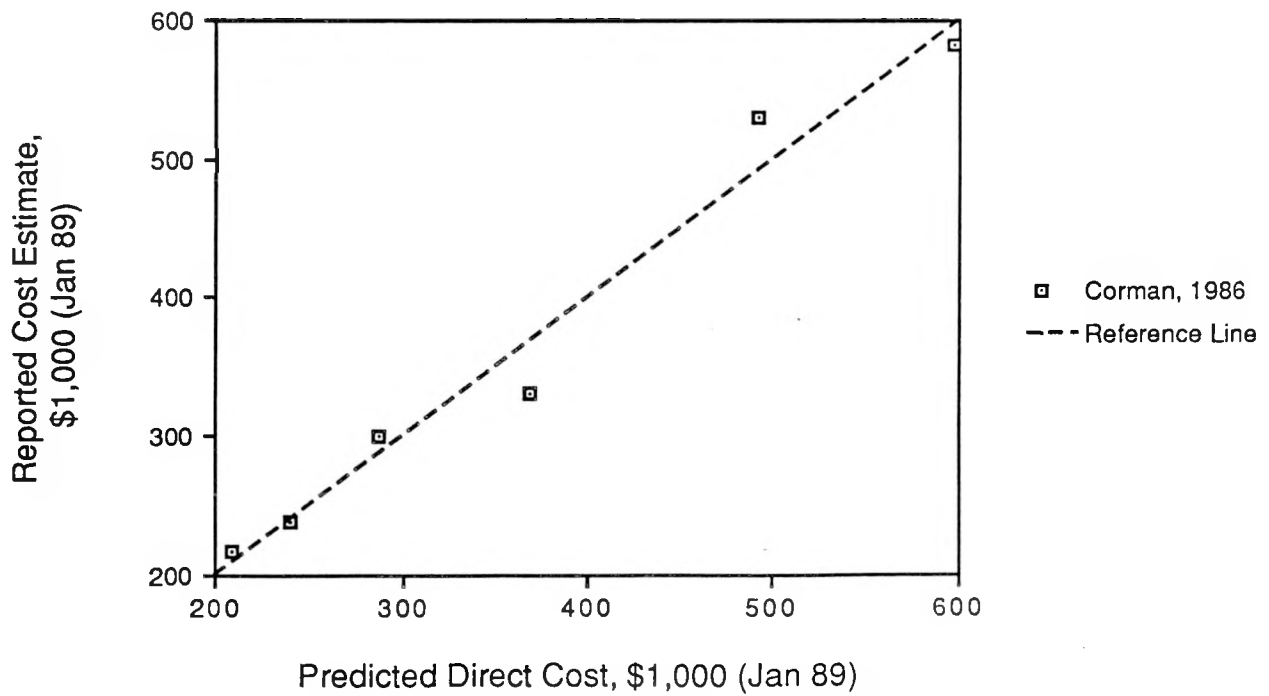


Figure 22. Comparison of Reported vs. Predicted Precooler Costs



secondary gas cooler
cooling gas compressor
recycle gas condenser
recycle gas compressor
knock-out drum
steam drum.

Sulfur removal during gasification using calcium-based sorbents has been demonstrated in the KRW process development unit (PDU). The sulfur absorption rate depends on the molar calcium-to-sulfur ratio (Ca/S) of the sorbent, the type of sorbent material used, and the sulfur content of the coal, among other factors. For a given sorbent, the sulfur removal increases as the Ca/S ratio is increased; however, insufficient data are reported in the literature reviewed to date to develop a regression relationship between sulfur removal and Ca/S ratio. A few different types of calcium-based sorbents have been used in testing, including high calcium, magnesium, and dolomitic limestones. The dolomitic limestones have shown the most consistent sulfur removal performance. The sulfur removal efficiency is limited by equilibrium constraints, and the equilibrium sulfur removal efficiency increases as the coal sulfur content increases (Schmidt, Sadhukhan, and Lin, 1989).

The nominal expected in-bed desulfurization for a high sulfur coal is 90 percent (Haldipur et al, 1988). For dolomitic limestone, this removal rate is commonly assumed to occur at a Ca/S ratio of 1.8 (e.g., Haldipur et al, 1988; Earley and Smelser, 1988). However, the actual Ca/S ratio required to achieve 90 percent sulfur removal for a given coal and sorbent also depends on the residence time in the gasifier. Based on a graph presented by Haldipur et al (1988) for the PDU using a 4.5 percent sulfur coal, 90 percent sulfur removal may be achievable with dolomitic limestone at a Ca/S ratio as low as 1.4. In contrast, for a high calcium limestone, a minimum Ca/S ratio is approximately 2.4, and a conservative assumption would be a ratio of 4.

There is general agreement that calcium-based sorbents catalyze the gasification reactions, increasing the reactivity of eastern coals. For example, the reactivity of Pittsburgh No. 8 coal appears to triple with the use of a limestone sorbent (Floyd and Agrawal, 1989). This implies that carbon conversion efficiencies may be higher with in-bed desulfurization. A sorbent may also reduce the caking tendency of bituminous coals, thereby allowing the gasifier to operate at higher temperatures, which also would tend to increase carbon conversion (Shinnar, Avidan, and Weng, 1988). Based on PDU tests with limestone, the

design operating temperature for the Appalachian clean coal technology demonstration project featuring a KRW gasifier using a high sulfur eastern coal was raised from 1,850 °F to 1,900 °F (Banchik, Buckman, and Rath, 1988).

There is less consensus on the effect of calcium sorbents on the environmental performance of the gasifier. For example, a conceptual design study (Earley and Smelser, 1988) reports that ammonia production is less with in-bed desulfurization than without, particularly for high calcium limestone. But an environmental study by Radian (Scheffel and Skinner, 1988) based on testing of the PDU does not indicate any reduction in ammonia production with dolomite compared to gasification without a sorbent. For the KRW system with cold gas cleanup, ammonia yield from the gasifier is not an air pollution concern because the ammonia is almost completely removed by wet scrubbing. However, for the hot gas cleanup system, the ammonia yield will affect NOx emissions in the gas turbine combustor (e.g., Cincotta, 1984). Earley and Smelser (1988) also report that pilot plant data indicate reduced production of methane with sorbents. The increased gasification rate resulting from use of a calcium sorbent is reported to increase fines consumption in the gasifier and reduce fines elutriation (Banchik, Buckman, and Rath, 1988).

To accommodate in-bed desulfurization, the gasifier vessel may be slightly increased in size compared to no in-bed desulfurization. For example, one study assumed a gasifier size of 101 feet overall length and 14 feet maximum outside diameter without in-bed desulfurization, and 115 feet overall length and 14 feet maximum outside diameter with in-bed desulfurization (Smith and Smelser, 1987). The fluidized bed height for the in-bed desulfurization case is approximately 4 feet higher, due to increased bed volume. The limestone addition results in high levels of ash in the bed and higher bed densities than the conventional gasifier. The higher bed density permits a slightly higher superficial velocity (1.72 ft/s vs. 1.6 ft/s) in the freeboard (uppermost) zone of the gasifier.

Because only three cost estimates for an oxygen-blown coal gasification system with in-bed desulfurization and hot gas cleanup are currently available, the number of variables that can be included in a regression model is limited. The gasifier coal feed rate was identified as the single most important parameter affecting cost. All three data points are based on a Pittsburgh No. 8 coal feedstock. Two alternative simple models of the coal gasification cost were developed:

$$\text{Case 1: } DC_G = 12,200 + 0.205 m_{cm,G,i} \quad (R^2 = 0.999; n = 3) \quad (43)$$

$$\text{Case 2: } DC_G = 16.5 \left(m_{cm,G,i} \right)^{0.68} \quad (R^2 = 0.999; n = 3) \quad (44)$$

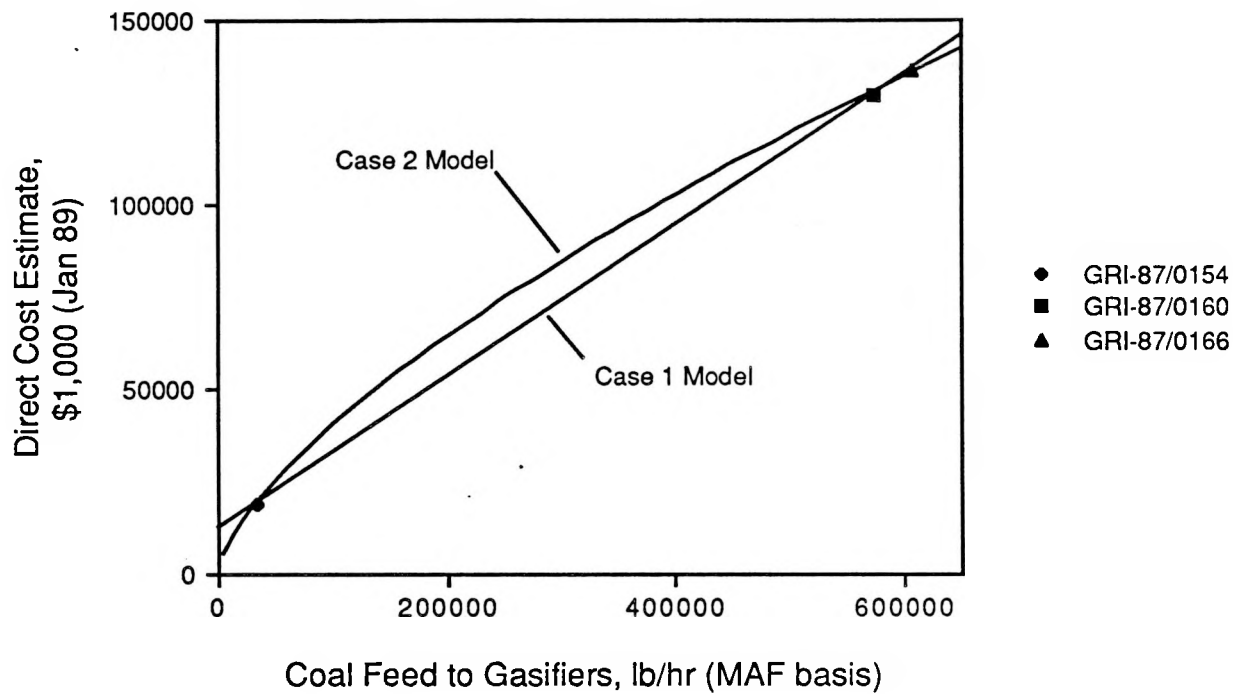
The Case 1 model is a simple linear fit to the data, and the Case 2 model is a traditional "capacity-exponent" model commonly assumed in chemical engineering cost estimation. A graph of the three cost estimate data points and the two alternative cost models is given in Figure 23. The two cost models produce similar results over the range of coal feed rates shown in the figure, with the Case 2 model slightly conservative compared to the Case 1 model for coal feed rates below about 600,000 lb/hr on a moisture and ash free basis. The Case 2 model implies that there is an economy of scale to building larger plants. The standard error of these models is negligible. The models are statistically significant at the 0.01 significance level, based on the F test.

Equations (43) and (44) are valid only for oxygen-blown systems. In order to handle the larger oxidant and syngas volume flow rates, the cost of an air blown system would be higher than a comparable oxygen-blown system. In an air-blown system, the syngas flow rate is substantially increased. The components of the gasification section that are affected by significant changes in the syngas volume flow rate include the cyclones, heat exchangers, and particulate filters. The gasifier vessel dimensions are also affected. The coal feed and ash removal systems would be unaffected.

A correction factor for the cost of an air-blown gasification system compared to an oxygen-blown gasification system can be estimated based on a few assumptions. Comparing results from the two KRW ASPEN simulations (oxygen-blown with cold gas cleanup vs. air-blown with hot gas cleanup), the syngas flow rate for the air-blown system is larger than for the oxygen-blown system by a factor of 2.5. If approximately one-quarter of the total direct cost of the gasification section is affected by increases in the syngas flow rate, and if the relationship between cost and syngas flow rate can be modeled using a capacity-exponent formulation, then it is reasonable to assume, based on the Case 2 model presented as Equation (44), that the scaling exponent is approximately 0.7. If the syngas flow rate increases by a factor of 2.5 for an air-blown system, then the direct cost of an air-blown gasification section is approximately 20 percent greater than for an oxygen-blown system.

There is an additional discrepancy between the ASPEN flowsheet design and the available cost data. The flowsheet assumes gas cooling by quenching the high temperature raw gas with a sufficient quantity of boiler feedwater. The cost data includes high temperature

Figure 23. Direct Cost of Oxygen-Blown Gasification Section



gas cooling using heat exchangers to generate steam. Therefore, the capital cost for this process area would tend to be over-estimated for a given syngas flow rate. Because cost estimates are not directly available for an air-blown system with raw gas quench, some judgment is required to adjust the costs accordingly.

Some basis for the adjustment is available from the studies for Texaco-based IGCC systems, in which alternative gas cooling systems were considered (Fluor, 1984; Fluor, 1986). In place of a radiative syngas cooler and a convective syngas cooler, a case is considered in which the raw gas enters a quench chamber, where water is introduced. From the Texaco study, it appears that the high temperature gas cooling equipment is a significant part of the total cost. In the Texaco system, the raw syngas has a temperature of 2,400 to 2,600 °F, compared to 1,850 °F typically assumed for KRW-based systems. The gas is cooled to 450 °F in the Texaco design. For the KRW-based system with hot gas cleanup, cooling is required only to about 1,100 °F. Therefore, the proportional decrease in cost of the quench vs. radiative plus convective design for the Texaco system is probably higher than for a KRW system with hot gas cleanup. The quench system has a direct cost lower than the radiative plus convective system by a factor of 0.43. We might then assume that for the KRW system, the cost of a quench system is on the order of a factor of 0.75 of the cost of a system with syngas heat exchangers.

For initial analyses, this implies that uncertainty in the cost of the gasification process area due to process configuration can be represented by two multiplicative factors. The first factor adjusts the costs of a system with syngas coolers to the cost of a system with raw gas quench, and is initially assumed to be 0.75. The second factor adjusts the costs of an oxygen-blown system to the cost of an air-blown system, based on the difference in volumetric gas flow rates. This factor is initially assumed to be 1.2. These uncertainties can be resolved by developing a more detailed cost model based on the design assumed in the ASPEN simulation. Such a model is not possible at this time because there is no available cost estimate of an air-blown system with raw gas quench with which to develop a design basis and validate the model. Thus, the preliminary cost model assumed for the air-blown KRW-based system with raw gas quench is:

$$DC_G = (0.75)(1.2)[16.5(m_{cm,G,i})^{0.68}] \quad (45)$$

Clearly, this is an approximate, study-grade estimate. Additional detail on the costs of specific equipment in the gasification section would be required to refine the correction factors or to develop a new model.

3.4 Desulfurization

Three studies of KRW-based systems with zinc ferrite desulfurization have been prepared for GRI (Smelser, 1986a; Smith and Smelser, 1987; Earley and Smelser, 1988b). These studies are based on a fixed-bed design, in which a minimum of two reactor vessels are required. In the fixed-bed design, one vessel operates in the sulfur absorption mode, while the other vessel operates in the sorbent regeneration mode. In the absorption mode, hydrogen sulfide and carbonyl sulfide react with zinc ferrite, and are thus removed from the syngas stream. In the regeneration mode, air and steam are reacted with the spent sorbent, resulting in an off-gas containing primarily water vapor with a concentration of about 5 percent sulfur dioxide. This off-gas is recycled to the gasifier in the KRW-based systems.

The design basis for the zinc ferrite system was significantly changed in the latest of the three studies. The change was in the amount of sulfur loading capacity in the sorbent at which regeneration of the sorbent is required. In the two earlier studies, it was assumed that a nearly theoretical level of over 30 weight percent sulfur in the spent sorbent could be achieved. However, based on recent tests the design basis was updated to a sulfur loading of only 10 percent, substantially increasing the sorbent requirement. As a result, the number of reactor vessels per gasifier train was increased from two (one absorbing and one regenerating at any given time) to four (two absorbing and two regenerating). The likely reason for the increase in the number of vessels (although not stated explicitly in the report) is a constraint on the maximum vessel diameter which is economically transportable (limited by railroad clearances) for shop fabricated pressure vessels, as discussed by Kasper (1988).

Cost estimates for fixed bed zinc ferrite systems were developed as part of analyses of fixed bed gasification IGCC systems (Corman, 1986). However, insufficient definition of the design basis, and aggregation of desulfurization and sulfur recovery costs, precludes use of these estimates for development of cost models. For example, design parameters such as the absorption cycle time and the maximum assumed sulfur loading in the sorbent just prior to regeneration are not given.

A study of hot gas desulfurization processes was performed for METC (Klett et al, 1986). However, in this report, all cost estimates were normalized, with no indication of how the normalized costs convert into absolute dollar costs.

Because a fixed-bed zinc ferrite design is assumed in the cost studies, it is adopted as the basis for a preliminary cost model. Because so few cost estimates of this technology are available, and because the zinc ferrite plant section consists primarily of vertical pressure vessels, a preliminary estimate was developed based on published cost curves for vertical pressure vessels. These costs were used with the KRW-based system studies to estimate costs for sorbent handling, piping, valving, and the control system. The cost model is based on equipment cost curves in Ulrich (1984), a design basis for fixed bed zinc ferrite systems developed by Kasper (1988), and cost and design data from the most recent IGCC study with zinc ferrite hot gas cleanup (Earley and Smelser, 1988b). In addition, a comparison of the cost model prediction and the cost estimate from another hot gas cleanup gasification system study (Smith and Smelser, 1987) was made to verify the model.

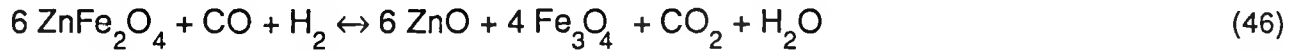
Cost curves for vertical pressure vessels from Ulrich (1984) were used to develop an analytic expression for pressure vessel cost as a function of vessel internal diameter, length, and operating pressure. Assumptions in the design basis governing the maximum allowable syngas velocity through the vessels determine the minimum vessel diameter. The maximum vessel diameter is assumed to be limited to the largest diameter that can be shop fabricated and rail transported, which is 12.5 feet. A length-to-diameter ratio of 4 is assumed for simplicity, representing a reasonable economical aspect ratio (Kasper, 1988).

The number of operating absorber vessels is assumed to be at least as great as the number of operating gasifiers, but may be larger by a factor of two, four, etc, depending on limitations imposed by maximum gas velocity or large sorbent requirements. The cost of auxiliary equipment, such as sorbent handling, piping, valving, and control systems, was estimated by difference between the pressure vessel cost estimate and the total direct cost estimate in the most recent IGCC cost study with hot gas desulfurization (Earley and Smelser, 1988b). This study includes five operating and one spare gasification train. Each train includes a gasifier, associated raw gas handling equipment, and one zinc ferrite process train. Each zinc ferrite process train consists of four vertical reactor vessels, of which two are in an absorption mode and two are in a regeneration mode at any given time. The plant has a total of 24 zinc ferrite vessels, of which ten are in operation during full plant load.

The raw syngas flow rate into the zinc ferrite section in the Earley and Smelser (1988b) study is estimated to be 61,200 lbmole/hr, on a plant basis. Ten zinc ferrite absorbers are operated to remove sulfur from this syngas, with an inlet temperature of 1,125 °F and inlet pressure of 453 psia. Therefore, the volume flow rate per absorber is approximately 64 ft³/sec, assuming that the syngas is an ideal gas.

The absorption cycle is assumed in the study to be 168 hours. The sulfur flow rate in the syngas is 1,734 lb/hr of sulfur, which is in the form of hydrogen sulfide and carbonyl sulfide. Therefore, 291,400 lb of sulfur must be absorbed on a plant basis every 168 hours.

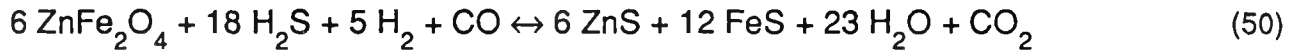
To estimate the amount of sorbent required to absorb this amount of sulfur, a simple relationship was developed based on the stoichiometry of the absorption reactions. The main absorption reactions are:



and it is assumed that all carbonyl sulfide reacts to form hydrogen sulfide according to:



The overall absorption reaction is then:



If 6 moles of zinc ferrite, weighing 1,447 pounds, react completely, then 576 pounds of sulfur are absorbed and the spent sorbent (consisting of ZnS and FeS) weighs 1,638.5 pounds, assuming no inert materials in the sorbent. From these relationships it is apparent that the maximum theoretical sulfur loading is 35 percent, if all the sorbent reacts. However, test results and expectations for commercial scale design are for less than theoretical loading, with long term loading as low as 10 percent (Kasper, 1988). Let the fresh sorbent charge to the operating absorber vessels be S_c , the weight fraction of sulfur loading in the spent sorbent be L_s , the absorption time be t_a , and the weight of the spent sorbent at the end of the absorption cycle be S_s , then:

$$M_{s,ZF,i} = (M_{\text{HS},ZF,i} + M_{\text{COS},ZF,i}) t_a \quad (51)$$

$$m_{s,ZF,i} = M_{s,ZF,i} \left(\frac{32 \text{ lb S}}{\text{lbmole S}} \right) \quad (52)$$

$$S_s = S_c + 192 \left(\frac{M_{s,ZF,i}}{18} \right) \quad (53)$$

$$L_s = \frac{m_{s,ZF,i}}{S_s} \quad (54)$$

substituting Equations (52) and (53) into (54) and simplifying, the following result is obtained:

$$S_c = \left(\frac{32 - 10\frac{2}{3} L_s}{L_s} \right) M_{s,ZF,i} \quad (55)$$

On a mass, rather than molar basis, Equation (55) becomes:

$$S_c = \left(\frac{1 - \frac{1}{3} L_s}{L_s} \right) m_{s,ZF,i} \quad (56)$$

Equation (55) or (56) can be used to determine the initial charge of fresh sorbent required for the operating absorber vessels as a function of the sulfur capture in the absorbers and the long term sorbent loading expected. As a check on the equations, for a maximum theoretical sorbent loading of 35.1 percent and a sulfur capture of 576 pounds, the estimated sorbent charge required is 1,447 pounds, which is the result previously obtained.

At a 10 percent weight loading of sulfur on the spent zinc ferrite sorbent and a sorbent capture of 291,400 pounds per cycle, the total sorbent requirement for the example IGCC plant is estimated to be equivalent to 2.8 million pounds of fresh sorbent charged to the ten on-line absorbers.

The volume of the sorbent charge in each reactor vessel is given by the following equation:

$$V_c = \frac{S_c}{D_b} \quad (57)$$

At a sorbent bulk density, D_b , of 82 lb/ft³ (Kasper, 1988), the total sorbent volume is estimated to be 34,350 ft³, or 3,435 ft³ for each on-line reactor vessel. For simplicity, this is assumed to

be the vessel volume. The design basis developed by METC indicates that the reactor vessels are only slightly larger in volume than the sorbent bed volume.

The minimum diameter is constrained to keep the superficial gas velocity below 2 ft/sec (Kasper, 1988). In ASPEN, the gas volume flow rate can be obtained directly. However, as an approximation for model use outside of ASPEN, the raw syngas is assumed to behave like an ideal gas. The following equation can be used to estimate the minimum acceptable diameter for each zinc ferrite reactor vessel:

$$d_{\min} = 1.128 \left(\frac{M_{\text{syn,ZF},i} (T_{\text{ZF},i} + 460) 10.73}{V_s P_{\text{ZF},i} N_{\text{O,ZF}} 3,600} \right)^{\frac{1}{2}} \quad (58)$$

Kasper (1988) recommends a vessel length-to-diameter ratio of two to four, which is a range of the most economical aspect ratios. Furthermore, Kasper recommends that the vessel diameter should not exceed 12.5 feet, which is an upper limit for the size of rail-transportable shop-fabricated vessels. If a length-to-diameter ratio of 4 is assumed, then the reactor vessel inner diameter is given by:

$$d = \left(\frac{V_c}{\pi} \right)^{\frac{1}{3}} \quad (59)$$

If the estimated vessel diameter is smaller than the minimum diameter, the number of absorber vessels should be decreased. If the estimated diameter is larger than 12.5 feet, the number of absorber vessels should be increased.

Using the above relations for the example IGCC case, the dimensions of each vessel are estimated to be 10.3 feet inner diameter and 41.2 feet minimum length. The estimated diameter is larger than the required minimum diameter of 6.4 feet, so no adjustment in the number of vessels is required.

To estimate the cost of a range of reactor vessel sizes for the zinc ferrite system, the cost curves in Ulrich (1984) have been approximated as analytic expressions using regression analysis. The cost of the pressure vessel depends on the operating pressure, diameter, and length. The diameter and length are determined based on the sorbent

requirements, which in turn depends on an assumption regarding sulfur loading, as indicated in the previous discussion. The cost of the auxiliaries is assumed to scale with the cost of the reactor vessels; for simplicity it is assumed that the cost of the auxiliaries is a simple multiplier of the cost of the pressure vessels. This assumption can be refined if more data become available.

The method for estimating pressure vessel cost is first to estimate the purchased cost of a vertical vessel for operation at low pressures, then to adjust the cost for higher pressures, and finally to adjust the cost to include installation costs. Eight curves for the purchased cost of a low pressure vertical vessel are given in Ulrich (1984, p 307), one for each of eight diameters as a function of vessel height. These curves were approximated using linear regression models, of the form:

$$y = a + bH \quad (60)$$

The dependent variable in these estimates was the purchased cost of the vessel, and the independent variable was the vessel height. The coefficient of determination of these estimates ranged from 0.988 to 0.999. The parameters of Equation (60) are a function of the vessel diameter. A regression analysis of the slope and intercept of all eight linear cost approximations using a third degree polynomial yielded a single analytical expression for pressure vessel cost as a function of diameter and height, in units of feet. This expression for a single vessel (in thousands of June 1982 dollars) is:

$$PC_{ZF,v} = (4.1 - 4.35d + 0.958d^2 - 0.0391d^3) + (0.23 + 0.166d - 0.018d^2 + 7.87 \times 10^{-4}d^3)H \quad (61)$$

The cost estimate from Equation (61) is adjusted to January 1989 dollars using a multiplier of 1.122, based on the *Chemical Engineering* plant cost index. To adjust the cost for increased pressure and installation costs, the following correction factor was developed based on curves presented in Ulrich (1984, p. 308):

$$f_{p,i} = 2.5 + 0.093 \left(P_{\text{syn,ZF,i}} \right)^{0.64} \quad (62)$$

The direct cost of the zinc ferrite plant section also includes piping, valves, and a control system. These auxiliary items were estimated by difference between the vessel cost and the total direct cost estimate presented in Earley and Smelser (1988b). As already discussed, 24

reactor vessels are assumed in the detailed cost study. The total cost for pressure vessels is estimated to be \$10.6 million, in January 1989 dollars. This compares with a reported total direct cost of \$14.1 million, implying that the cost of auxiliaries is \$3.5 million, or 32 percent of the vessel costs.

An attempt was made to verify the cost model against other cost estimates. However, limited design data is available for the study by Corman (1986) and the exact value of the weight percent sulfur loading used in the Smith and Smelser (1987) study is not known, except that it is above 30 percent (Earley and Smelser, 1988b). The upper limit is 35 percent, as already discussed. The cost model was applied to try to reproduce the results of the Smith and Smelser (1987) study.

For the Smith and Smelser study, it is assumed that a sulfur loading of 32 percent was used. The syngas flow rate and sulfur flow rate for each vessel are given, which are very close to the values used in the Earley and Smelser (1988b) study, as is the assumed cycle time of one week. There are five operating vessels and a total of twelve vessels, associated with five operating and one spare gasifier. For each gasifier, there are two zinc ferrite vessels, allowing one to operate in the absorption mode and the other in the regeneration mode. The cost estimated from the equations presented above is \$4.7 million in June 1985 dollars, including the multiplier of 32 percent for auxiliaries. The cost estimate in the GRI study is reported with only one significant figure as \$4 million, also in June 1985 dollars. These costs agree fairly well, and also indicate the sensitivity of the zinc ferrite equipment capital costs to the long term attainable sulfur loading in the sorbent.

The direct cost model for the zinc ferrite section can be summarized in the following equation:

$$DC_{ZF} = 1.477 f_{p,i} [A(d) + B(d) H] N_{T,ZF} \quad (63)$$

where

$$\begin{aligned} f_{p,i} &= \text{from Equation (61);} \\ A(d) &= 4.1 - 4.35d + 0.958d^2 - 0.0391d^3; \\ B(d) &= 0.23 + 0.116d - 0.018d^2 + 7.87 \times 10^{-4}d^3. \end{aligned}$$

As previously discussed, the vessel diameter should not exceed 12.5 feet, and the minimum diameter for which this model is valid is 1.6 feet. The factor 1.477 in Equation (63) includes the multiplier of 1.324 representing the cost of auxiliary equipment. It also adjusts the cost year from mid 1982 to January 1989. The vessel dimensions are determined based on the sorbent requirement per vessel, which depends upon the sulfur flow rate in the syngas, the

absorption cycle time, and the long term sulfur loading capacity of the sorbent. The number of vessels should be selected so that the superficial velocity of the syngas is below a threshold value (e.g., 2 ft/sec), and so that the diameter of the vessels is small enough (e.g., less than 12.5 feet) to be shop fabricated and rail transported to the site.

3.5 Sulfation

For the air-blown KRW-based system with in-bed desulfurization, the spent sorbent is a waste stream. This sorbent contains calcium sulfide (CaS) which is not acceptable for landfilling. The purpose of the sulfation unit is to oxidize the calcium sulfide to the more stable calcium sulfate (CaSO₄), which is then suitable for disposal. The spent sorbent is contained in the gasifier ash waste stream, which also includes unconverted carbon. Therefore, a sulfation unit can recover the heating value of unconverted carbon. A circulating fluidized bed combustor has been proposed for the sulfation application. The heat released from the sulfation reaction and combustion of unconverted carbon is used to generate steam. However, the current ASPEN simulation does not include a model of this system; therefore, the recoverable chemical energy in the ash and spent limestone leaving the gasifier are not included in the plant energy balance.

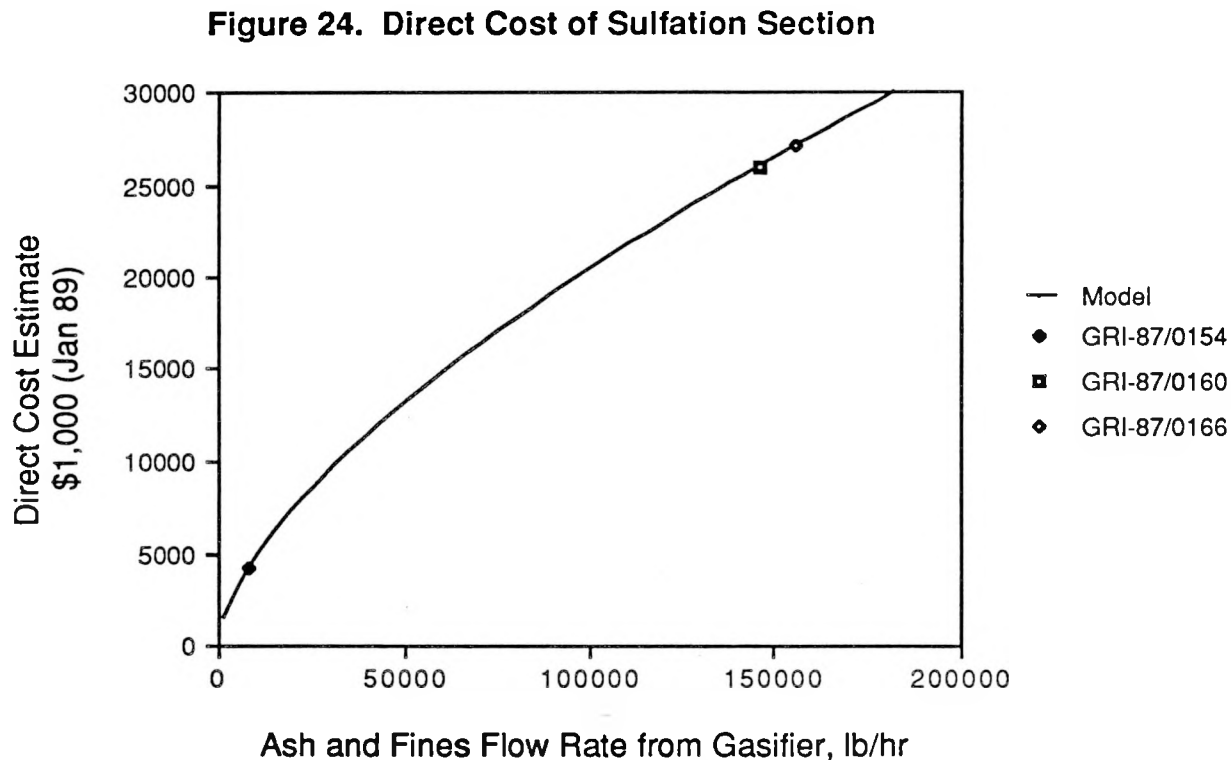
The sulfation unit is an additional emission source. Fluidized bed combustors generally have more uniform flame temperatures than conventional types of combustion systems; therefore, the NO_x emissions from these systems are comparatively low. The sulfur emissions from the sulfation unit depend on the sulfur content of the feed stream, which is expected to be low. The SO₂ and NO_x emission rates are expected to be 0.01 lb SO₂/MMBtu and 0.15 lb NO_x/MMBtu, respectively. These are well below the NSPS limits for steam generators fired with bituminous coal. Particulate emissions are controlled by a fabric filter to less than 0.03 lb/MMBtu (Earley and Smelser, 1988).

There is uncertainty regarding the effectiveness of the sulfation unit in converting calcium sulfide to calcium sulfate. A test of ash and spent limestone in a research reactor yielded only 25 to 35 percent conversion of the calcium sulfide at solids residence times from one to two hours. A proprietary additive has been formulated to improve the conversion rate, and it is expected to result in 85 percent conversion of the calcium sulfide. However, no information is currently published on the amount required or the cost of the additive. Furthermore, there is currently no assessment of the effect of an incomplete conversion rate or of the additive on the environmental characteristics of the sulfation unit waste stream.

A simple cost curve for the sulfation section was developed based on the three available KRW-based gasification system studies with hot gas cleanup. This model is:

$$DC_{SF} = 13.0 (m_{ash} + m_{fines})^{0.639} \quad (R^2 = 1.00; n = 3) \quad (64)$$

The standard error of the model is negligibly small. The F ratio of this model is significant at the 0.001 significance level. A graph of this model is shown in Figure 24.



4.0 CAPITAL COST OF A LURGI-BASED SYSTEM WITH HOT GAS CLEANUP

The plant design assumed as the basis for cost model development is illustrated in Figure 25. This design is based on an ASPEN flowsheet simulating an air-blown Lurgi gasifier IGCC system with hot gas cleanup (Klara, Rastogi, and Craig, 1988), and on a design study of simplified IGCC plants (Corman, 1986). However, there are some differences between the design bases of the ASPEN simulation and the available cost data. The largest of these differences is in the design of the high temperature desulfurization process. The flowsheet is based on a moving bed zinc ferrite system, whereas the available cost data are for fixed bed systems.

In the air-blown Lurgi system with hot gas cleanup, coal gasification is accomplished in a fixed-bed, dry ash Lurgi gasifier. The Lurgi dry-ash gasifier, which was developed in the 1930's, features a counter-current flow of coal past the steam and oxidant feed streams. This results in a widely varying temperature profile through the reactor, and also in a low syngas exit temperature compared to other gasifiers (e.g., KRW). The Lurgi gasifier has only a limited ability to handle coal fines, which tend to become entrained in the exiting syngas or deposited in the exiting tars (which are a characteristic product of this design). Because entrained fines represent a loss of energy efficiency (due to unconverted carbon), typical designs for Lurgi systems involve sizing of the coal input to the gasifier. The typical coal feed is between 1/4 and 2 inches in size (SFA, 1983). Alternatively, a portion of the coal may be fed to the gasifier in the form of fines (coal particles less than 1/4 inch in size). Provisions for cyclones to collect the fines that pass through the gasifier, and the agglomerating and recycling of fines to the gasifier, are necessary to improve carbon conversion efficiency. This option is assumed in the available cost study, and is therefore adopted as a basis for the cost model.

Because the Lurgi gasifier has a relatively low syngas outlet temperature (typically near 1,000 °F), high temperature syngas cooling is not required prior to hot gas cleanup. The plant design assumed here includes high temperature particulate removal using high efficiency cyclones, and high temperature desulfurization using the zinc ferrite process, as assumed in a recent study (Corman, 1986). The clean syngas is then fed directly to GE Frame 7F gas turbines. The exhaust from the gas turbines passes through heat recovery steam generators, which supply steam for the steam turbine, the zinc ferrite process, and the gasifier. Bleed air from the gas turbine is used as the oxidant in the gasifiers. The off-gas from the zinc ferrite process is delivered to a sulfuric acid plant for byproduct recovery.

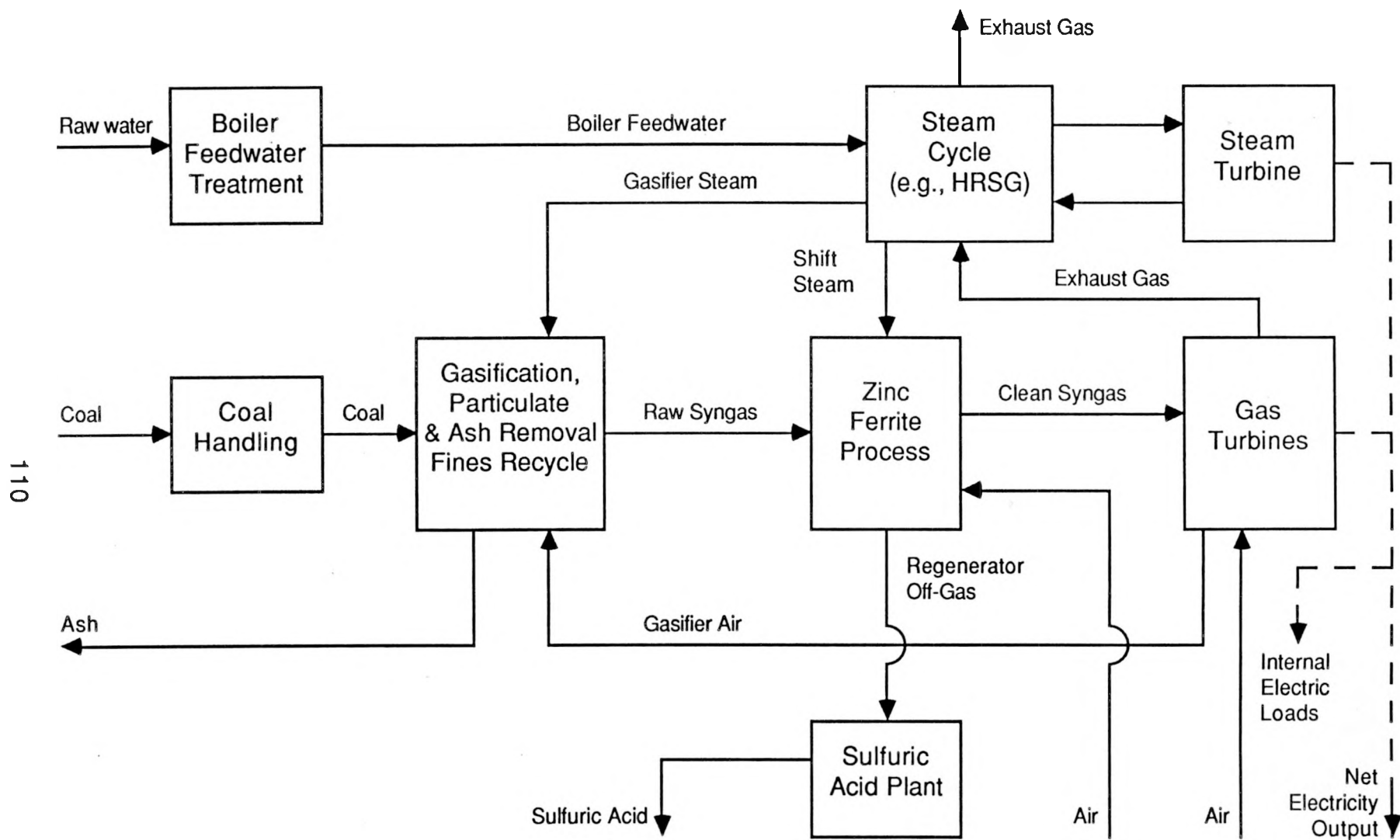


Figure 25. Simple Schematic of Air-Blown Dry-Ash Lurgi IGCC System with Hot Gas Cleanup

An important difference between the moving bed zinc ferrite design assumed in the performance model and the fixed bed zinc ferrite design for which detailed performance and cost information are available is the regeneration system. In the moving bed design, regeneration occurs continuously, using air as a reactant and recycled off-gas as a diluent to prevent excessive temperatures from sintering the sorbent. The regeneration reactions occur at a lower pressure than the desulfurization reactions; therefore, a separate air compressor is assumed for direct feed of ambient air to the regenerator (based on the ASPEN simulation). The off-gas contains mostly nitrogen and approximately 10 percent sulfur dioxide, which can be fed directly to a sulfuric acid plant. In the fixed bed design, regeneration occurs in cycles, using steam as a diluent and producing an off-gas stream containing primarily water vapor and nitrogen. Air required for the regeneration reactions is supplied by the air booster compressor. The sulfur dioxide concentration is typically below five percent (Corman, 1986). While sulfuric acid plants have processed off-gases containing as little as three percent sulfur dioxide in the copper smelting industry (EPA, 1981), the costs increase significantly as the off-gas volume flow rate increases.

As an approximation, the zinc ferrite cost model developed in Section 3.0, which represents a fixed bed design, is assumed in the cost basis for the desulfurization section of an air-blown Lurgi system with hot gas cleanup. At such time when detailed performance and cost data are developed for the moving bed zinc ferrite system, a more applicable cost model can be developed.

Many of the direct cost models developed in Sections 2.0 and 3.0 are applicable to air-blown Lurgi-based systems with hot gas cleanup. From Section 2.0, the applicable direct cost models are:

1. Boiler Feedwater System (Area 80);
2. Gas Turbine (Area 91);
3. Heat Recovery Steam Generator (Area 92);
4. Steam Turbine (Area 93);
5. General Facilities (Area 100).

From Section 3.0, the applicable direct cost models are:

1. Air Boost Compression (Area 10);
2. Zinc Ferrite Desulfurization (Area 50).

In this section, additional direct cost models are presented for the following plant sections:

1. Coal Handling (Area 20);
2. Gasification (Area 30);
3. Coke, Fines, and Ash Handling (Area 31);
4. High Temperature Cyclones (Area 32);
5. Sulfuric Acid Plant (Area 60).

The direct cost models for each of these additional plant sections are presented below. These models are intended only for the purpose of estimating the cost of an air-blown Lurgi-based IGCC system with hot gas cleanup.

4.1 Coal Handling

Lurgi gasifiers are constrained in the amount of coal fines that can be included in the gasifier coal feed, due to entrainment of fines in the exit gas and possible instabilities in the gasifier bed (SFA, 1983; Corman, 1986). To accommodate these limitations, which apply to both dry-ash and slagging Lurgi gasifiers, four general approaches have been used in different studies for utilizing coal in Lurgi-based plants. These include: (1) separation and return of all or a portion of the fines to the coal supplier (e.g., Cover et al, 1985b; Smelser, 1986b and 1986c); (2) separation of sized coal for use in Lurgi gasifiers and fines for use in other types of gasifiers, or for combustion in conventional boilers (e.g., Cover et al, 1985b; Smelser, 1986b and 1986c); (3) separation of a portion of the fines for injection directly into the gasifier bed via "tuyeres" (e.g., Bechtel, 1983a); and (4) agglomeration of separated coal fines into briquettes for feed to the gasifier coal surge bin (e.g., Parsons, 1985; Bechtel et al, 1988). All systems require coal receiving, storage, and reclaim facilities. All systems typically include some type of screening to separate coal fines (less than 1/4 inch) from the sized coal (typically 2 x 1/4 inch). The systems vary in the size of the received coal, with many studies assuming that the coal is two inches or smaller when received. In a few studies, larger sized coal is first crushed to the minus two inch size recommended for use in Lurgi gasifiers (Bechtel, 1983; Zahnstecher, 1984).

The assumption in a recent General Electric study (Corman, 1986) is that the coal is received as minus two inch size, and that all of the coal, including fines, are fed directly to the gasifier. In this study, only the fines that are captured from the raw syngas in a primary cyclone are agglomerated into briquettes, using bentonite (a type of clay) as a binder. In other studies that include fines agglomeration, the fines in the coal feed are agglomerated prior to entering the gasifier. The coal assumed in the General Electric study is an Illinois No. 6, which is a caking coal. It is reported that up to 30 percent fines by weight in the coal feed could be handled in the Lurgi gasifier if the coal is a caking type and if the small fines (less than 1/8 inch) met certain size criteria, which are not reported. Corman reports that at most 5 percent of the coal feed is expected to carry-over in the form of fines recovered in the primary cyclone.

The Corman study scaled cost estimates for the coal handling section from a previous study performed by General Electric (Cincotta, 1984). In the Cincotta study, it is assumed that sized coal (2 x 1/4 inch) is delivered by a dedicated unit train operating between the power

plant and a supplier of sized coal. The coal handling system includes an unloading hopper for railroad bottom dump cars, vibrating feeders for the hopper cones, a dual train of belt conveyors between the vibrating feeders and the sampling system, and a belt conveyor from the sampling system to the stacker. Two 100 percent coal reclaiming systems are included. Coal is reclaimed by gravity feed from the active storage pile to underground reclaim hoppers. A vibrating feeder supplies a reclaim belt conveyor to a series of belt conveyors that distribute the coal to the gasifier surge bins using traveling trippers.

Corman (1986) reports that the coal handling section costs can be estimated based on the coal feed mass flow rate (lb/hr basis) to the gasifiers and the following relationship:

$$DC_{CH} = 4.3 \times 10^{-4} (m_{cf,G,i})^{0.85} \quad (65)$$

The reported costs of this coal handling system are high compared to the cost reported for coal handling for the KRW-based systems (e.g., for a 420,000 lb/hr system, the KRW coal handling cost is estimated to be \$24 million compared to \$26 million for the Lurgi system). This is unreasonable because, in addition to coal receiving, storage, and reclaiming, the KRW-based coal handling section also includes systems for coal pulverization and coal drying. Furthermore, the General Electric estimate is high compared to the estimates prepared for the Gas Research Institute, which include crushing of minus six inch coal to minus two inches, plus fines separation. However, because no detail on equipment costs is given in any of the studies, it is not possible to determine which study is "correct".

For the purpose of comparing Lurgi and KRW systems, it is important that the relative costs be reasonable. Therefore, a cost correlation based on the Gas Research Institute estimates was developed. This provides consistency with the data used in developing the coal handling cost curve for the KRW-based systems, which included data from four studies prepared for GRI. The fact that that GRI estimates include allowances for coal crushing, fines separation, and fines export can be considered as conservative in cost for a system intended to receive sized coal and deliver all of the coal, including fines, to the gasifiers. Such a coal handling system may be required, however, if the gasifiers are unable to perform as predicted with large fines loading in the feed coal. Furthermore, such a coal handling system allows receipt of less expensive coal, such as minus 6 inch, rather than requiring that the coal be sized prior to delivery to the plant.

An additional data point for the coal handling systems was obtained from a study by Bechtel (Bechtel, 1983a) for a plant size better suited for IGCC systems. Using all available

data, a simple relationship between the cost of the coal handling section and the coal feed rate to the gasifier was developed:

$$DC_{CH} = 11,200 + 1.70 m_{cf,CH,i} \quad (R^2 = 0.943; n = 11) \quad (66)$$

where,

$$5,000 \leq m_{cf,CH,i} \leq 36,000 \text{ tons/day}$$

The standard error of the estimate is \$5 million. Unlike the cost equation developed for KRW-based systems, this one is based on the coal feed into the coal handling system, rather than into the gasifier. However, in the case where all of the coal, including fines, are fed to the gasifier, the as-received mass flow and the gasifier feed mass flow are equivalent. A comparison of the cost data and the regression model is shown graphically in Figure 26. The data used to develop the model in Equation (66) include lignite and Illinois No. 6 coals.

4.2 Gasification

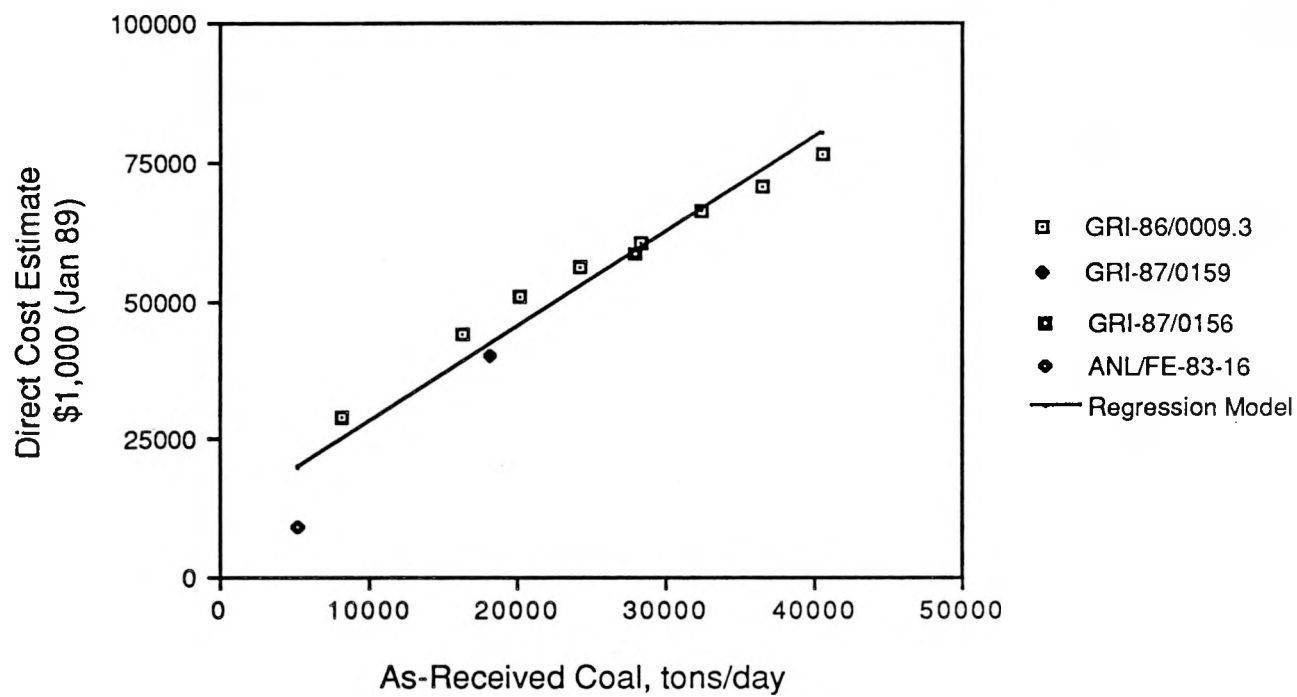
The only currently available detailed design study of a commercial-size air-blown Lurgi-based IGCC system with hot gas cleanup is Corman (1986). Other studies of similar Lurgi-based systems have been done, but either no details are given on the development of performance and cost estimates (e.g., Craig and Koch, 1988), or too little information is reported from which to develop a cost model (e.g., Klett et al, 1987).

For the KRW-based systems, the gasification section costs include the gasifier pressure vessels, coal feed hoppers, ash removal hoppers, particulate removal systems, and other equipment. In the Corman study, costs are reported separately for three component parts of the gasification system. These are: (1) the gasifier pressure vessels, coal feed hopper, and ash removal hoppers; (2) coke handling, fines recycle and agglomeration, and ash removal from the ash lockhopper; and (3) particulate removal from the syngas using high temperature, high pressure cyclones. The costs for these gasification subsections are treated separately here.

4.2.1 Gasifier and Associated Subsystems

From Corman (1986) and Zahnstecher (1984), the cost per train of a gasifier subsystem can be estimated. The gasifier subsystem includes a coal surge bin, the gasifier pressure vessel, and an ash lockhopper. The coal surge bin is centrally mounted on top of the gasifier pressure vessel. Coal is charged to the surge bin from the coal handling system, and the coal flows from the surge bin to a rotating distributor plate in the gasifier. The gasifier vessel

Figure 26. Direct Cost of Coal Handling for a Lurgi-Based System



includes a jacket for generation of steam. Ash is removed from the gasifier as a dry solid to a centrally located ash lockhopper located at the bottom of the gasifier.

The typical gasifier design assumed as a basis in cost studies is a Lurgi Mark IV system. This is a commercial offering for oxygen-blown systems. It has a nominal diameter of 12.7 feet, which is at the upper limit of rail-shippable vessel size. The cost per unit of these systems is approximately the same in several cost estimates. However, the capacity of each gasifier, and therefore the total number of gasifiers required for a given plant, depends on the coal throughput that can be handled per vessel. For an Illinois No. 6 coal, the coal throughput estimated by Lurgi, as reported by Corman, is 300 lb/hr-ft² of grate area for a system operating at 300 psia, and 500 lb/hr-ft² of grate area for a system operating at 600 psia. The grate area can be approximated based on the working diameter of the vessel, which can be assumed to be 12.7 feet. Thus, the coal throughput per gasifier operating on Illinois No. 6 coal may be estimated as follows, assuming a simple linear model:

$$CTP = 127 \left(100 + \frac{2}{3} P_{\text{syn,G,o}} \right) \quad (67)$$

Corman states that the throughput values estimated by Lurgi may be optimistic for air-blown systems. Therefore, in actual model applications, the coal throughput estimated by Equation (67) should be viewed as an upper bound. This equation is only applicable for the Mark IV gasifier operating with Illinois No. 6 coal and gasifier pressures ranging from 300 to 600 psia. The coal throughput per gasifier is in units of lb/hr.

The cost of the coal surge bin, gasifier vessel, and ash lockhopper is approximately \$7.7 million (Jan 89) per train based on three estimates for the Mark IV design presented in Corman (p III-3). The cost of the gasifier system from Zahnstecher (1984) is approximately 7.85 million (Jan 89) per train; however, this is for a conventional oxygen-blown system with cold gas cleanup, including a raw gas wash cooler to remove tars and oils which would condense on downstream equipment. Because gas cooling is not required in hot gas cleanup systems, condensation of tars and oils is not expected to pose a problem; therefore no raw gas wash cooling is required. Furthermore, the heating value of the syngas is increased by the presence of tars and oils in the vapor phase, compared to the syngas in a cold gas system in which tars and oils are removed. Therefore, the cost estimate from Corman will be used. The number of operating trains of gasifiers is determined based on the coal throughput per gasifier. The number of total trains is determined based on the number of operating trains and the availability of the gasifier, which is 87 percent (Zahnstecher, 1984). Thus:

$$N_{O,G} = \text{INT}\left(\frac{m_{cf,G,i}}{CTP}\right) \quad (68)$$

$$N_{T,G} = \text{INT}\left(\frac{N_{O,G}}{AV}\right) \quad (69)$$

where $AV = 0.87$. The function INT indicates that the number of trains should be rounded to the nearest integer. The direct cost equation for the gasification section will be presented after the discussion of coke, fines, and ash handling and particulate removal using cyclones.

4.2.2 Coke, Fines, and Ash Handling

The gasifier train costs do not include coke handling, fines briquetting and recycle, or ash removal. Coke is required as a startup fuel for Lurgi gasifiers. A typical coke handling system would include a receiving hopper for coke delivered by truck, and a means for reclaiming the coke and delivering it to the gasifier during startup. A similar system is required for handling bentonite, which is used as a binder for fines agglomeration. The fines agglomeration section includes fines surge bins, binder storage, mixers, roll briquetters, and conveyors. Briquetted fines are delivered to the gasifier coal surge bin for feed back to the gasifier. No detail on binder requirements are given in the Corman study. Ash removal includes facilities to removal ash from the gasifier ash lockhopper and convey it for disposal (e.g., to trucks). The costs of these systems are included in the Corman study, but are not detailed. They are estimated here by difference between the total direct cost reported for the gasification section of the plant and the total direct cost of the gasifier vessels and associated equipment. The cost of the coke handling system is expected to be proportional to the size of the gasifier, as measured by the coal feed rate. The cost of the agglomeration system is expected to be proportional to the fines recycle rate (which, for the Corman study, is proportional to the coal flow rate). The cost of the ash removal system is expected to be proportional to the ash content of the coal and the coal feed rate to the gasifier. A simple linear model of the cost of the coke, agglomeration, and ash removal subsystems as a function of plant coal feed rate was developed:

$$DC_{SS} = -1,200 + 2.5 m_{cf,G,i} \quad (R^2 = 1.00; n = 4) \quad (70)$$

where,

$$1,000 \leq m_{cf,G,i} \leq 5,200 \text{ tons/day}$$

The standard error of this estimate is negligible. A graphical representation of the sum of the costs of these subsystems and the coal feed rate to the gasifiers is given in Figure 27 on page 119. This model is applicable only to Illinois No. 6 coal.

4.2.3 High Temperature Particulate Removal

In the Corman study, particulate removal occurs without gas cooling and in two stages. Primary cyclones, located between the gasifier outlet and the zinc ferrite inlet, are used to remove approximately 98 percent of the particulates in the raw syngas. Secondary cyclones, located between the zinc ferrite absorber outlet and the gas turbine inlet, are used for removing any entrained sorbent material from the zinc ferrite process as well as a significant portion of the remaining particulates not removed in the primary cyclone. The cost estimate is based on single-stage cyclones offered by General Electric, which consist of a stainless steel cyclone enclosed in a refractory lined carbon steel pressure vessel (Corman, 1986). The inner diameter of the cyclones ranges from 50 to 60 inches in the six cases considered in the study.

Each cyclone unit consists of a cyclone, solids lockhopper, solids ball valve, and support structural steel. The cost of each unit is expected to depend on the syngas volume flow rate, the syngas pressure, and to a lesser extent, the solids throughput. A multi-variate regression of the six cost estimates for cyclone systems was developed based on the syngas volume flow rate through each cyclone, and the operating pressure of the gasifier (which was assumed as the design pressure for the cyclones). For a particulate removal system in which both primary and secondary cyclones are used, the volume flow rate of the syngas through each cyclone is equal to the total syngas volume flow rate divided by one-half the number of cyclones. The size of the primary and secondary cyclones is assumed to be the same in this analysis, because the syngas volume flow rate changes only slightly between the desulfurization system inlet and outlet. The direct cost model is:

$$DC_{CY} = 0.98 N_{T,CY} \left(P_{G,o} \right)^{0.28} \left(\frac{2 V_{syn,G,o}}{N_{O,CY}} \right)^{0.43} \quad (R^2 = 0.997; n = 6) \quad (71)$$

where,

$$260 \leq P_{G,o} \leq 630 \text{ psia, and}$$

$$3,900 \leq \left(\frac{2 V_{syn,G,o}}{N_{O,CY}} \right) \leq 6,000 \text{ acfm.}$$

The number of operating cyclones should be twice the number of operating gasifiers, and the number of total cyclones should be twice the number of total gasifiers to ensure that there are two stages of cyclones per gasification train. The regression models should not be

**Figure 27. Direct Cost of Lurgi Gasification Subsystems:
Coke Handling, Fines Agglomeration, and Ash Removal**

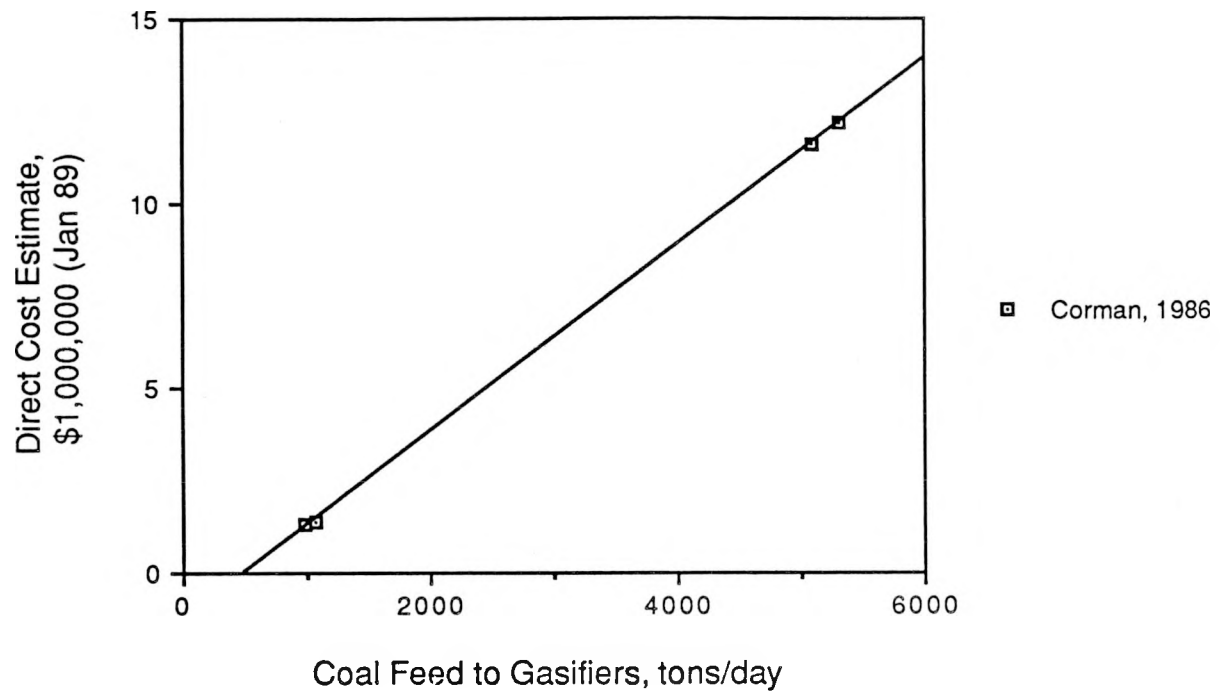
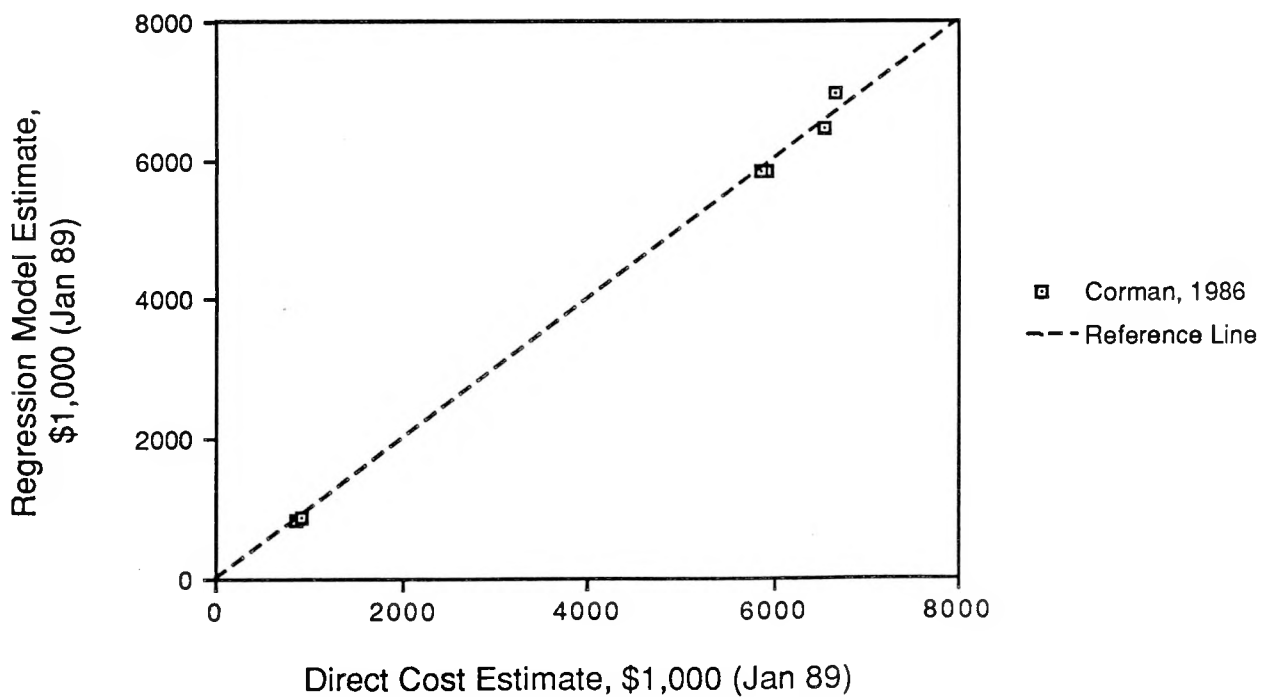


Figure 28. Predicted vs. Actual Direct Cost for High Temperature Cyclones



extrapolated beyond the ranges shown above. A comparison of the direct cost estimated from the multi-variate regression of Equation (71) and the direct costs reported in Corman (1986) are shown graphically in Figure 28 on the previous page.

The total direct cost for the Lurgi gasification section is given by the sum of its components:

$$DC_G = 7,700 \left(\frac{f_{p,i}}{7.14} \right) N_{T,G} + DC_{ss} + DC_{CY} \quad (72)$$

The term $f_{p,i}$ is the pressure correction factor presented in Equation (62). This factor is assumed to adjust the cost of the gasifiers for different pressure levels. The factor is divided by 7.14, which is the value of the factor at a pressure of 450 psia. This is the design pressure which is the basis for the cost estimate. Because the pressure range for the coal throughput estimate is 300 to 600 psia, the pressure levels for the gasifier should not exceed this range.

4.3 Sulfuric Acid Plant

The zinc ferrite desulfurization system generates an off-gas stream containing sulfur dioxide which must be processed to remove or recover the sulfur. The proposed method for recovering the sulfur is a sulfuric acid recovery plant (Corman, 1986; Cincotta, 1984; O'Hara et al, 1987). For a fixed bed zinc ferrite process, the sulfur dioxide concentration in the off-gas may be 1 to 4 percent on a volume basis. For a moving-bed zinc ferrite design, in which off-gas is recycled as a diluent for the exothermic reactions in the regenerator, the concentration of sulfur dioxide is higher, on the order of 10 percent. The ASPEN simulation is based on a moving bed zinc ferrite process.

A detailed performance and cost model of sulfuric acid plant was developed by Frey (1987). The model is sensitive to a number of factors affecting the cost of a sulfuric acid plant, including the volume flow rate of the off-gas stream, the amount of combustible gases in the off-gas, the amount of sulfur dioxide in the off-gas, and the inlet temperature of the off-gas. The sulfuric acid plant model is based on a detailed design and cost estimate developed for DOE by Monsanto under subcontract to Science Management Corporation (SMC, 1983). The plant is an interpass absorption contact design with off-gas conditioning.

The detailed sulfuric acid plant model was run for 100 cases with variation in the off-gas molar flow rate, sulfur dioxide concentration, moisture concentration, and off-gas temperature. A variant of Monte Carlo analysis employing Latin hypercube sampling was

used to select combinations of parameters for the 100 cases. From the resulting cost estimates, regression analysis was used to develop a simple expression for the cost of the sulfuric acid plant as a function of key performance parameters. The cost of the sulfuric acid plant is most sensitive to the total flow rate of the gas streams throughout the system. The total off-gas molar flow rate, concentration of sulfur dioxide, and temperature were identified as parameters with significant influence on cost. The simplified cost model is:

$$DC_{SA} = 52 + 18.2 M_{SA,i}^{0.67} f_{SO_2}^{0.12} T_{SA,i}^{0.12} \quad (R^2 = 0.956; n = 100) \quad (73)$$

where,

$$1,480 \leq M_{SA,i} \leq 70,000 \text{ lbmole/hr};$$

$$0.01 \leq f_{SO_2} \leq 0.13; \text{ and}$$

$$500 \leq T_{SA,i} \leq 1,500 \text{ } ^\circ\text{F}.$$

The model should not be extrapolated beyond the ranges of the predictive variables shown above. The standard error of the estimate is \$1.4 million. The plant does not include provision for off-gas particulate removal.

5.0 TOTAL CAPITAL COST MODEL

A framework for estimating the total capital requirement (TCR) for an IGCC power plant is presented in this section. The method is based on the EPRI Technical Assessment Guide (TAG) (1986), and is presented in the following section. A review of process contingency factors commonly assumed in the gasification systems cost studies is presented in Section 5.2. Section 5.3 discusses how uncertainty in the capital costs can be represented more rigorously using probabilistic methods.

5.1 Estimating the Total Capital Requirement

Table 9 lists the items included in the total capital requirement (TCR) for an IGCC system. The total direct cost (TDC) is the summation of the plant section direct costs (presented in previous chapters) and the general facilities cost. This is given by:

$$TDC = \left(\sum_{i=1}^n DC_i + GF \right) \left(\frac{I_{PCI}}{351.5} \right) \quad (74)$$

The total direct cost is referenced to January 1989 dollars, using the *Chemical Engineering Plant Cost Index*. The cost can be adjusted for other years using the appropriate value of the cost index in Equation (74).

Indirect construction costs include workers 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 the workers. From an analysis of the cost estimate for a KRW-IGCC plant located in Chicago, the indirect construction costs are approximately 25 percent of the total direct cost (Fluor, 1985). For the Lurgi-based system, the only indirect costs reported are for field supervision labor (Corman, 1986). Therefore, for this analysis it is assumed that the indirect construction cost is given by:

$$C_{ICC} = f_{ICC} TDC \quad (75)$$

where, as a nominal (default) value,

$$f_{ICC} = 0.25.$$

Table 9. Items Included in the Total Capital Requirement

Total Plant Cost (TPC)

Summation of Direct Costs for All Process Areas
 General Facilities
 Sales Tax
 Environmental Permits
 Indirect Construction Costs
 Engineering and Home Office Overhead and Fees
 Process Contingency
 Project Contingency

Total Plant Investment (TPI)

TPC
 Allowance for Funds Used During Construction (AFUDC)

Total Capital Requirement (TCR)

TPI
 Prepaid Royalties
 Spare Parts Inventory
 Preproduction (or startup) Costs
 Inventory Capital (fuel storage, etc.)
 Initial Chemicals and Catalyst Charges
 Land

Basis: EPRI (1986)

Table 10. Process Contingency Factors Recommended by EPRI^a

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

^a Cost estimates using these contingency factors are intended to represent the cost of commercialized (e.g., fifth of a kind) process plants. Source: EPRI (1986)

The cost of sales tax is specific to the state where the power plant is to be constructed. A common assumption is the Illinois sales tax of six percent, applicable to all material costs. Material costs comprise typically 80 percent of the total direct cost and 10 percent of the indirect costs, based on data in Fluor (1985). Therefore, the sales tax can be estimated as:

$$C_{\text{tax}} = r_{\text{tax}} (0.8 \text{ TDC} + 0.1 C_{\text{ICC}}) \quad (76)$$

where,

$$r_{\text{tax}} = 0.06, \text{ as a default value.}$$

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; (5) fees or profit to the architect/engineer. EPRI recommends that a value of 7 to 15 percent of the total direct cost, indirect construction cost, and sales tax be used. For this analysis, a value of 15 percent is assumed. The engineering and home office cost is thus given by:

$$C_{\text{EHO}} = f_{\text{EHO}} (\text{TDC} + C_{\text{ICC}} + C_{\text{tax}}) \quad (77)$$

where,

$$f_{\text{EHO}} = 0.15 \text{ as a default value.}$$

An item not commonly quantified in capital cost estimates is the cost of obtaining environmental permits required for the power plant. An allowance for permitting costs is included here, with a default value of one million dollars. This represents the costs 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 assumed here is a rough, order-of-magnitude estimate only. The cost of environmental permits is thus given by:

$$C_{\text{EP}} = 1,000 \quad (78)$$

The total indirect cost (TIC) is the sum of the indirect construction costs, engineering and home office costs, sales tax, and environmental permitting costs:

$$TIC = C_{ICC} + C_{tax} + C_{EHO} + C_{EP} \quad (79)$$

A major cost item for advanced technology plants is the process contingency. As discussed in Section 1.7.1, the process contingency is used in deterministic cost estimates to quantify the expected increase in the capital cost of an advanced technology due to uncertainty in performance and cost for the specific design application. In the EPRI cost method, the process contingency is estimated based on separate consideration of contingencies for each process section. The contingency is expressed as a multiplier of the sum of the direct and indirect capital costs for each plant section. Recommended ranges of process contingency factors are shown in Table 10 on page 123. The process contingency decreases as the commercial experience with a process area increases. For example, in a fully commercialized process, which has been used in similar applications, the process contingency may be zero. For a new concept early in the development stage, the process contingency may be over 40 percent of the process area cost. Experience has shown that cost estimates for innovative technologies early in the development phase tend to be low by a factor of two or more compared to the cost of the first commercial-size demonstration plant (EPRI, 1986; Merrow, Phillips, and Myers, 1981). However, the cost for subsequent plants tends to decrease, which is known as the "learning curve" effect. 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:

$$C_{PC,i} = f_{PC,i} \left[DC_i + TIC \left(\frac{DC_i}{TDC} \right) \right] \quad (80)$$

where,

$f_{PC,i}$ = process contingency for plant section i.

Equation (80) includes a term which prorates the total indirect costs to each plant section based on the ratio of the plant section direct cost to the plant total direct cost. This approach was used in Fluor (1985). The total process contingency allowance for the plant is given by the sum of process contingencies for each plant section:

$$C_{PC} = \sum_{i=1}^n C_{PC,i} \quad (81)$$

The process contingencies for each plant section are discussed in Section 5.2.

Table 11. Project Contingency Factors Recommended by EPRI^a

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

^a Expressed as a percentage of the total of total direct, total indirect, and process contingency.

Source: EPRI (1986)

In contrast to the process contingency, the *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. EPRI defines four levels of cost estimates, based on the type of information used to develop the estimate. These are listed, with brief explanatory notes, in Table 11. The type of estimates developed in this work are best classified as "preliminary." 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. In the EPRI study of KRW IGCC systems (Fluor, 1985), a project contingency of 15 percent was assumed, which is at the low end of the recommended range for preliminary estimates. Based on the overlap between recommended project contingencies for preliminary and "detailed" cost estimates, it is assumed here that the project contingency has a typical value of 20 percent. This value is also consistent with the range of recommended values for preliminary estimates. Furthermore, an upward adjustment in the contingency factor is consistent with the findings of a number of studies by the Rand Corporation, which indicate that contingency factors typically are under-estimated for advanced technology process plants (e.g., Milanese, 1987). Reasons for under-estimates often include: a lack of a organized record comparing the cost estimates of past projects with actual costs (which would provide a basis for correcting future cost estimates); the mis-application of a standard contingency factor (e.g., 10 percent) regardless of the type of estimate; and deliberate under-estimation of costs to ensure further consideration of the project by management. The project contingency is given by:

$$C_{PJ} = f_{PJ} (TDC + TIC + C_{PC}) \quad (82)$$

where,

$f_{PJ} = 0.20$ as the default value.

The total plant cost (TPC) is the sum of the total direct cost, total indirect cost, process contingency, and project contingency:

$$TPC = TDC + TIC + C_{PC} + C_{PJ} \quad (83)$$

The total plant cost is an "instantaneous" estimate; i.e., it 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.

The total plant investment, as shown in Table 9, 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 the following:

$$TPI = AF \times TPC \quad (84)$$

where,

$$AF = \frac{Z^N - 1}{N(Z - 1)} \quad (85)$$

$$Z = \frac{(1+i)}{(1+e_a)} \quad (86)$$

and,

i = interest cost for spent funds,

e_a = annual escalation rate for plant equipment (e.g., inflation rate)

The total process capital 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.

As shown in Table 9, the total capital requirement includes the total plant investment plus several other items. Prepaid royalties are fees paid to the owners of proprietary process technology designs, and are typically estimated as a fraction of the total plant investment if specific data are not available. The spare parts inventory is also estimated as a fraction of the

total plant investment. 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, and two percent of the total plant investment. The operating costs and fuel costs are developed in Section 6.0. The total preproduction cost is given by the following:

$$PPC = PP_{FC} + PP_{OC} + PP_{Fuel} + 0.02 TPI \quad (87)$$

The preproduction costs for fixed operating costs, variable operating costs, and fuel are presented in Section 6.3 in Equations (149), (150), and (151), respectively.

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 make-up catalyst and chemicals (e.g., for a Claus sulfur plant or Selexol process) and water treatment chemicals. The fuel and consumable requirements are developed in Section 6.0 for each of the three IGCC power plant systems. The inventory capital is given by the following:

$$IC = \sum_{i=1}^n IC_i \quad (88)$$

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, and fuel oil. For a KRW-based system with cold gas cleanup, the inventory capital also includes Selexol solvent, Claus plant catalyst, and either SCOT or Beavon-Stretford catalyst and chemicals. For a KRW-based system with hot gas cleanup, the inventory capital also includes limestone and zinc ferrite sorbent. For a Lurgi-based system with hot gas cleanup, the inventory capital also includes limestone, coke, bentonite, and zinc ferrite sorbent. The annual operating requirements for these processes are discussed in Section 6. A detailed expression for the inventory capital is developed and presented in Section 6.3, Equation (152).

The initial catalyst and chemicals charge is distinct from the inventory capital, and includes the cost of catalyst or chemicals that are contained in process equipment. Examples of initial catalyst or chemicals are the required catalyst inventory for a Claus plant or zinc ferrite process or the initial charge of solvent for the Selexol process.

The initial catalysts and chemicals required for a KRW-based system with cold gas cleanup include fuel oil for gas turbine startup, Selexol solvent, Claus plant catalyst, and SCOT catalyst and chemicals. If a Beavon-Stretford unit is used in place of a SCOT unit, then Beavon catalyst and the Stretford chemical are required. Data from a number of cost studies have been analyzed to determine the initial requirement for catalyst and chemicals as a function of key performance parameters.

The initial requirement for Selexol solvent is expected to depend primarily on the mass flow of hydrogen sulfide, the primary sulfur species in raw syngas, and on the concentration of the hydrogen sulfide. A multivariate regression yielded the following result for the initial solvent requirement, expressed in pounds:

$$\text{CHEM}_{i,S} = -25,200 + 16.6 \left(\frac{M_{\text{HS},S,i}^{0.935}}{f_{\text{HS}}^{1.04}} \right) \quad (R^2 = 0.959; n = 12) \quad (89)$$

where,

$$50 \leq M_{\text{HS},S,i} \leq 900 \text{ lbmole/hr,}$$

$$0.004 \leq f_{\text{HS}} \leq 0.012.$$

The standard error of the estimate is 68,000 lb of solvent. The solvent requirement estimated from the regression model is compared to the reported solvent requirement in Figure 29.

The initial catalyst requirement for two-stage Claus plants was found to depend on the recovered sulfur mass flow rate. The initial catalyst requirement, in tons, is given by:

$$\text{CAT}_{i,C} = 5.03 \times 10^{-3} m_{s,C,o} \quad (R^2 = 0.988; n = 13) \quad (90)$$

where,

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

The standard error of the estimate is 4.1 tons. The regression model is shown graphically in Figure 30.

Because the SCOT process is proprietary, no detailed data are reported regarding the performance of this process. The recovered sulfur mass flow rate is the only performance parameter that is widely reported in the cost studies that is relevant to the SCOT process. Therefore, single-variate regression analyses based on this parameter were used to estimate

Figure 29. Initial Solvent Requirement for the Selexol Process

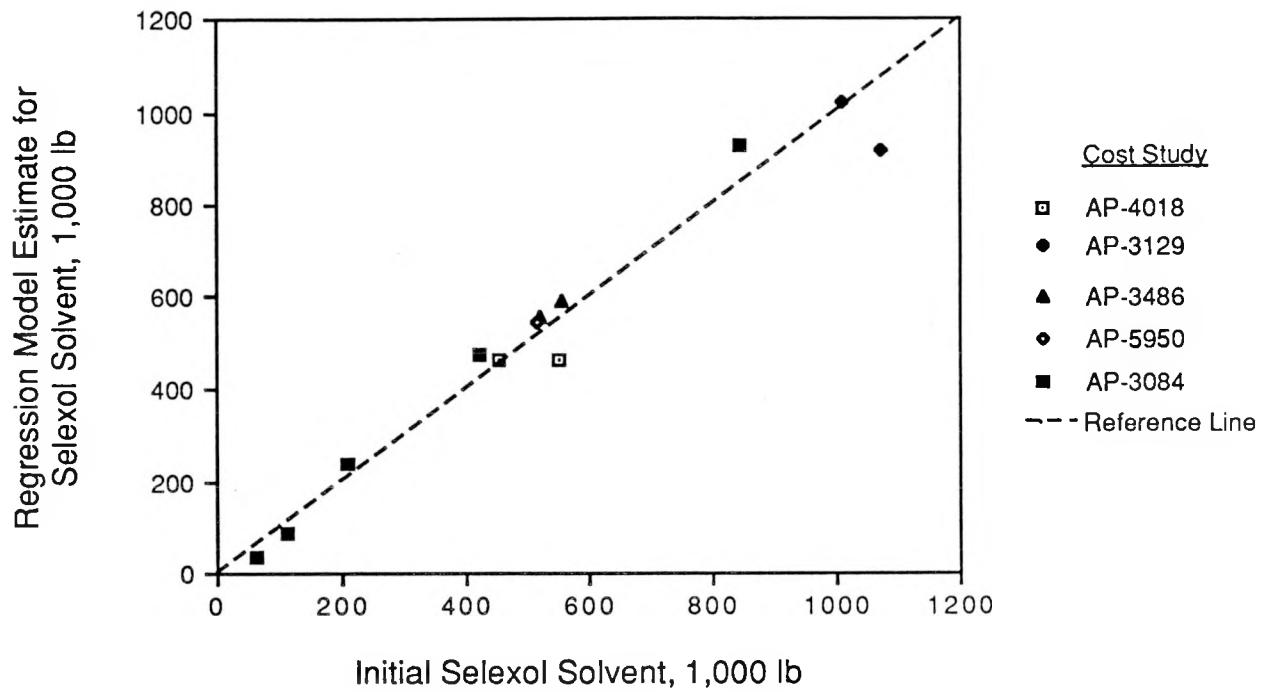
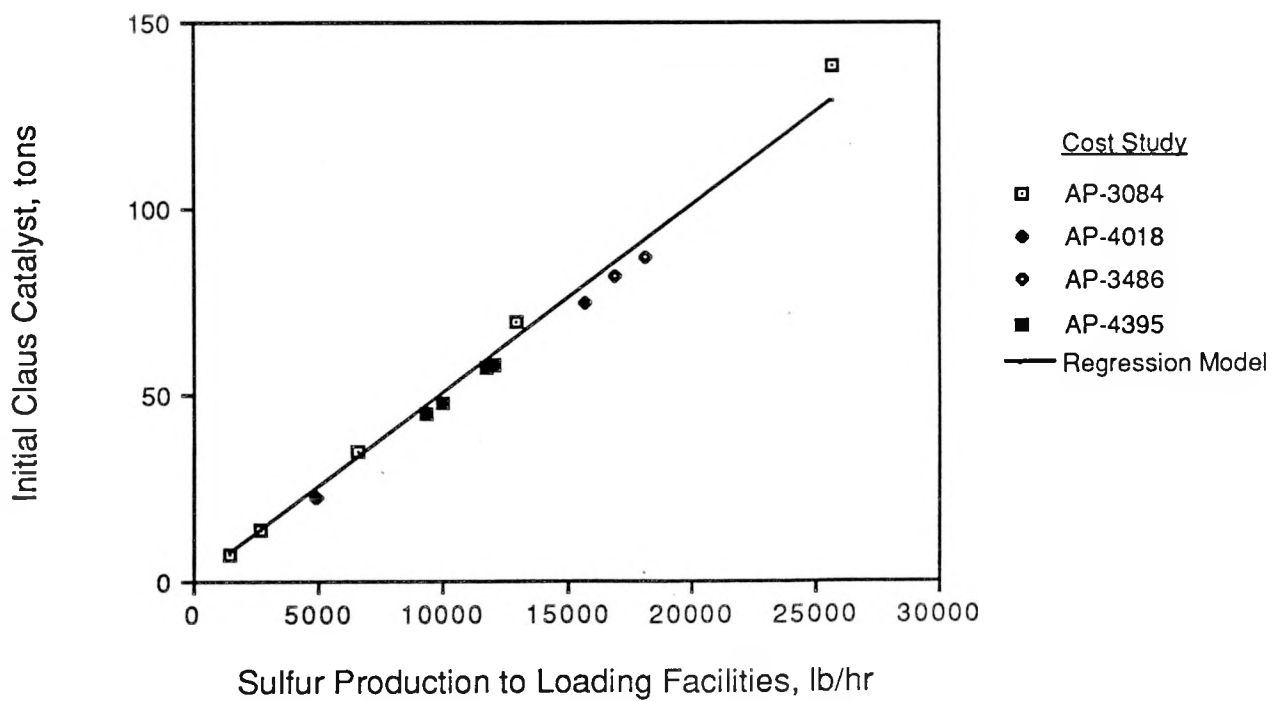


Figure 30. Initial Catalyst Requirement for Two-Stage Claus Plant



both the initial SCOT catalyst and SCOT chemical requirements. Data from three of the studies (Fluor, 1984; Fluor, 1985; Fluor, 1986) satisfied a simple straight-line approximation for catalyst or chemical requirement as a function of the recovered sulfur flow rate. Data from an earlier study (Fluor, 1983b) appear inconsistent with the three more recent studies, and were not included in the regression analysis. The resulting equations for the initial SCOT catalyst and chemical are:

$$CAT_{i,SC} = 19.3 + 0.161 m_{s,C,o} \quad (R^2 = 1.00; n = 7) \quad (91)$$

$$CHEM_{i,SC} = 1,270 + 10.7 m_{s,C,o} \quad (R^2 = 1.00; n = 7) \quad (92)$$

where,

$$5,000 \leq m_{s,C,o} \leq 18,000$$

The catalyst requirement is in units of cubic feet, and the initial chemical requirement is in units of pounds. The standard errors of the estimates are negligibly small. The regression equations are shown graphically in Figures 31 and 32 for the initial catalyst and chemical requirement, respectively. The regression equations reproduce almost exactly the reported requirements for most of the data shown, but over-estimate the requirements for the three data points not included in the regression model.

As an alternative to the SCOT process, some cost studies specify the Beavon-Stretford process for Claus plant tail gas treatment. The Beavon-Stretford process requires a catalyst for the Beavon unit and a special chemical for the Stretford unit. The initial catalyst and chemical requirements for the Beavon-Stretford process were estimated from the values reported in Fluor (1983a), which includes data for a range of plant sizes. From these data, a simple linear relationship of catalyst and chemical requirements as a function of the sulfur recovered in the Beavon-Stretford unit was identified. In the case of the Beavon catalyst, the mass requirement as a function of sulfur flow rate can be estimated. In the case of the Stretford chemicals, the mass requirement is not given. However, the cost of the initial Stretford chemicals as a function of the recovered sulfur flow rate was developed. The resulting regression models for the initial catalyst requirement, in cubic feet, and the initial chemical requirement, in dollars, are:

$$CAT_{i,BS} = -1.3 + 0.641 m_{s,BS,o} \quad (R^2 = 1.00; n = 5) \quad (93)$$

$$C_{i,BS,Chem} = 85.8 m_{s,BS,o} \quad (R^2 = 1.00; n = 5) \quad (94)$$

Figure 31. Initial Catalyst Requirement for the SCOT Process

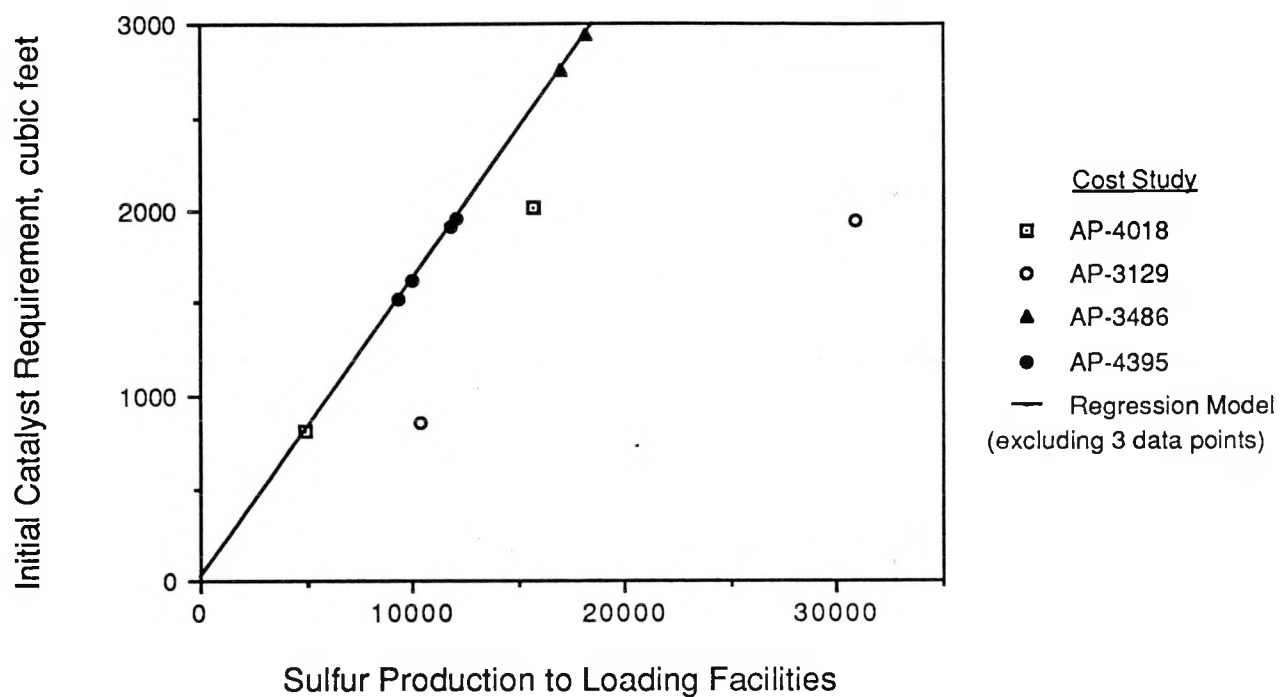


Figure 32. Initial SCOT Chemical Requirement

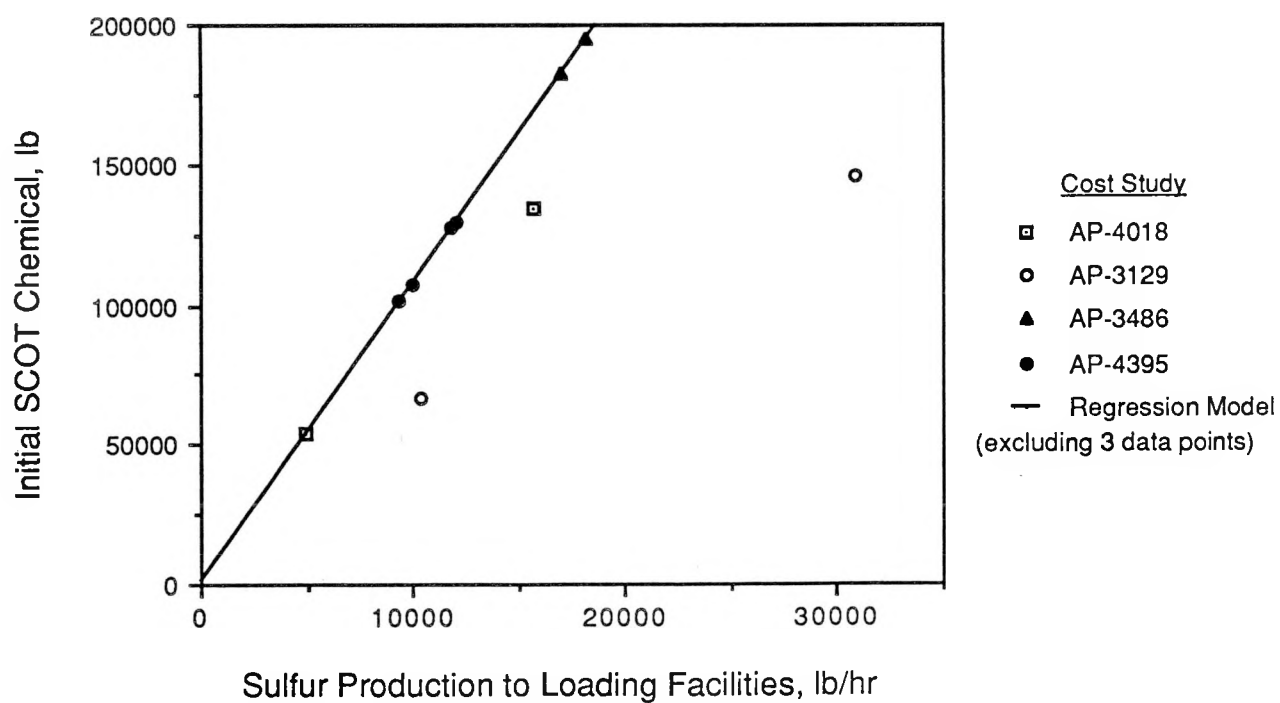
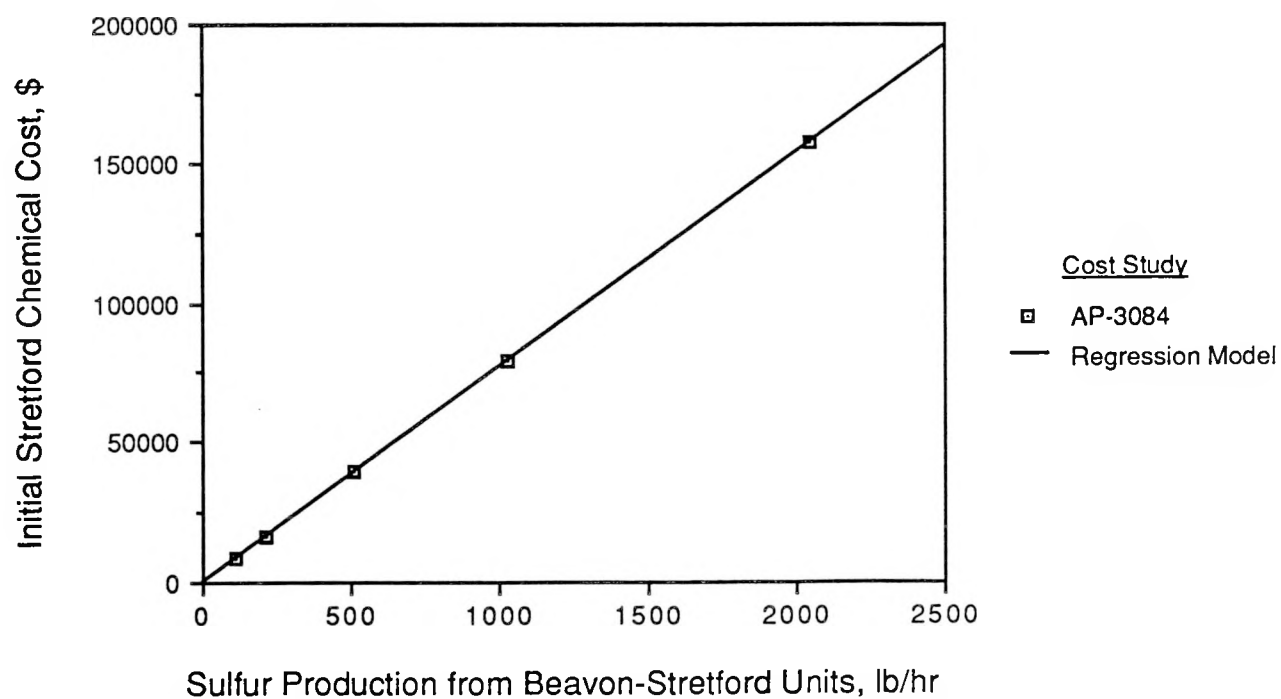


Figure 33. Initial Catalyst Requirement for the Beavon-Stretford Process



Figure 34. Initial Stretford Chemical Cost for the Beavon-Stretford Process



where,

$$100 \leq m_{s,BS,o} \leq 2,100 \text{ lb/hr.}$$

The standard error of the estimate for these models is negligible. The regression models for the initial catalyst and initial chemical requirements are shown graphically in Figures 33 and 34, respectively.

Fuel oil is required as a startup and auxiliary fuel for the 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:

$$FO_i = 80,000 \left(\frac{MW}{550} \right) \quad (95)$$

In addition to the initial catalysts and chemicals summarized above, chemicals are also required for treatment of the makeup cooling water. These chemicals include sulfuric acid (93 percent purity), corrosion inhibitor, and surfactant. However, the initial inventory of these chemicals is small, and in most cost estimates the cost of these is approximately 1,000 dollars (e.g., Fluor, 1985). Therefore, an allowance of 1,000 dollars for the initial charge of these chemicals is given by:

$$C_{CH,i,CW} = 1 \quad (96)$$

The total cost of initial catalysts and chemicals for a KRW-based system with cold gas cleanup is then given by:

$$C_{IC\&C} = UC_{FO} FO_i + C_{CH,i,CW} + \sum_j UC_j CHEM_{i,j} + \sum_k UC_k CAT_{i,k} \quad (97)$$

The subscript j refers to the name of the chemical and the subscript k refers to the name of the catalyst. The unit costs of chemicals and catalysts are discussed in Section 6.

For a KRW-based system with hot gas cleanup, the initial catalyst and chemical requirement includes zinc ferrite sorbent, fuel oil, and cooling water treatment chemicals. The required initial zinc ferrite sorbent charge for the on-line desulfurization units is discussed in Section 3.4 and is presented in Equation (36). The total amount of sorbent required is given by:

$$\text{CHEM}_{i,\text{ZF}} = S_c \left(\frac{N_{\text{T,ZF}}}{N_{\text{O,ZF}}} \right) \quad (98)$$

The fuel oil and cooling water treatment chemical requirements are estimated in the same manner as for the KRW-based system with hot gas cleanup.

For a Lurgi-based system with hot gas cleanup, coke and bentonite are required in addition to zinc ferrite sorbent, fuel oil, and cooling water treatment chemicals. Coke is required as a gasifier startup fuel. For a nominal 180 MW power plant, the initial coke requirement is 200 tons, based on Bechtel et al (1988). Therefore, the initial coke requirement may be estimated by:

$$\text{CHEM}_{i,\text{coke}} = 200 \left(\frac{\text{MW}}{180} \right) \quad (99)$$

Bentonite is used as a binder for fines agglomeration, based on the design specified in Corman (1986). However, the amount of binder required is not given. From Parsons (1985), the ratio of bentonite to coal fines is estimated to be 0.035 on a weight basis. The initial charge is equivalent to one week of consumption at full load. Therefore, the initial charge of bentonite in pounds is estimated by:

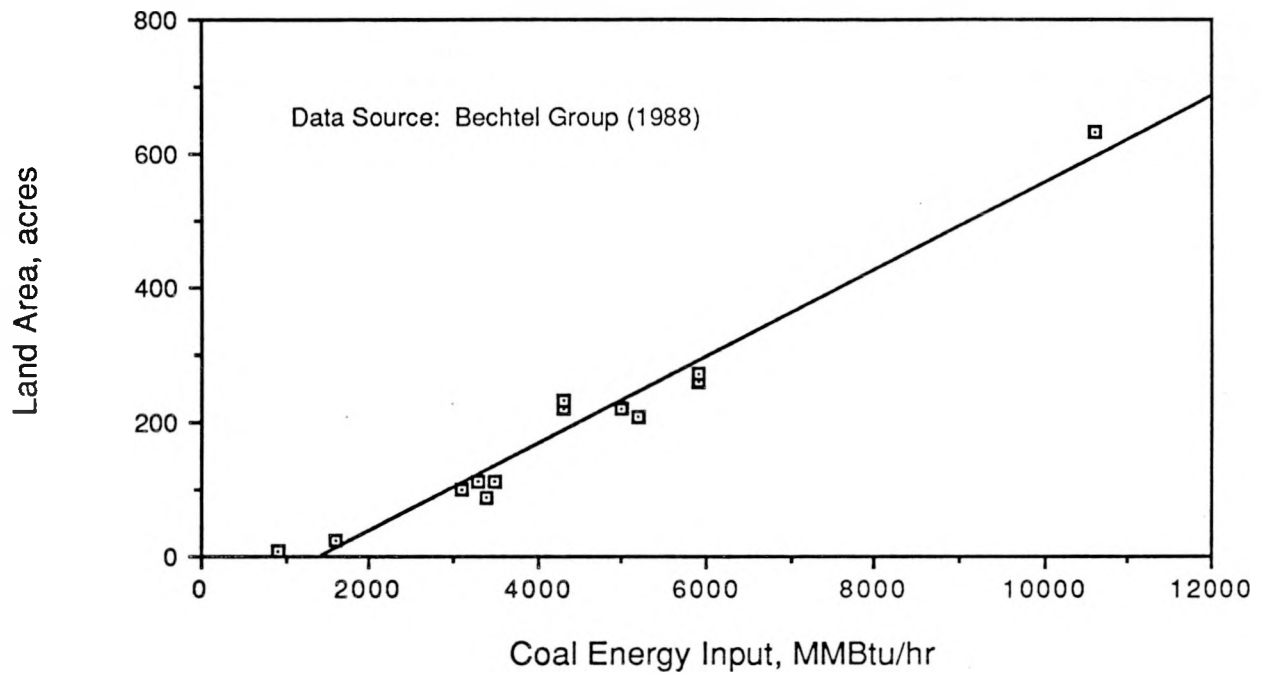
$$\text{CHEM}_{i,\text{Bent}} = 5.88 m_{\text{fines}} \quad (100)$$

The requirements for the other initial chemicals and catalysts are estimated using the equations developed for the KRW-based systems. The total cost of initial catalysts and chemicals are then estimated by multiplying the total requirements by the appropriate unit costs and summing, as in Equation (75).

The cost of land is estimated based on the total land required for an IGCC power plant and the cost per acre of land. Plant site requirements have been estimated by Bechtel Group (1988) based on analysis of data in 25 studies of gasification systems. The data are presented graphically in the Bechtel Group study. With the exception of two outlier points for phased systems, regression analysis was used to develop an analytic expression for the land requirement as a function of the plant coal energy input. The land requirement can be estimated from the following correlation:

$$A_L = -93 + 0.065 Q_{\text{coal}} \quad (R^2 = 0.962; n = 13) \quad (101)$$

Figure 35. Land Requirements for IGCC Plants



where,

$$900 \leq Q_{\text{coal}} \leq 11,000 \text{ MMBtu/hr}$$

The standard error of the estimate is 32 acres. The data and the regression equation are shown graphically in Figure 35.

Finally then, the total capital requirement (TCR) for an IGCC system is given by:

$$\text{TCR} = 1.01 \text{ TPI} + \text{PPC} + \text{IC} + C_{\text{IC\&C}} + A_L \text{UC}_L \quad (102)$$

The first term includes an allowance of 0.5 percent of the total process investment for prepaid royalties and 0.5 percent for spare parts inventory. The unit cost of land is assumed to be \$6,500 per acre, based on EPRI (1986).

It is common to express the total capital requirement on a normalized (\$/kW) basis. For IGCC plants, it is important to specify the ambient temperature for which the normalized cost is reported, because the gas turbine power output is a function of ambient temperature.

5.2 Process Contingencies

Reported process contingencies for the major plant sections of the three IGCC systems considered in this study are summarized in Table 12. These contingency factors are based on a review of several studies (Corman, 1986; Earley and Smelser, 1988a; Earley and Smelser, 1988b; Fluor, 1985; Smelser, 1986; Smith, Hanny, and Smelser, 1986; Smith and Smelser, 1987). For many commercially available systems, the process contingencies are zero. Contingency factors tend to be largest for those process areas which have not been built on a commercial scale, such as the gasification, sulfation, and zinc ferrite sections. As noted earlier, the process contingency is to represent the expected cost increases for a mature, "fifth-of-a-kind" plant. However, there is considerable variability in the contingencies used in different studies. The studies prepared for GRI tend to have the largest contingencies, while the studies of Lurgi-based systems with hot gas cleanup prepared by General Electric had the lowest overall process contingency factors of any of the studies examined.

The contingency factors developed in the GRI studies are based on a disaggregated consideration of several factors that pose a risk of increased process area costs. These factors are summarized in Table 13, which is taken from GRI (1983). The process development allowance, which is equivalent to the process contingency used in EPRI studies,

Table 12. Reported Process Contingency Factors for Selected IGCC Systems^a

Area Description	Area No.	KRW, Cold Gas Cleanup	KRW, Hot Gas Cleanup	Lurgi, Hot Gas Cleanup
Oxidant Feed	10	0	<5	<5
Coal Handling	20	0 - 12.5	12.5	0
Limestone Handling	25	--	0	--
Gasification	30	20 - 37.5	37.5	5
Sulfation	35	--	35 - 60	--
Low Temp. Gas Cooling	40	0	--	--
Selexol	50	0 - 12.5	--	--
Zinc Ferrite	50	--	40	5.8
Claus Plant	60	0 - 8.75	--	--
Sulfuric Acid Plant	60	--	--	5.9
SCOT	70	0 - 8.75	--	--
Beavon-Stretford	70	10	--	--
Boiler Feed Water System	80	0	0	0
Process Condensate Treatment	85	10 - 50	--	--
Combined Cycle System	90	2.7	2.7	2.7

References: Corman, 1986; Earley and Smelser, 1988a; Earley and Smelser, 1988b; Fluor, 1985; Smelser, 1986; Smith, Hanny, and Smelser, 1986; Smith and Smelser, 1987

^a These contingency factors are intended to be representative of commercial, fifth-of-a-kind plants.

is based on an average (expert judgment) "score" for four attributes of a technology. These attributes are state of development, availability of experimental data, untested assumptions used in the design, and operability and control difficulty. Recommended scores for each of these attributes are indicated in Table 13. Contingency factors estimated in this manner (i.e., based on a disaggregated consideration of several factors) are likely to be more accurate than contingency factors estimated as a single multiplier, as in the EPRI studies. The GRI estimates also tend to be significantly higher than the EPRI estimates. Although Lurgi gasifiers are commercially available, the contingency factors reported by Corman for the Lurgi-based system appear to be unrealistically low, particularly for the hot gas cleanup system, resulting in a downward bias of the capital cost estimates.

Table 13. Process Development Allowances Recommended by GRI^a

State of Development	Experimental Data	Assumptions	Operability and Control	Value (Percent of Direct Cost)
Conceptual	Phenomena identified	Critical assumptions for key processes or equipment	Not defined and not demonstrated	90 to 100
Laboratory	Laboratory only	Critical assumptions for supporting equipment or processes	Poorly defined and not demonstrated	70 to 90
PDU	PDU Level	Assumptions for supporting equipment or processes not demonstrated (and provided by others)	Preliminary definition, demonstration in development stage	50 to 70
Pilot plant	Pilot Plant Level	Assumptions for equipment or processes demonstrated in significantly different applications	Fair definition, demonstration may not be at design conditions	30 to 50
Demonstration plant	Demonstration plant level	Assumptions for equipment or processes demonstrated in similar but not identical applications	Good definition demonstrated in similar but not identical process	10 to 30
Commercial plant	Complete, demonstrated, commercial plant level	No assumptions: commercially demonstrated for the design case	Well defined, and commercially demonstrated for the design case	0 to 10

^a The process development allowance is a weighted sum of the values assigned for state of development, experimental data, assumptions, and operability and control. The PDA may be estimated based on a simple average of these values. The PDA is intended to account for expected cost increases for a commercial (mature) plant.

SOURCE: GRI (1983)

Because contingency factors for new processes tend to be significantly underestimated (Milanese, 1987), it is reasonable to use the largest contingency factors reported in the literature in first-pass cost analyses. However, the traditional approach to estimating contingency factors is inadequate for properly quantifying the risk of cost increases in new process technology. An alternative approach, using probabilistic analysis, is described below.

5.3 Representing Uncertainty in Capital Cost Estimates

The concept of a contingency factor is based on adjusting the capital cost estimate such that the probability of a cost over-run is below some acceptable threshold value (e.g., 50 percent). A low contingency factor for a new technology therefore implies a high probability of cost overrun. Therefore, contingency factors should be developed based on a probabilistic analysis of the capital cost of a process plant. A probabilistic analysis facilitates the disaggregated and quantitative consideration of uncertainties for specific parameters in an engineering model. Engineers are better able to make judgments about specific parameters than about an aggregate contingency factor. In fact, the GRI approach to estimating contingency costs involves disaggregation of the contingency factor into specific components, about which judgments are more easily made.

In a probabilistic modeling approach, any parameter in an engineering model may be treated as uncertain and quantified using probability distributions instead of point estimates. Uncertainties are particularly important for advanced concepts in an early phase of development, for which significant scale-up (or possibly redesign) is required for commercial application.

An important application of the cost models developed in this report will be the development of probabilistic estimates of IGCC system cost based on a disaggregated consideration of uncertainties in the performance and cost parameters of the engineering models. Although some of these uncertainties are indicated in the open literature, in most cases probability distributions for key performance and cost parameters must be estimated in consultation with process developers or other experts. The development of probabilistic performance and cost estimates for IGCC systems is addressed in a later phase of the work described here.

6.0 OPERATING COST MODEL

IGCC plant operating costs are also estimated using the method presented in the EPRI Technical Assessment Guide (EPRI, 1986). The items included in the operating costs are listed in Table 14. These can be divided into fixed and variable costs. Fixed costs are independent of the plant capacity factor or plant load. Variable costs, which include fuel, consumables, ash disposal, and byproduct credit, are directly proportional to the amount of energy produced by the plant. Fixed operating costs are discussed in Section 6.1. Variable operating costs are discussed in Section 6.2. Preproduction costs, which are required for the capital cost estimate, are summarized in Section 6.3. Because the ASPEN simulations include little or no detail on the power consumption of several plant sections, regression models for power consumption of selected plant sections are given in Section 6.4. In Section 6.5, a discussion of how uncertainty may be incorporated into the operating cost model is presented.

6.1 Fixed Operating Costs

Fixed operating costs include operating labor, maintenance labor and 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 common to assume that four shifts per day are required for plant operation, allowing two hours overlap for transition between shifts. Furthermore, an allowance for personnel on sick leave or vacation can be incorporated into the "shift factor." A shift factor of 4.75 is assumed as a default in this study, based on Bechtel (1988).

The number of operators required per shift for major plant sections are indicated for the three IGCC systems in Table 15. These estimates are based on data reported in several studies (Bechtel, 1983b; Bechtel, 1988; Bechtel et al, 1988; Corman, 1986; and Fluor, 1985). In some cases, the numbers are taken directly from the studies, while in other cases the numbers represent a judgment based on a review of several data sources. For the gasification and combined cycle areas, an assumption was made that the number of operating personnel are directly proportional to the number of operating trains for the process area.

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 applying the average labor rate as follows:

Table 14. Items Included in the Operating Cost Model

Fixed Operating Costs

Operating Labor
Maintenance Labor
Maintenance Materials
Administrative and Support Labor

Variable Operating Costs

Consumables
 Ex: Raw water, chemicals, catalyst, limestone
Ash Disposal
Fuel
Byproduct Credit

Basis: EPRI, 1986.

Table 15. Operating Labor Requirements for Selected IGCC Systems^a

Description	Area No.	Number of Operators per Shift		
		KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Oxidant Feed	10	5	--	--
Coal Handling	20	5	4	3
Limestone Handling	25	--	1	--
Gasification	30	2 N _{O,G}	2 N _{O,G}	2 N _{O,G}
Sulfation	35	--	1	--
Gas Cooling	40	--	--	--
Selexol	50	2	--	--
Zinc Ferrite	50	--	5	5
Claus Plant	60	2	--	--
Sulfuric Acid Plant	60	--	--	2
Tail Gas	70	1	--	--
Boiler Feed Water	80	1	1	1
Process Cond. Treat.	85	1	--	--
Combined Cycle	90	4 N _{O,GT} + 1	4 N _{O,GT} + 1	4 N _{O,GT} + 1
General Facilities	100	3	3	3

^a Based on a nominal 500 MW plant.

Sources: Bechtel (1983b); Bechtel (1988); Bechtel et al (1988); Corman (1986); Fluor (1985).

$$OC_L = ALR \left(\frac{2,080 \text{ hours}}{\text{year}} \right) SF \sum_i O_i \quad (103)$$

where,

ALR = Average labor rate (default is \$19.70/hr based on EPRI, 1986),

SF = Shift factor (default is 4.75, based on Bechtel, 1988), and

O_i = Number of operators in process area i , from Table 15.

The cost for maintenance material and labor for new technologies is typically estimated as a percentage of the installed capital cost for each process section. Recommended maintenance cost factors for the three IGCC systems are presented in Table 16. These maintenance cost factors are based on information reported in two cost studies (Fluor, 1985; Corman, 1986). The total maintenance cost for the plant is given by:

$$OC_M = \sum_i f_{M,i} (DC_i + IDC_i + C_{PC,i} + C_{PJ,i}) \quad (104)$$

where,

$$IDC_i = TIC \left(\frac{DC_i}{TDC} \right) \quad (105)$$

$$C_{PJ,i} = C_{PJ} \left(\frac{DC_i}{TDC} \right) \quad (106)$$

and,

$f_{M,i}$ = Maintenance cost factor for plant section i (from Table 16).

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). Thus:

$$OC_{MM} = 0.60 OC_M \quad (107)$$

$$OC_{ML} = 0.40 OC_M \quad (108)$$

The administrative and support labor cost is assumed to be 30 percent of the operating and maintenance labor cost:

$$OC_{AS} = 0.30 (OC_L + OC_{ML}) \quad (109)$$

Table 16. Annual Maintenance Cost Factors for Selected IGCC Systems

Description	Area No.	Maintenance Cost Factor ^a		
		KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Oxidant Feed	10	2.0	2.0	2.0
Coal Handling	20	3.0	3.0	3.0
Limestone Handling	25	--	3.0	--
Gasification	30	4.5	4.5	4.5
Coke, Fines, and Ash	31	--	--	3.0
Cyclones	32	--	--	1.5
Sulfation	35	--	4.0	--
Gas Cooling	40	3.0	--	--
Selexol	50	2.0	--	--
Zinc Ferrite	50	--	4.5	4.5
Claus Plant	60	2.0	--	--
Sulfuric Acid Plant	60	--	--	2.0
Tail Gas	70	2.0	--	--
Boiler Feed Water	80	1.5	1.5	1.5
Process Cond. Treat.	85	2.0	--	--
Combined Cycle	90	1.5	1.5	1.5
General Facilities	100	1.5	1.5	1.5

^a Annual maintenance cost as a percent of plant section direct, indirect, and contingency costs, as presented in Equation (81).

Sources: Corman (1986); Fluor (1985).

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

$$FOC = OC_L + OC_M + OC_{AS} \quad (110)$$

6.2 Variable Operating Cost

Variable operating cost includes fuel, consumables, ash disposal, and byproduct credits. Default unit costs of fuels, consumables, ash disposal, and byproduct credits are given in Table 17. In the following sections, the total requirements for fuel and consumable materials are developed, followed by a discussion of ash disposal and byproduct quantities. The material requirements are then used in conjunction with the unit costs in Table 17 to estimate the total variable cost.

6.2.1 Fuel Consumption

Fuel consumption for the IGCC power plants 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:

$$\hat{Q}_{coal} = 8,760 c_f m_{cf,CH,i} HV_{coal} \quad (111)$$

6.2.2 Feed Water Treating Consumables

Consumption of chemicals required for water treatment are a significant portion of the operating cost of an IGCC plant. The water streams that require treatment include boiler feed water and cooling water. In the boiler feed water system, raw water is treated and mixed with steam condensate. The combined stream is then chemically polished. Raw water used in the cooling water system must be treated to avoid fouling and corrosion in the cooling water system.

All of the IGCC systems require consumables for treatment of boiler feed water and steam cycle condensate. The chemicals required for raw boiler feed water demineralization include 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 primarily a combination of the raw water feed rate

Table 17. Unit Costs of Fuel, Consumables, Ash Disposal, and Byproducts

Item	Units	Cost	Basis	Reference	1/89 ^a
<u>Fuel</u>					
Illinois No. 6 Coal	\$/MMBtu	1.55	1/85	EPRI, 1986	1.61
Gulf Lignite	\$/MMBtu	1.50	1/85	EPRI, 1986	1.64
Wyoming Subbituminous	\$/MMBtu	1.85	1/85	EPRI, 1986	1.99
<u>Consumables</u>					
Sulfuric Acid (93%)	\$/ton	89.40	1/87	BGE, 1989	110
NaOH (50%)	\$/ton	175	1/87	BGE, 1989	220
Na ₂ HPO ₄	\$/lb	0.55	1/87	BGE, 1989	0.70
Hydrazine	\$/lb	2.50	1/87	BGE, 1989	3.20
Morpholine	\$/lb	1.02	1/87	BGE, 1989	1.30
Lime	\$/ton	65	1/85	EPRI, 1986	80
Soda Ash	\$/ton	125	1/87	BGE, 1989	160
Corrosion Inhibitor	\$/lb	1.55	1/83	Fluor, 1985	1.90
Surfactant	\$/lb	1.05	1/83	Fluor, 1985	1.25
Chlorine	\$/ton	200	1/87	BGE, 1989	250
Biocide	\$/lb	3.00	1/83	Fluor, 1985	3.60
Selexol Solvent	\$/lb	1.50	1/83	Fluor, 1985	1.80
Claus Catalyst	\$/ton	365	1/83	Fluor, 1985	440
Sulfuric Acid Catalyst	\$/liter	1.70	4/88	Higgins, 1988	1.90
SCOT Catalyst	\$/ft ³	192	1/83	Fluor, 1985	230
SCOT Chemicals	\$/lb	0.30	1/83	Fluor, 1985	0.36
Beavon-Stretford Catalyst	\$/ft ³	154	6/81	Fluor, 1983a	170
Beavon-Stretford Chemicals	\$/ (lb/hr sulfur) ^b	154	6/81	Fluor, 1983a	170
Zinc Ferrite Sorbent	\$/lb	3.00	6/88	Kasper, 1988	3.30
Plant Air Adsorbent	\$/lb	2.30	1/83	Fluor, 1985	2.80
Flare - LPG	\$/bbl	9.24	1/87	BGE, 1989	11.70
Wastewater Chemicals	\$/gpm ww	695	1/83	Fluor, 1985	840
Fuel Oil	\$/bbl	42	1/83	Fluor, 1985	42
Water	\$/1,000 gal	0.60	1/85	EPRI, 1986	0.73
Limestone	\$/ton	15	1/85	EPRI, 1986	18
Bentonite	\$/lb	0.024	6/84	Parsons, 1985	0.029
Coke	\$/lb	0.0175	6/84	Parsons, 1985	0.021
<u>Ash Disposal</u>					
Ash Disposal	\$/ton	8	1/85	EPRI, 1986	10
Sludge	\$/ton (dry)	9.25	1/85	EPRI, 1986	11.30
<u>Byproduct Credits</u>					
Sulfur	\$/ton	125	1/89	Frey, 1989	125
Sulfuric Acid	\$/ton	40	1/89	Frey, 1989	40

^a Prices except for coal adjusted to 1/89 basis using the industrial chemicals producer prices index in *Chemical Engineering* magazine. The values of this index are: 6/81 = 369.4; 1/83 = 339.9; 6/84 = 345.5; 1/85 = 337.7; 1/87 = 323.9; 4/88 = 367.6; 6/88 = 372.8; 1/89 = 411.25. Coal prices adjusted based on interpolation of data in EPRI (1986), Table B.4-2.

^b At 100% capacity factor.

and the flow rate of condensate from the steam turbine, with a minor contribution (e.g., less than 5 percent) from condensate in other process units.

The required quantity of each steam cycle water treatment chemical has been estimated as a function of the raw or polished water flow rate based on detailed data available in a number of cost studies (Fluor, 1983a; Fluor, 1983b; Fluor, 1984; Fluor, 1985; Fluor, 1986). For each chemical and application, the regression model is presented and summarized below. Each regression model is shown graphically in the indicated figures.

Sulfuric Acid for Raw Water Dimineralization

$$m_{sa,BF,i} = c_f(47.0 + 2.09 \times 10^{-3} m_{rw}) \quad (R^2 = 0.969; n = 14) \quad (112)$$

where,

$$24,200 \leq m_{rw} \leq 613,000 \text{ lb/hr.}$$

The sulfuric acid flow rate is expressed in units of tons per year. The standard error of the estimate is 70 tons/year. The regression model is shown graphically in Figure 36.

Sodium Hydroxide for Raw Water Dimineralization

$$m_{sh,BF,i} = c_f(9.5 + 4.20 \times 10^{-4} m_{rw}) \quad (R^2 = 0.969; n = 14) \quad (113)$$

where,

$$24,200 \leq m_{rw} \leq 613,000 \text{ lb/hr.}$$

The sodium hydroxide flow rate is expressed in tons per year. The standard error of the estimate is 15 tons/year. The regression model is shown graphically in Figure 37.

Sodium Phosphate for Raw Water Treating

$$m_{sp,BF,i} = c_f(115 + 3.61 \times 10^{-3} m_{rw}) \quad (R^2 = 0.962; n = 14) \quad (114)$$

where,

$$24,200 \leq m_{rw} \leq 613,000 \text{ lb/hr.}$$

The sodium phosphate flow rate is expressed in units of lb/yr. The standard error of the estimate is 140 lb/yr. The regression model is shown graphically in Figure 38.

Hydrazine for Raw Water Treating

$$m_{hy,BF,i} = c_f(529 + 0.0174 m_{rw}) \quad (R^2 = 0.898; n = 14) \quad (115)$$

where,

$$24,200 \leq m_{rw} \leq 613,000 \text{ lb/hr.}$$

Figure 36. Sulfuric Acid Requirement for Boiler Feed Water Dimineralization

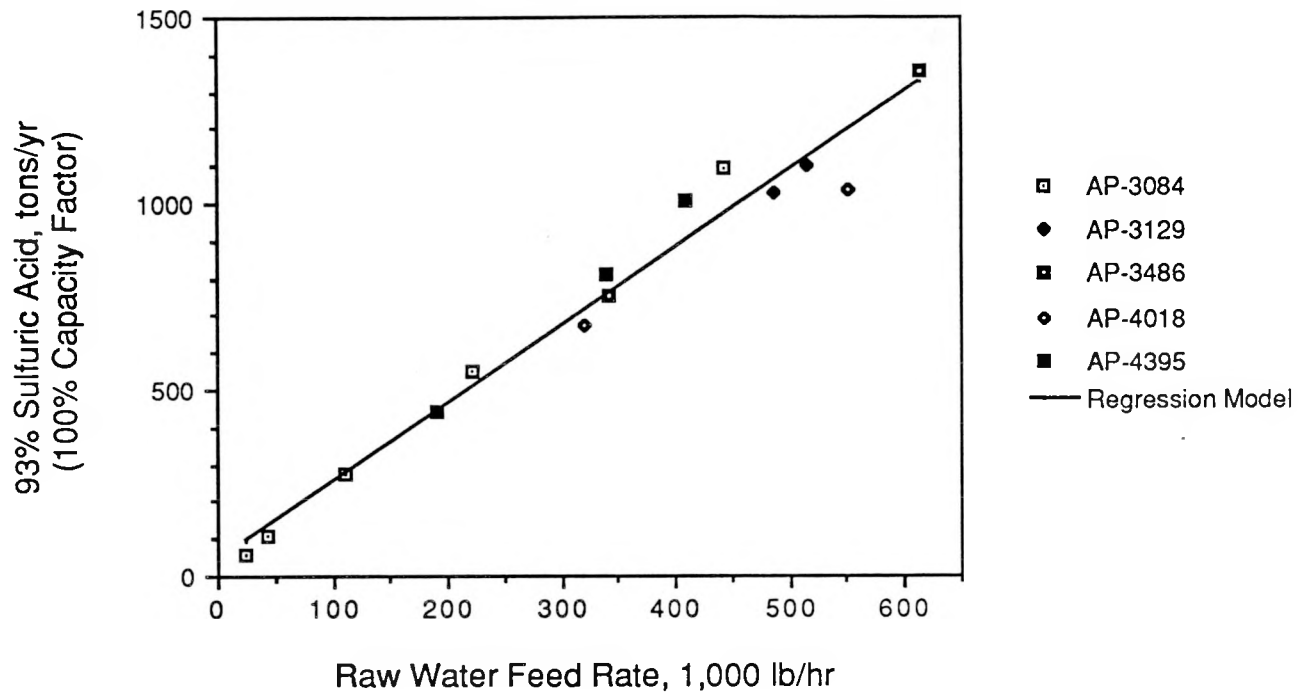


Figure 37. Sodium Hydroxide Requirement for Boiler Feed Water Dimineralization

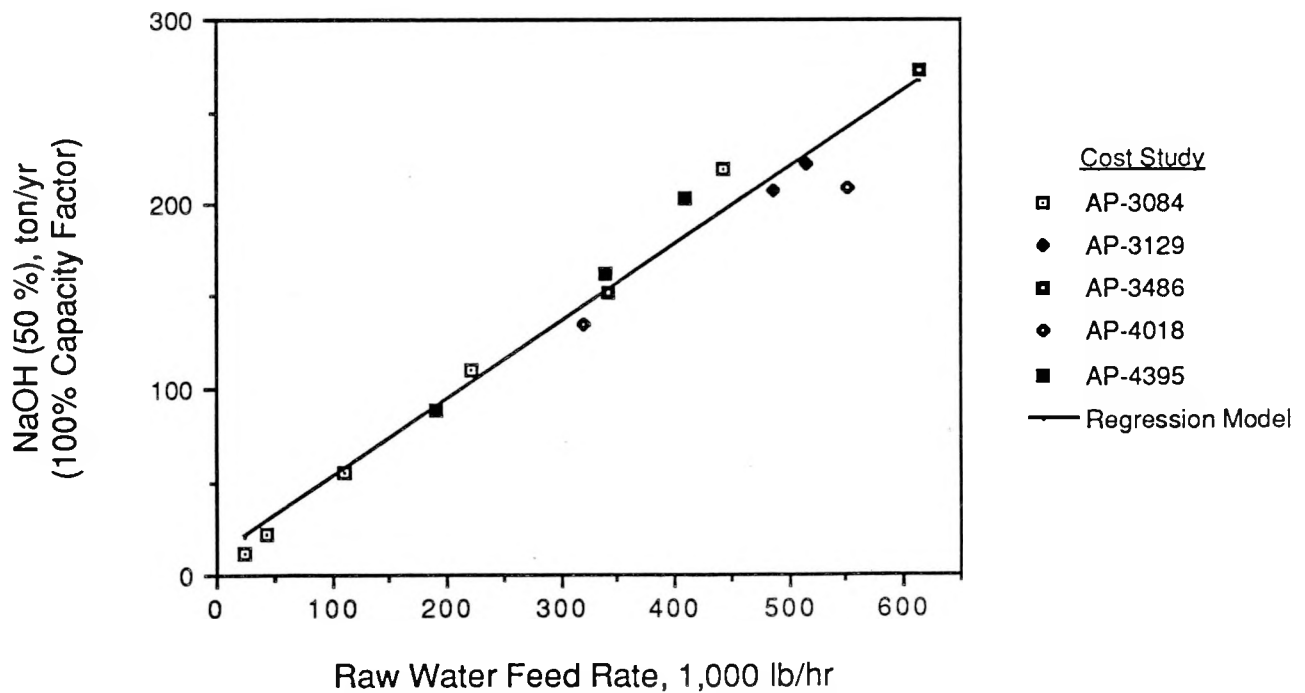


Figure 38. Sodium Phosphate Requirement for Boiler Feed Water Treating

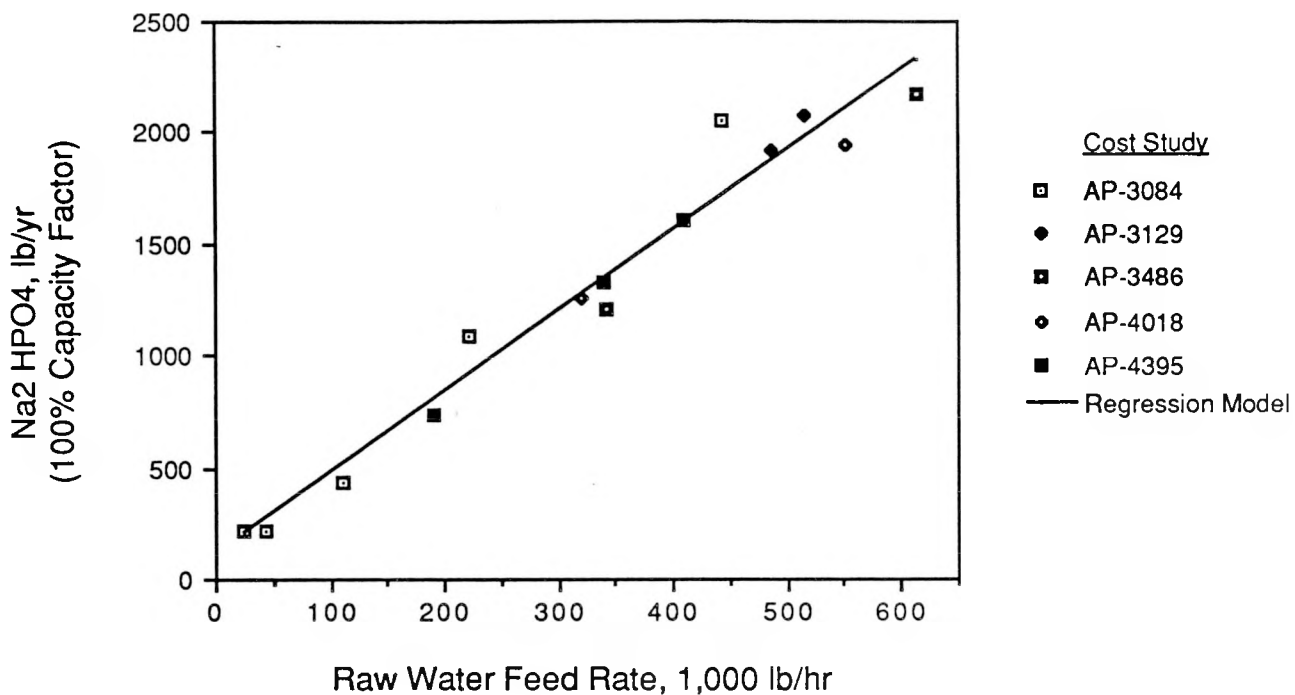
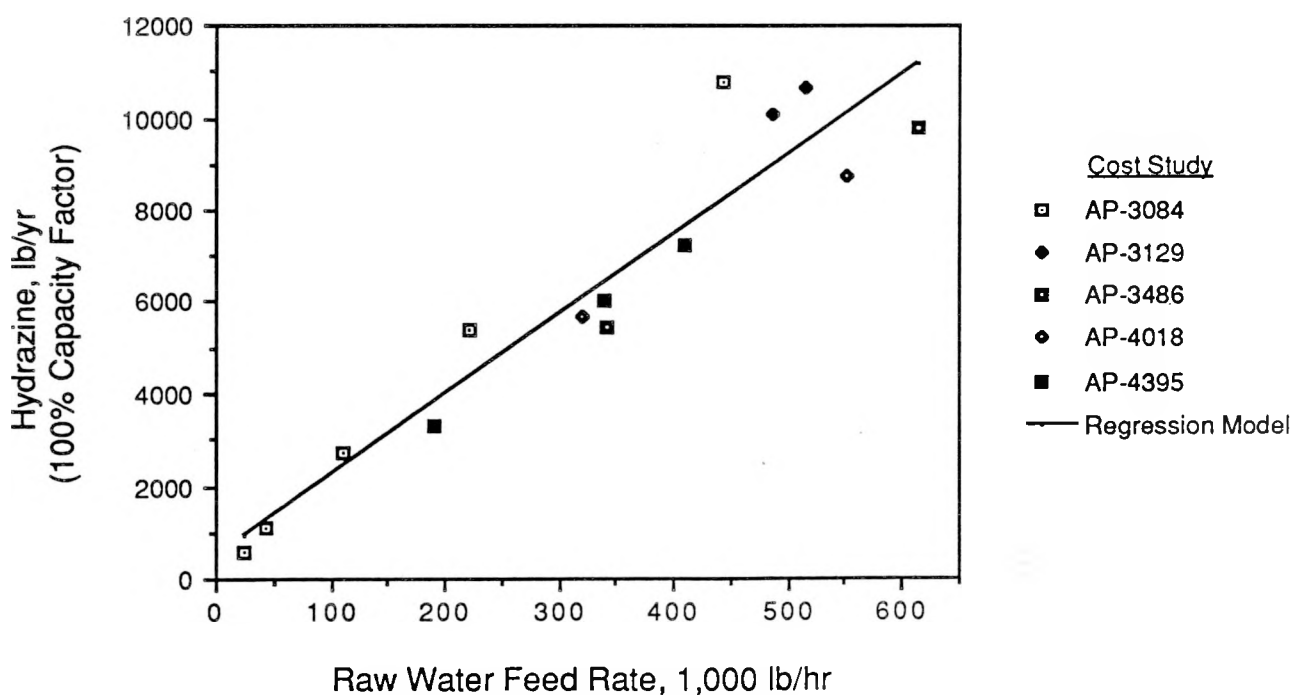


Figure 39. Hydrazine Requirement for Boiler Feed Water Treating



The hydrazine flow rate is expressed in units of lb/yr. The standard error of the estimate is 1,200 lb/yr. The regression model is shown graphically in Figure 39.

Morpholine for Raw Water Treating

$$m_{mo,BF,i} = c_f(420 + 0.0163 m_{rw}) \quad (R^2 = 0.965; n = 14) \quad (116)$$

where,

$$24,200 \leq m_{rw} \leq 613,000 \text{ lb/hr.}$$

The morpholine flow rate is expressed in units of lb/yr. The standard error of the estimate is 610 lb/yr. The regression model is shown graphically in Figure 40.

Sulfuric Acid for Condensate Polishing

$$m_{sa,BFP,i} = c_f(15 + 5.4 \times 10^{-5} m_{pw}) \quad (R^2 = 0.992; n = 7) \quad (117)$$

where,

$$1,200,000 \leq m_{pw} \leq 2,200,000 \text{ lb/hr.}$$

The sulfuric acid flow rate is expressed in units of tons per year. The standard error of the estimate is 2 tons/year. The regression model is shown graphically in Figure 41. As indicated in the figure, seven data points were excluded from the regression analysis. These data points are from the earliest of the five studies, and are inconsistent with the sulfuric acid requirement reported in the more recent studies. Furthermore, the data that were used in the regression model include the estimates developed specifically for a KRW-based system (Fluor, 1985).

Sodium Hydroxide for Condensate Polishing

$$m_{sh,BFP,i} = c_f(30 + 1.07 \times 10^{-4} m_{pw}) \quad (R^2 = 0.991; n = 7) \quad (118)$$

where,

$$1,200,000 \leq m_{pw} \leq 2,200,000 \text{ lb/hr.}$$

The sodium hydroxide flow rate is expressed in units of tons per year. The standard error of the estimate is 4 tons/year. The regression model is shown graphically in Figure 42. As indicated in the figure, seven data points were excluded from the regression analysis, for the same reasons noted above for the sulfuric acid requirement regression model.

Chemical treatment is also required for the cooling water system. From the same five studies used to development the boiler feed water chemical treatment requirements, the cooling water treatment requirements were estimated. The amount of each chemical required was found to be directly proportional to the makeup cooling water flow rate. The required chemicals include lime, soda ash, sulfuric acid, a corrosion inhibitor, a surfactant, chlorine,

Figure 40. Morpholine Requirement for Boiler Feed Water Treating

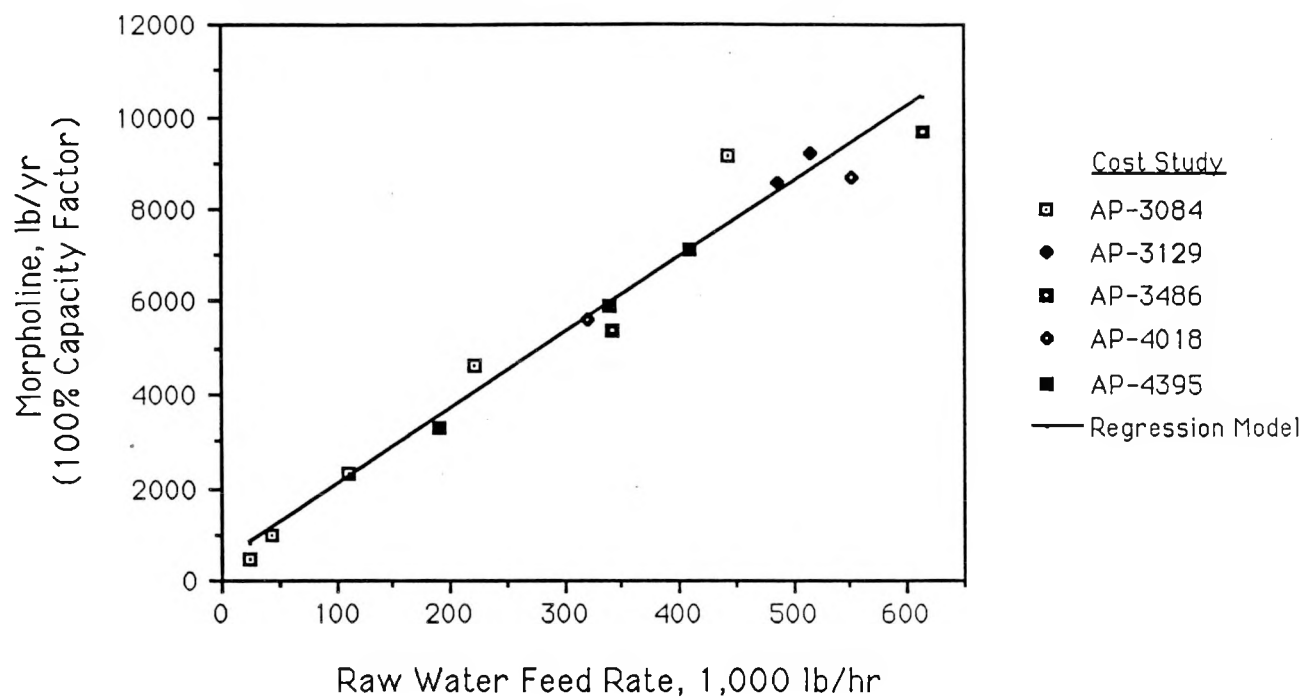


Figure 41. Sulfuric Acid Requirement for Polishing

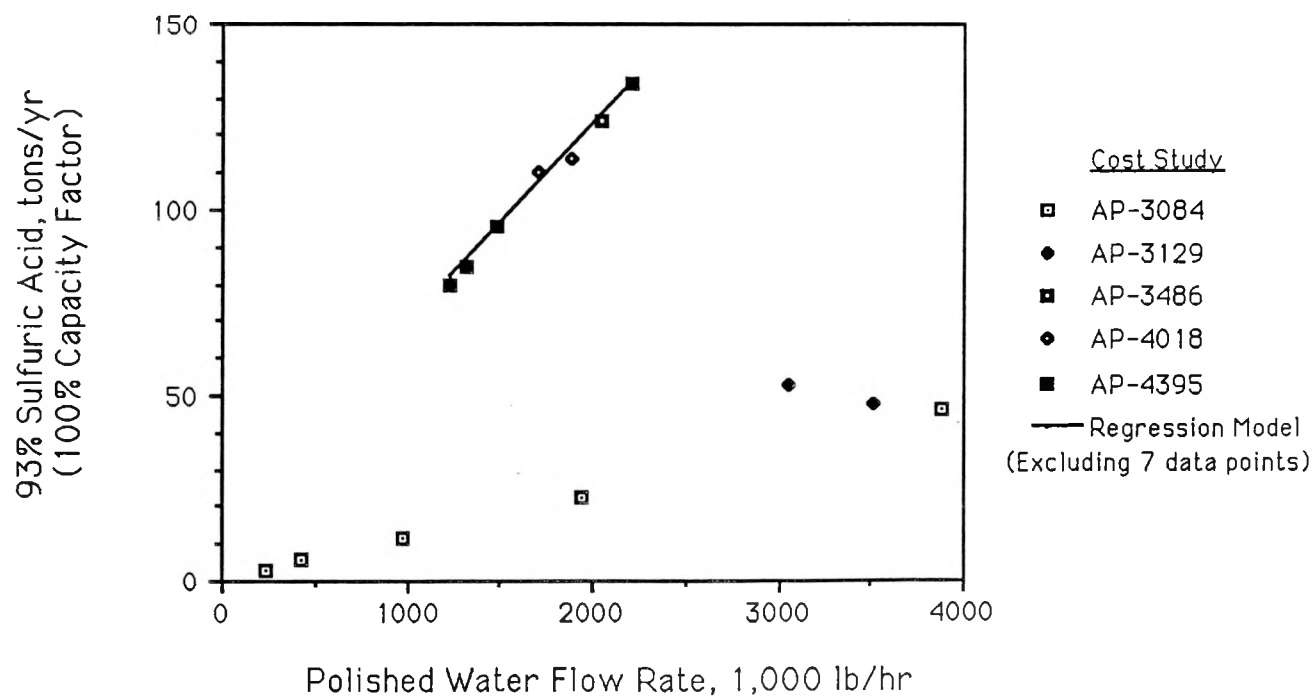


Figure 42. Sodium Hydroxide Requirement for Polishing

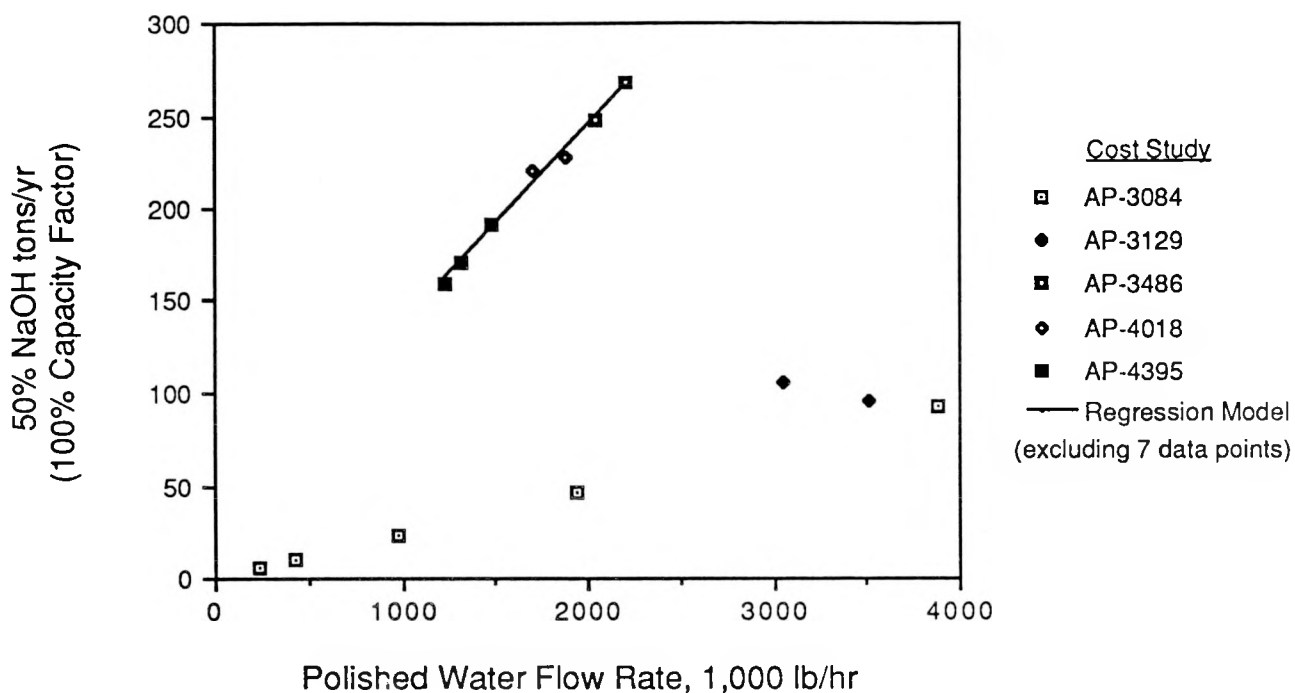
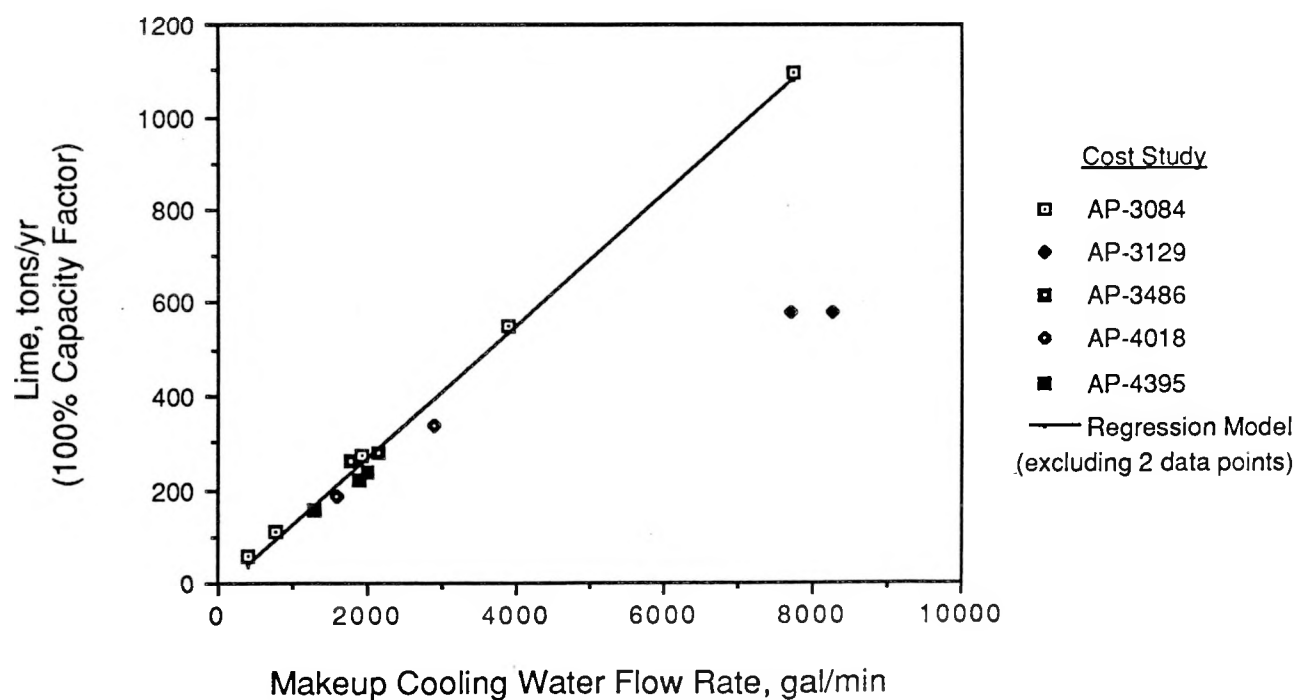


Figure 43. Lime Requirement for Makeup Cooling Water Treating



and a biocide. The regression models for the annual required flow rate of each of these treatment chemicals is presented below. Each regression model is represented graphically in the indicated figures.

Lime for Cooling Water Treating

$$m_{\text{lime,CW},i} = c_f(-26 + 0.143 m_{\text{cw},i}) \quad (R^2 = 0.991; n = 12) \quad (119)$$

where,

$$415 \leq m_{\text{cw},i} \leq 7,700 \text{ gal/min.}$$

The lime flow rate is expressed in units of tons per year. The standard error of the estimate is 27 tons/year. The regression model is shown graphically in Figure 43. As indicated in the figure, two data points, which are outliers, were excluded from the regression analysis.

Soda Ash for Cooling Water Treating

$$m_{\text{so,CW},i} = 0.154 c_f m_{\text{cw},i} \quad (R^2 = 0.989; n = 14) \quad (120)$$

where,

$$415 \leq m_{\text{cw},i} \leq 8,300 \text{ gal/min.}$$

The soda ash flow rate is expressed in units of tons per year. The standard error of the estimate is 47 tons/year. The regression model is shown graphically in Figure 44.

Sulfuric Acid for Cooling Water Treating

$$m_{\text{sa,CW},i} = c_f(6 + 0.147 m_{\text{cw},i}) \quad (R^2 = 0.984; n = 14) \quad (121)$$

where,

$$415 \leq m_{\text{cw},i} \leq 8,300 \text{ gal/min.}$$

The sulfuric acid flow rate is expressed in units of tons per year. The standard error of the estimate is 53 tons/year. The regression model is shown graphically in Figure 45.

Corrosion Inhibitor for Cooling Water Treating

$$m_{\text{ci,CW},i} = c_f(-7,280 + 28.8 m_{\text{cw},i}) \quad (R^2 = 0.964; n = 12) \quad (122)$$

where,

$$415 \leq m_{\text{cw},i} \leq 7,700 \text{ gal/min.}$$

The corrosion inhibitor flow rate is expressed in units of pounds per year. The standard error of the estimate is 11,100 lb/yr. The regression model is shown graphically in Figure 46. As indicated in the figure, two data points, which are outliers, were excluded from the regression analysis.

Figure 44. Soda Ash Requirement for Makeup Cooling Water Treating

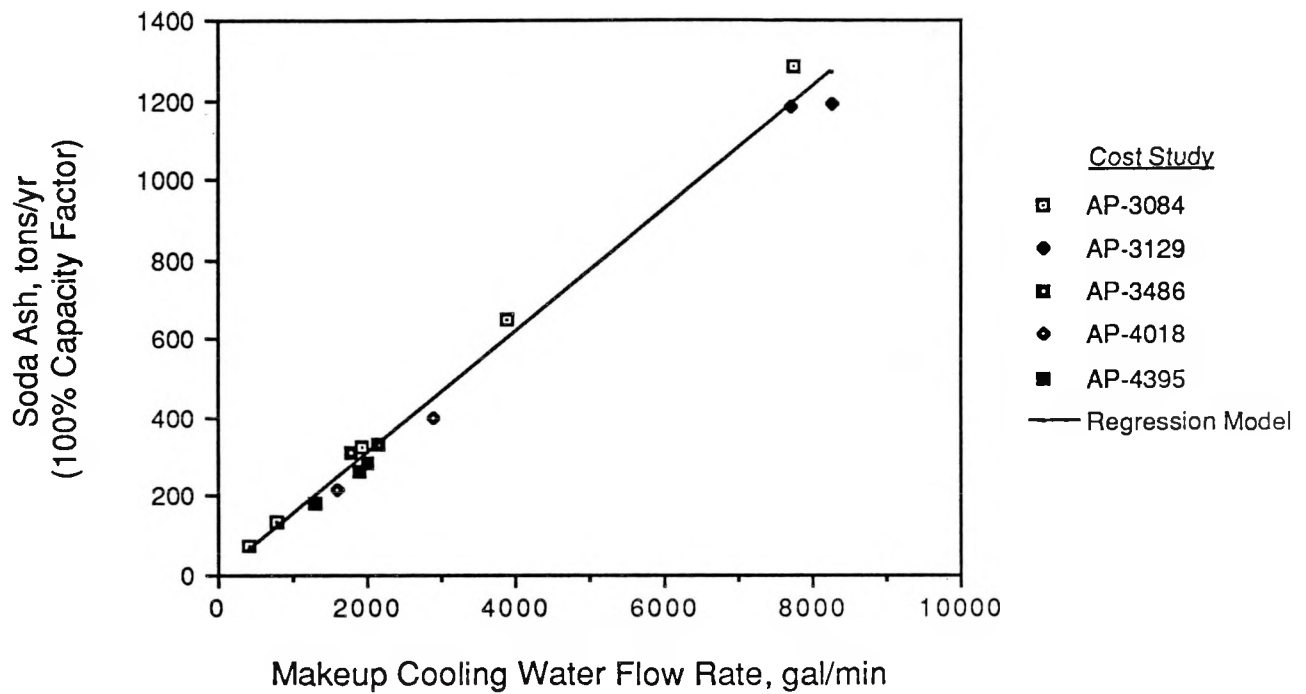


Figure 45. Sulfuric Acid Requirement for Cooling Water Treating

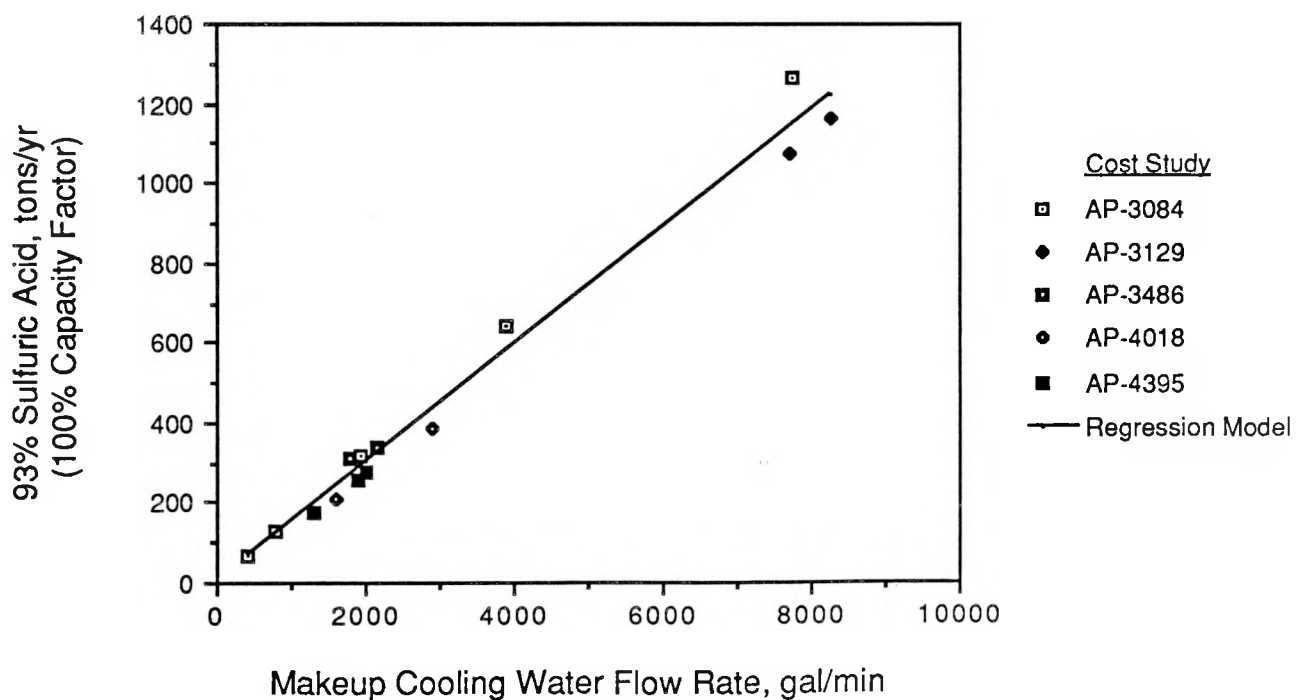


Figure 46. Corrosion Inhibitor Requirement for Cooling Water Treating

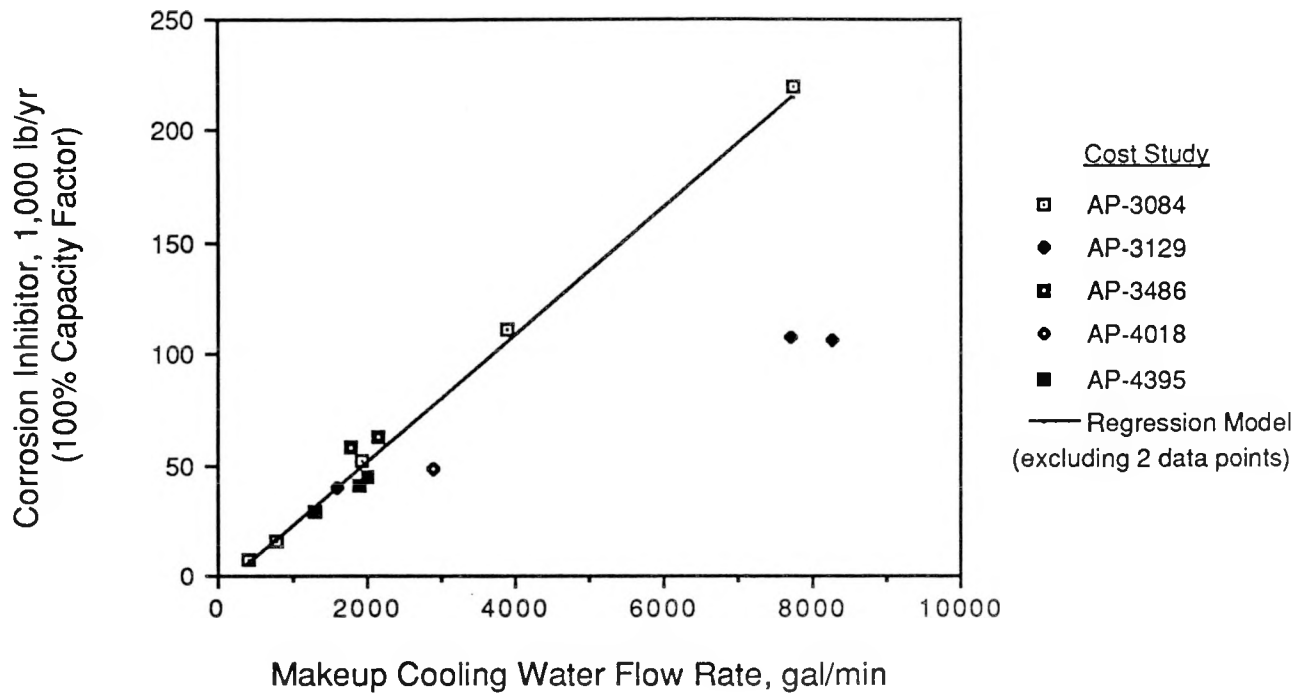
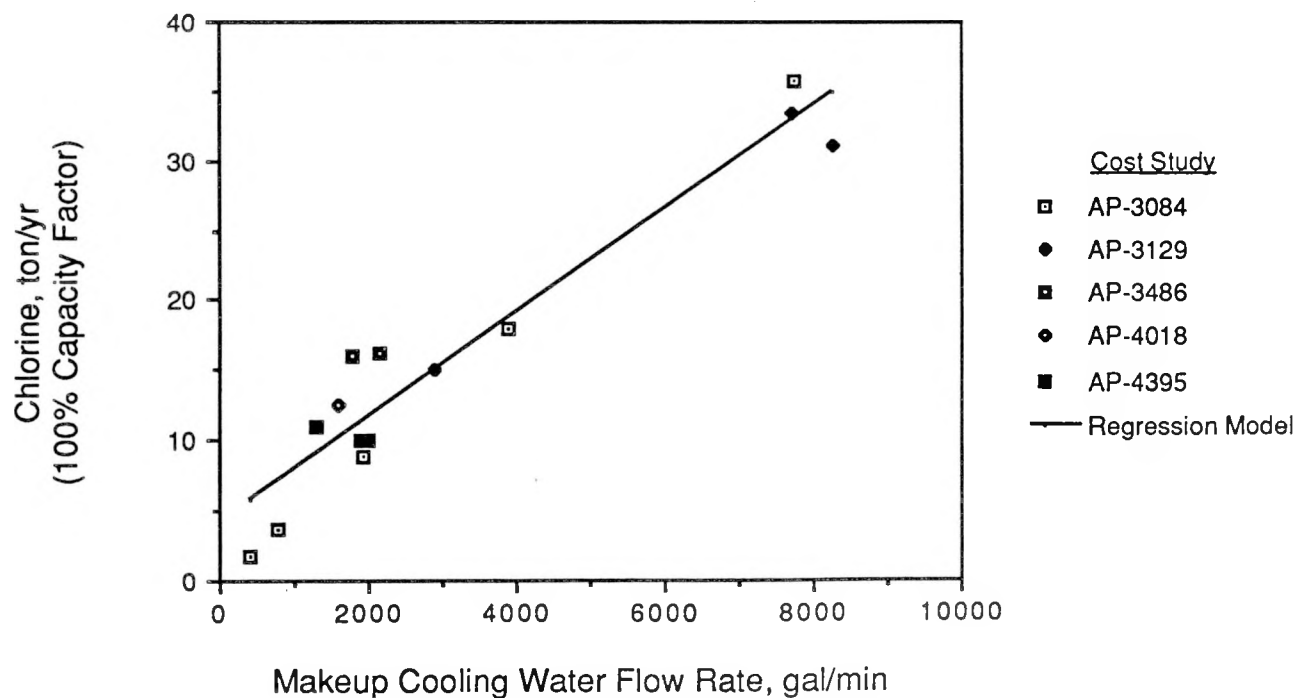


Figure 47. Chlorine Requirement for Cooling Water Treating



Surfactant for Cooling Water Treating

In all of the cost studies used to develop the cooling water treatment chemical requirements, the mass flow rate of surfactant is equal to the mass flow rate of corrosion inhibitor. Therefore, the surfactant requirement, in units of lb/yr, is given by:

$$m_{su,CW,i} = m_{ci,CW,i} \quad (123)$$

Chlorine for Cooling Water Treating

$$m_{Cl,CW,i} = c_f(4.1 + 3.74 \times 10^{-3} m_{cw,i}) \quad (R^2 = 0.920; n = 14) \quad (124)$$

where,

$$415 \leq m_{cw,i} \leq 8,300 \text{ gal/min.}$$

The chlorine flow rate is expressed in units of tons per year. The standard error of the estimate is 3.1 tons/year. The regression model is shown graphically in Figure 47.

Biocide for Cooling Water Treating

$$m_{bio,CW,i} = c_f(4,800 + 4.12 m_{cw,i}) \quad (R^2 = 0.922; n = 14) \quad (125)$$

where,

$$415 \leq m_{cw,i} \leq 8,300 \text{ gal/min.}$$

The biocide flow rate is expressed in units of pounds per year. The standard error of the estimate is 3,360 pounds/year. The regression model is shown graphically in Figure 48.

6.2.3 Sulfur Removal and Recovery

Makeup chemicals or catalysts are required for the sulfur removal and recovery systems in all IGCC designs. For cold gas cleanup systems, the makeup requirements include Selexol solvent, Claus plant catalyst, and either SCOT or Beavon-Stretford catalyst and chemicals. For the hot gas cleanup system with off-gas recycle, the only requirement is for makeup zinc ferrite sorbent. For a hot gas cleanup system with sulfuric acid recovery, makeup sulfuric acid catalyst is also required. The operating material requirements for these systems are summarized below.

Makeup Selexol Solvent

$$m_{ss,S,i} = c_f(-350 + 1.58 M_{syn,S,i}) \quad (R^2 = 0.989; n = 11) \quad (126)$$

where,

$$4,000 \leq M_{syn,S,i} \leq 74,500 \text{ lbmole/hr.}$$

Figure 48. Biocide Requirement for Cooling Water Treating

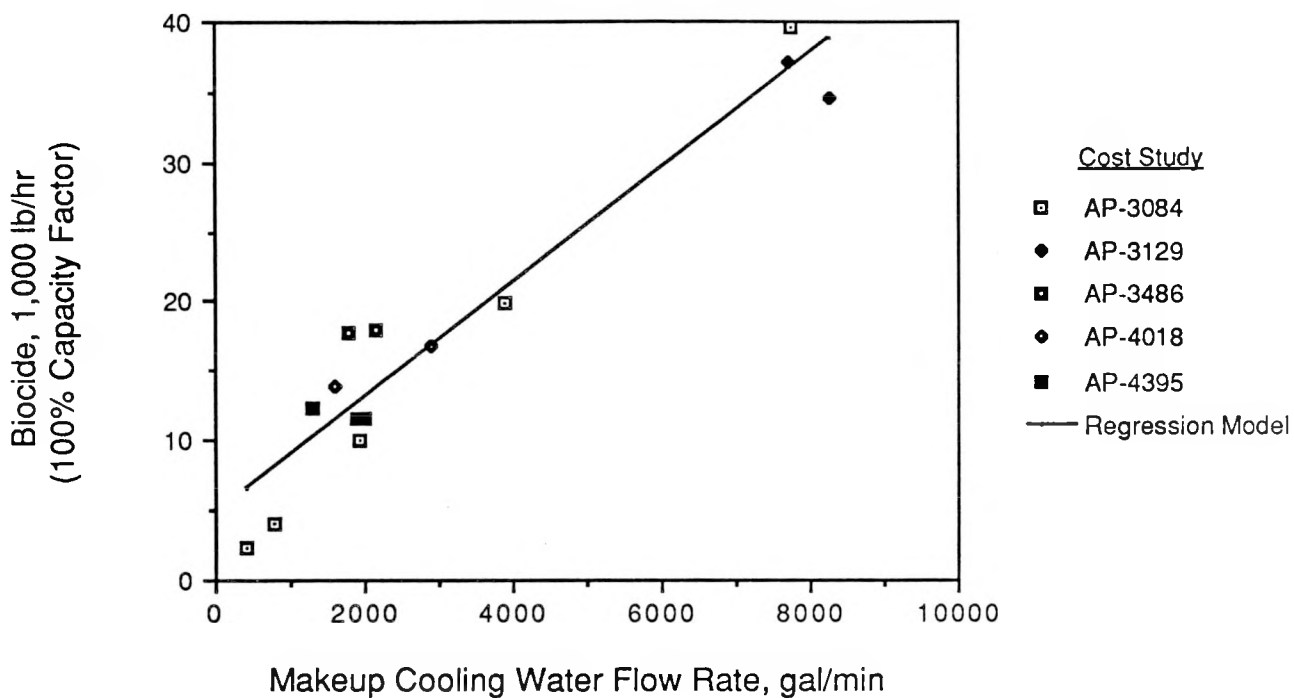
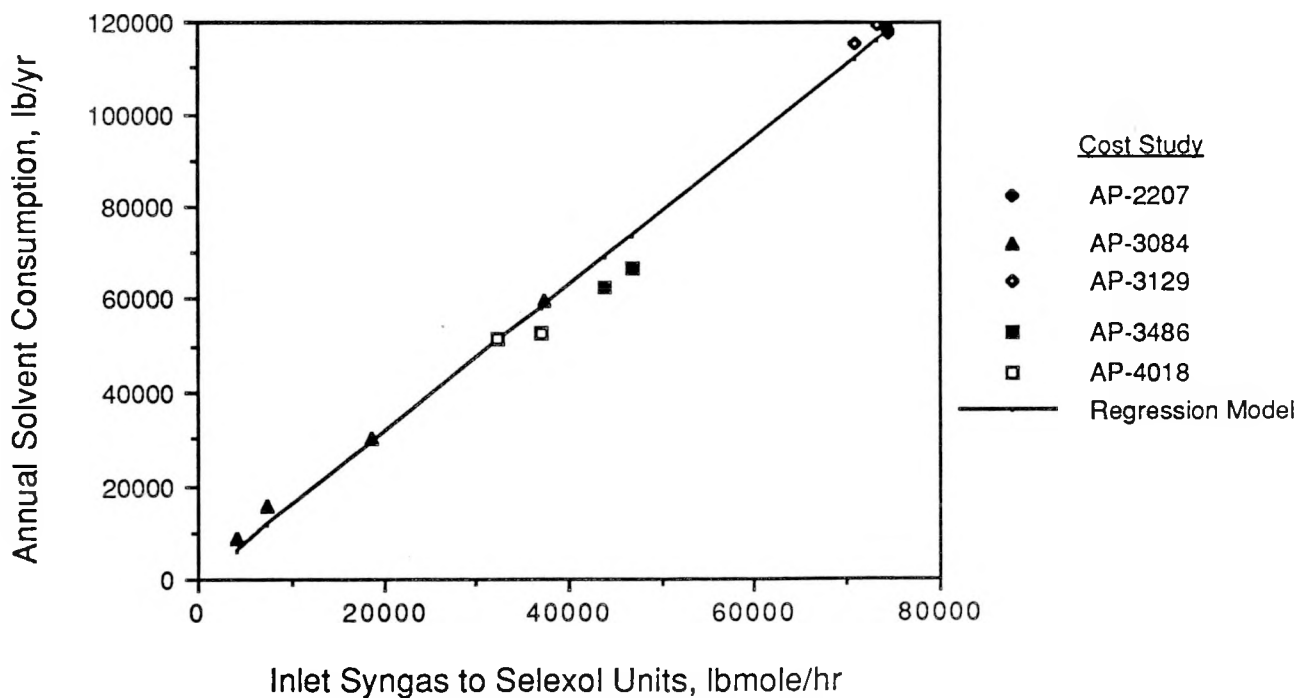


Figure 49. Annual Solvent Requirement for the Selexol Process



The makeup Selexol solvent flow rate is expressed in units of pounds per year. The standard error of the estimate is 4,400 pounds/year. The regression model is shown graphically in Figure 49.

Makeup Claus Plant Catalyst

$$m_{\text{cat,C,i}} = 9.61 \times 10^{-4} c_f m_{\text{s,C,o}} \quad (R^2 = 0.843; n = 13) \quad (127)$$

where,

$$1,000 \leq m_{\text{s,C,o}} \leq 26,000 \text{ lb/hr.}$$

The makeup Claus plant catalyst requirement is expressed in units of tons per year. The standard error of the estimate is 3 tons/year. The regression model is shown graphically in Figure 50.

Makeup SCOT Catalyst

$$m_{\text{cat,SC,i}} = 0.0822 c_f m_{\text{s,C,o}} \quad (R^2 = 0.993; n = 7) \quad (128)$$

where,

$$5,000 \leq m_{\text{s,C,o}} \leq 18,000 \text{ lb/hr.}$$

The makeup SCOT catalyst requirement is expressed in units of cubic feet per year. The standard error of the estimate is negligible. The regression model is shown graphically in Figure 51. As indicated in the figure, three outlier data points were excluded from the regression analysis. Two of these points are from the least recent study. The regression model over-estimates the outlier points, and therefore it is conservative for these outlier cases.

Makeup SCOT Chemicals

$$m_{\text{chem,SC,i}} = c_f (-270 + 1.10 m_{\text{s,C,o}}) \quad (R^2 = 0.883; n = 10) \quad (129)$$

where,

$$5,000 \leq m_{\text{s,C,o}} \leq 30,800 \text{ lb/hr}$$

The makeup SCOT chemical requirement is expressed in units of pounds per year. The standard error of the estimate is 3,000 lb/yr. The regression model is shown graphically in Figure 52.

Makeup Beavon-Stretford Catalyst

$$m_{\text{cat,BS,i}} = 0.0856 c_f m_{\text{s,BS,o}} \quad (R^2 = 1.00; n = 5) \quad (130)$$

where,

$$100 \leq m_{\text{s,BS,o}} \leq 2,000 \text{ lb/hr}$$

Figure 50. Annual Catalyst Requirement for Two-Stage Claus Plant

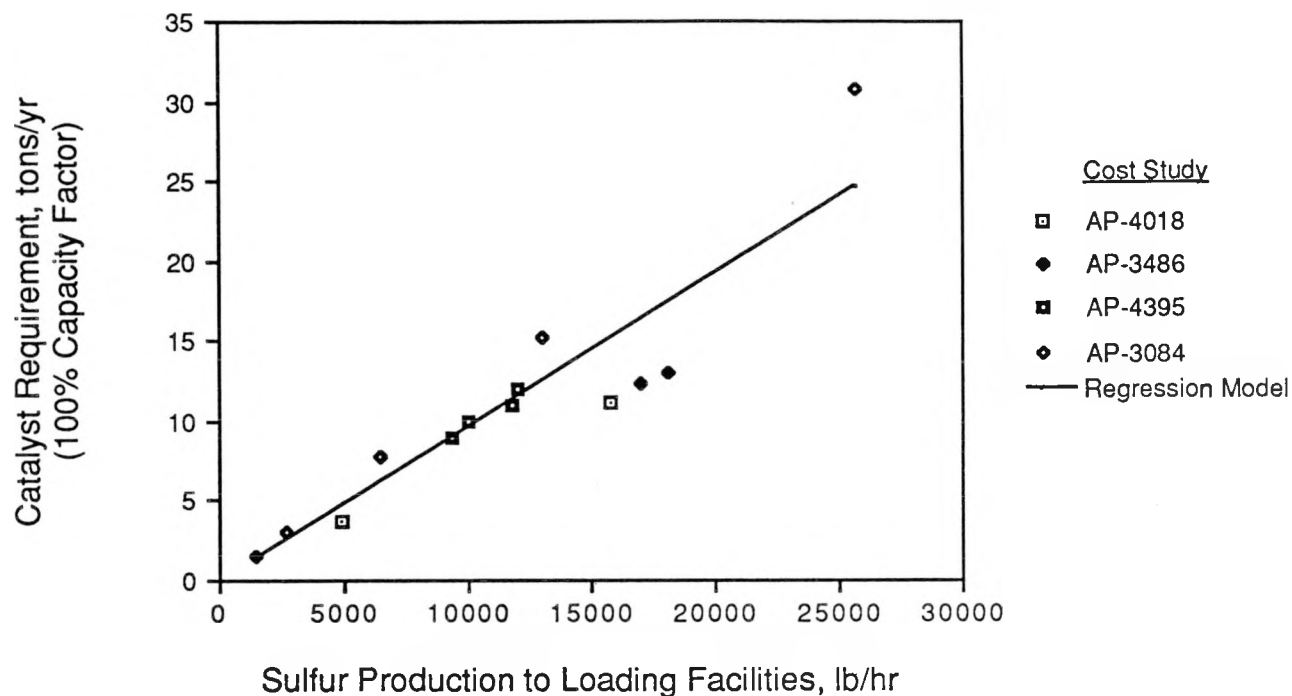


Figure 51. Annual Catalyst Requirement for the SCOT Process

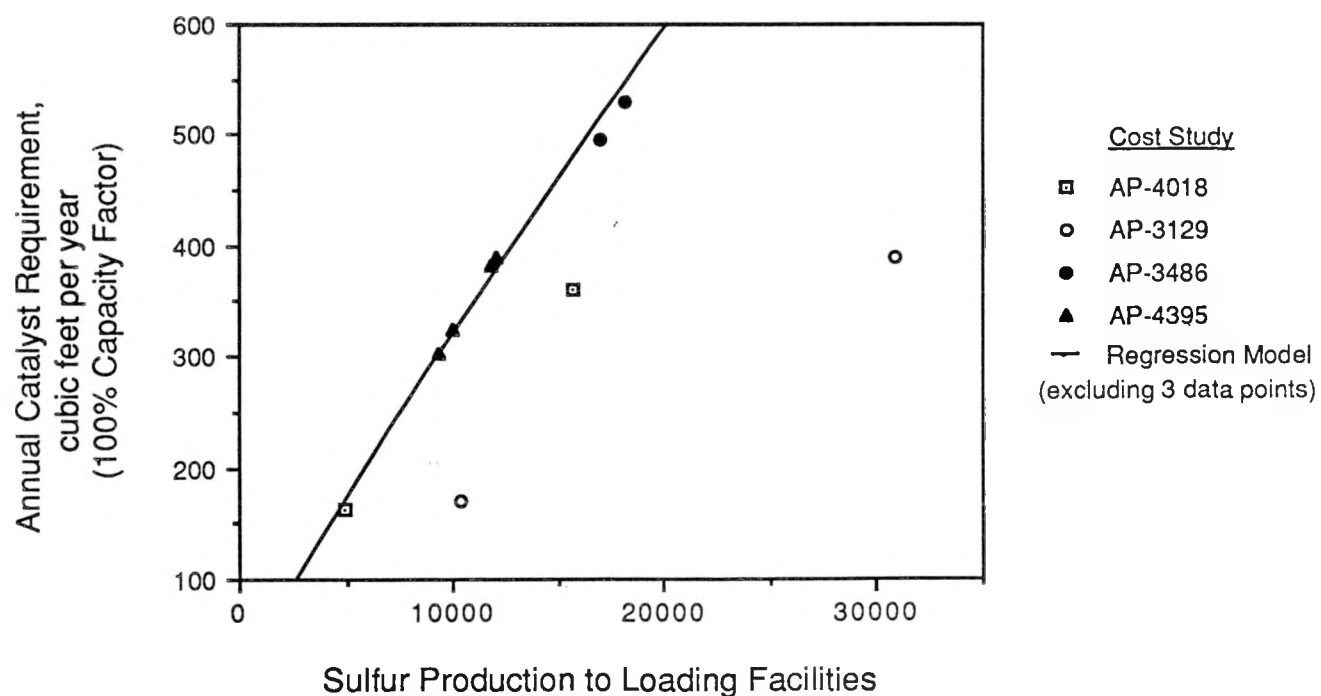


Figure 52. Annual SCOT Chemical Requirement

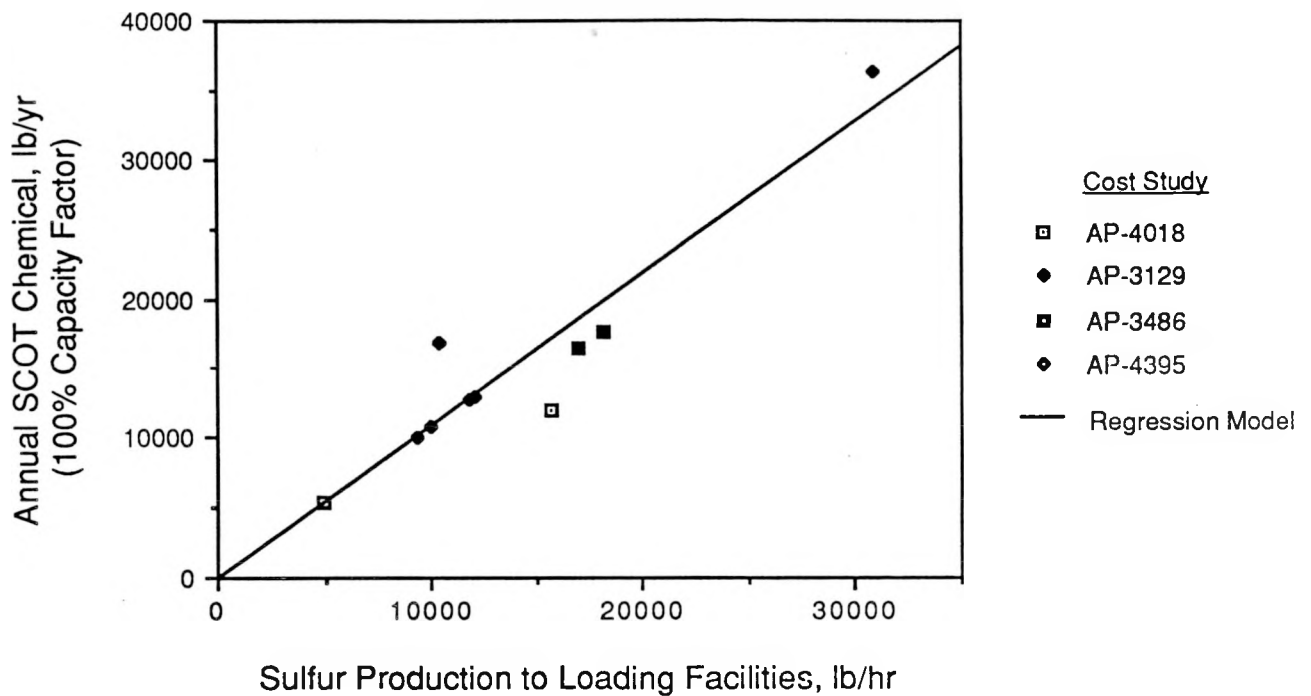
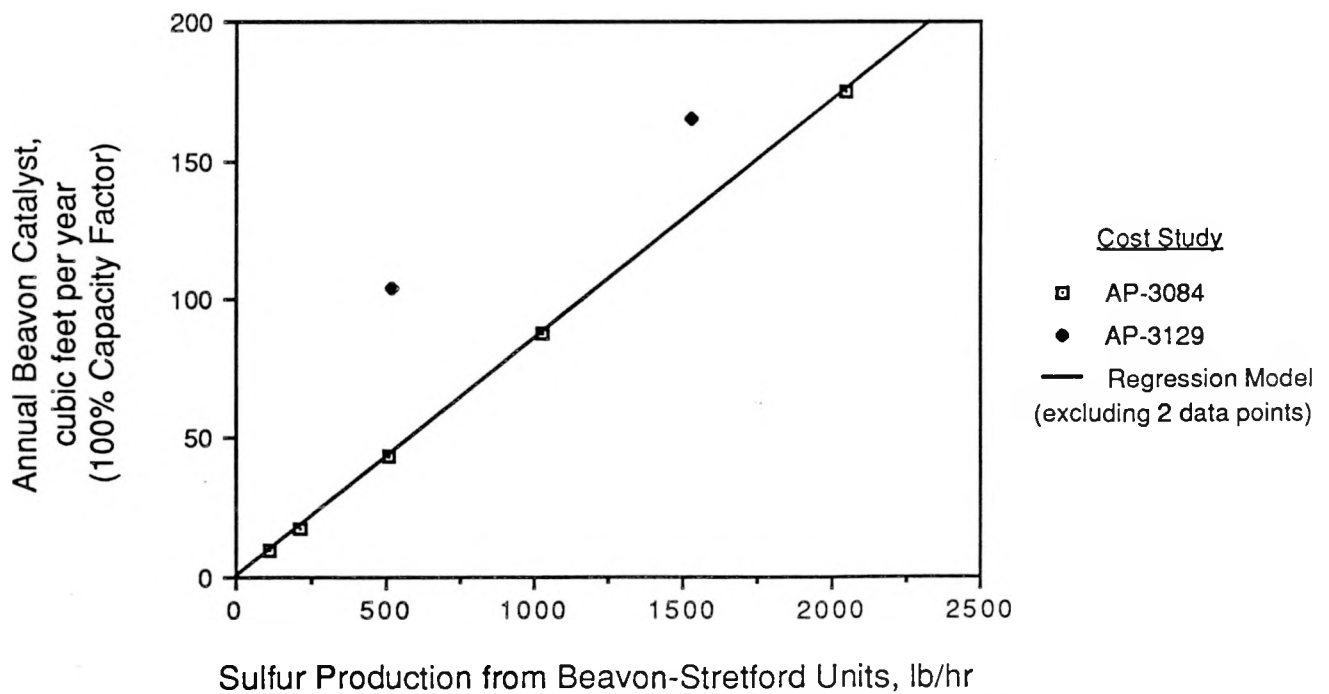


Figure 53. Annual Catalyst Requirement for the Beavon-Stretford Process



The makeup Beavon-Stretford catalyst requirement is expressed in units of cubic feet per year. The standard error of the estimate is negligible. The regression model is shown graphically in Figure 53. Two outlier data points were excluded from the analysis, as indicated in the figure. These points, both from the same study (Fluor, 1983b), appear inconsistent with the more extensive set of data from the other study (Fluor, 1983a).

Makeup Beavon-Stretford Chemicals

$$C_{\text{chem,BS},i} = 170 c_f m_{s,\text{BS},o} \quad (R^2 = 1.00; n = 5) \quad (131)$$

where,

$$100 \leq m_{s,\text{BS},o} \leq 2,000 \text{ lb/hr}$$

Unlike the other consumable chemicals and catalysts, data are not available regarding the makeup mass flow rate for the Stretford chemicals. However, data are available regarding the cost of the Stretford chemicals. The regression model shown in Equation (131) is the cost of the Stretford chemicals, in dollars, as a function of the sulfur recovered in the Beavon-Stretford process. The standard error of the model is negligible. The model is shown graphically in Figure 54.

Makeup Zinc Ferrite Sorbent

Makeup zinc ferrite sorbent is required to replace sorbent that disintegrates into fines and that may be entrained with the syngas or cause a significant increase in the pressure drop across the reactor. A replacement schedule of 20 percent of the sorbent in the reactors is suggested every 80 cycles (Kasper, 1988). Spent sorbent is assumed to be returned to the manufacturer for reprocessing (Banchik, Buckman, and Rath, 1988). The annual required replacement rate is therefore given by:

$$m_{\text{zf,ZF},i} = \frac{8,760 c_f}{2 t_a} \left(\frac{f_{\text{att}} S_c}{80} \right) \quad (132)$$

where, as a default, it is assumed that $f_{\text{att}} = 0.20$.

Makeup Sulfuric Acid Plant Catalyst

The sulfuric acid plant, which is assumed for sulfur recovery in the Lurgi-based system, requires makeup catalyst. The number of liters of makeup catalyst depends on the design capacity of the sulfuric acid plant. Based on data from Monsanto (Higgins, 1988), the annual makeup catalyst, in liters, is approximated as:

$$m_{\text{cat,SA},i} = 1.1 M_{\text{SA},i} \quad (133)$$

Figure 54. Annual Chemical Cost for the Beavon-Stretford Process



6.2.4 Other Consumables

Other consumables that are required for all systems are gas turbine startup fuel, plant and instrument air adsorbent, wastewater treatment chemicals, water, and liquified petroleum gas for flares. For the KRW-based system with hot gas cleanup, limestone is required for in-bed desulfurization. For the Lurgi-based system, coke is required as a gasifier startup fuel, and bentonite is required as a binder for fines agglomeration.

The consumables required for all systems are estimated based on a simple scaling relationship with plant size, on the premise that the requirement of these items scales directly with the plant capacity. The requirements for these items are based on data reported in Fluor (1985), for a nominal 550 MW plant.

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

$$m_{fo,GT,i} = 48,000 c_f \left(\frac{MW}{550} \right) \quad (134)$$

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

$$m_{ads,GF,i} = 3,600 c_f \left(\frac{MW}{550} \right) \quad (135)$$

The wastewater chemical requirement is given in dollar, not mass, terms. The reported cost of wastewater chemicals is given in Table 17 as an equivalent annual cost, at 100 percent capacity factor, based on the wastewater flow rate. The cost of wastewater chemicals is therefore given by:

$$C_{ww} = c_f UC_{ww} m_{ww} \quad (136)$$

The plant water consumption is given by the total of the raw water consumption for the steam cycle and the cooling water makeup water. The total is:

$$m_{water} = 8,760 c_f m_{rw} \quad (137)$$

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

$$m_{LPG} = 4,200 c_f \left(\frac{MW}{550} \right) \quad (138)$$

The limestone requirement for in-bed desulfurization depends on the sulfur content of the coal and the calcium-to-sulfur molar ratio required to achieve the design in-bed sulfur removal efficiency. For 90 percent sulfur removal, a molar ratio of 1.8 is commonly assumed (Smelser, 1986; Smith and Smelser, 1987; Earley and Smelser, 1988). Therefore, the annual limestone requirement, in pounds, is given by:

$$m_{limestone} = 3.13 \left(\frac{R_{Ca/S}}{f_{CaCO_3}} \right) m_{coal} f_s \quad (139)$$

where,

$$m_{coal} = 8,760 c_f m_{cf,CH,i} \quad (140)$$

and,

$R_{Ca/S}$ = molar ratio of calcium to sulfur, with a default value of 1.8.

f_{CaCO_3} = weight percent calcium carbonate in limestone. Default is 0.95.

The coke requirement for Lurgi-based systems is assumed to be directly proportional to the plant energy output, based on data in Bechtel et al (1988):

$$m_{coke} = 610 c_f \left(\frac{MW}{180} \right) \quad (141)$$

The required quantity of coke is expressed in tons.

The bentonite requirement is assumed to be directly proportional to the fines flow rate through the gasifier. The bentonite requirement for coal fines agglomeration is reported as 26 tons per day to agglomerate 600 tons per day of coal fines (Bechtel et al, 1988). Therefore, the annual bentonite requirement, in pounds, is assumed to be:

$$m_{bent} = 0.043 (8,760) c_f m_{fines} \quad (142)$$

6.2.5 Ash Disposal and Byproducts

The annual ash disposal requirement includes bottom ash from the gasifiers and ash recovered from the syngas in cyclones or scrubbers. In the case of the systems with hot gas cleanup, all ash is in dry form. In the case of the KRW-based system with cold gas cleanup, a portion of the ash is disposed of as filter cake from the syngas particulate scrubbing unit, which

contains typically about 60 percent dry ash (e.g., Fluor, 1985). The total ash to disposal may be approximately estimated as the total ash content in the coal:

$$m_{\text{ash,o}} = m_{\text{coal}} f_{\text{ash}} \quad (143)$$

The amount of sulfur recovered in the Claus plant (and Beavon-Stretford unit, if applicable) for the KRW-based system with cold gas cleanup, or the amount of sulfuric acid recovered for the Lurgi-based system, is the basis for estimating the byproduct credit. These flows are estimated in the ASPEN simulations.

6.2.6 Total Variable Operating Costs

To estimate the total variable operating cost, the annual material requirements appropriate to the given IGCC system must be multiplied by their respective unit costs. In two cases (Beavon-Stretford chemicals and wastewater treatment chemicals) the unit costs are based on a process flow rate (sulfur recovered in the Beavon-Stretford unit and the wastewater discharge rate) because the material requirements of the consumables themselves are not reported. The variable operating costs should be reported in the same categories as presented in Table 14 (i.e., fuel, consumables, ash disposal, and byproduct credit). The major components of the total variable operating cost, expressed in dollars, are:

$$OC_{\text{fuel}} = m_{\text{coal}} UC_{\text{coal}} \quad (144)$$

$$OC_{\text{cons}} = \sum_{\text{cons}} m_i UC_i \quad (145)$$

$$OC_{\text{ash}} = m_{\text{ash,o}} UC_{\text{ash}} \quad (146)$$

$$OC_{\text{byp}} = (1 - f_{\text{bm}}) m_{\text{byp}} UC_{\text{byp}} \quad (147)$$

where,

f_{bm} = fraction of byproduct revenue required for marketing and shipping costs. The default is assumed to be 10 percent, or $f_{\text{bm}} = 0.10$.

The total variable cost is then:

$$VOC = OC_{\text{fuel}} + OC_{\text{cons}} + OC_{\text{ash}} - OC_{\text{byp}} \quad (148)$$

6.3 Preproduction Costs and Inventory Capital

In Section 5, preproduction costs are summarized in Equation (87). These costs include one month of fixed operating costs; one month of variable operating costs excluding fuel and byproduct credit; one-quarter of the fuel requirement for one month of operation at full load; and an additional allowance based on a percentage of the total plant investment. Note that the operating costs are expressed in units of dollars, while the capital 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 presented earlier. One month of fixed operating cost (in thousands of dollars) is given by:

$$PP_{FC} = \frac{1}{12} \left(\frac{FOC}{1,000} \right) \quad (149)$$

Similarly, one month of variable operating costs, excluding fuel and byproduct credits, can be estimated as follows:

$$PP_{OC} = \left(\frac{0.083}{c_f} \right) \left(\frac{OC_{cons} + OC_{ash}}{1,000} \right) \quad (150)$$

The preproduction fuel cost is given by:

$$PP_{Fuel} = \left(\frac{0.021}{c_f} \right) \left(\frac{OC_{fuel}}{1,000} \right) \quad (151)$$

The preproduction capital costs are then fully specified by substituting Equations (149), (150), and (151) into Equation (87).

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 as follows:

$$IC = \left(\frac{0.164}{c_f} \right) \left(\frac{OC_{fuel} + OC_{cons}}{1,000} \right) \quad (152)$$

Equation (152) can be substituted for Equation (88) in Section 5.

6.4 Auxiliary Electrical Requirements

A number of process areas consume significant amounts of electric power, thereby reducing the saleable electrical output of the power plant. When reporting costs on a normalized basis (e.g., \$/kW or mills/kWh), it is important to use an accurate estimate of the net electrical production available for sale. For many process areas, the ASPEN performance models do not estimate the internal electrical loads. Therefore, simple regression models of power consumption versus key flow rates have been developed for major process sections. These models provide more accurate estimates of the plant electrical requirements than are currently available in the ASPEN IGCC simulation models.

For each major plant section for which power consumption data are available, a regression model was developed. These models are presented below.

KRW Coal Handling

$$W_{e,CH} = 378 + 0.533 m_{cf,CH,i} \quad (R^2 = 0.918; n = 6) \quad (153)$$

where,

$$5,900 \leq m_{cf,CH,i} \leq 27,000 \text{ tons/day}$$

The standard error of the estimate is 1,580 kW. The regression model is shown graphically in Figure 55.

Oxygen Plant

$$W_{e,OF} = 6.09 M_{O,G,i} \quad (R^2 = 0.967; n = 27) \quad (154)$$

where

$$1,250 \leq M_{O,G,i} \leq 22,700 \text{ lbmole/hr.}$$

The standard error of the estimate is 6,600 kW. The regression model is shown graphically in Figure 56. A factor to adjust the power consumption for changes in ambient temperature was developed based on analysis of data from Fluor (1985). This factor is:

$$f_{e,OF} = 0.9466 + 3.73 \times 10^{-3} T_a + 9.019 \times 10^{-6} T_a^2 \quad (155)$$

This factor has a value of 1.0 at an ambient temperature of 59 °F, and it is valid only for temperatures between 20 °F and 95 °F. The modified expression for oxidant feed power consumption is therefore given by:

$$W_{e,OF} = 6.09 (0.9466 + 3.73 \times 10^{-3} T_a + 9.019 \times 10^{-6} T_a^2) M_{O,G,i} \quad (156)$$

Figure 55. Power Requirement for KRW Coal Handling

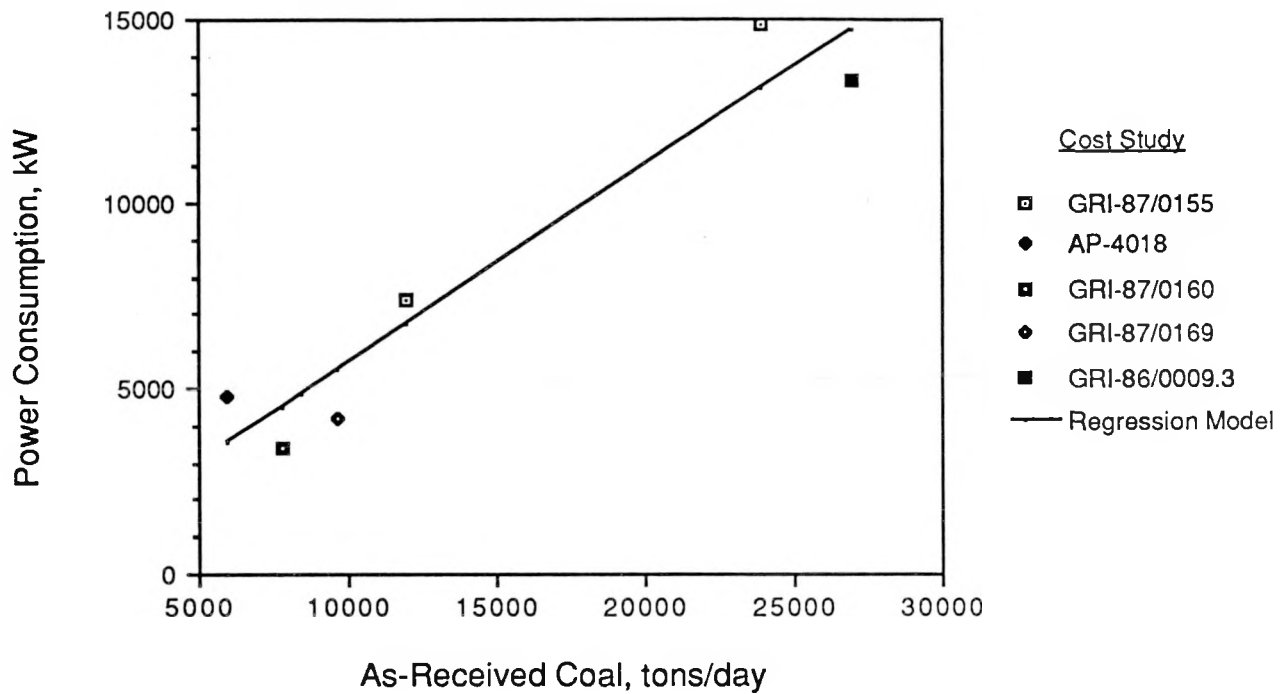
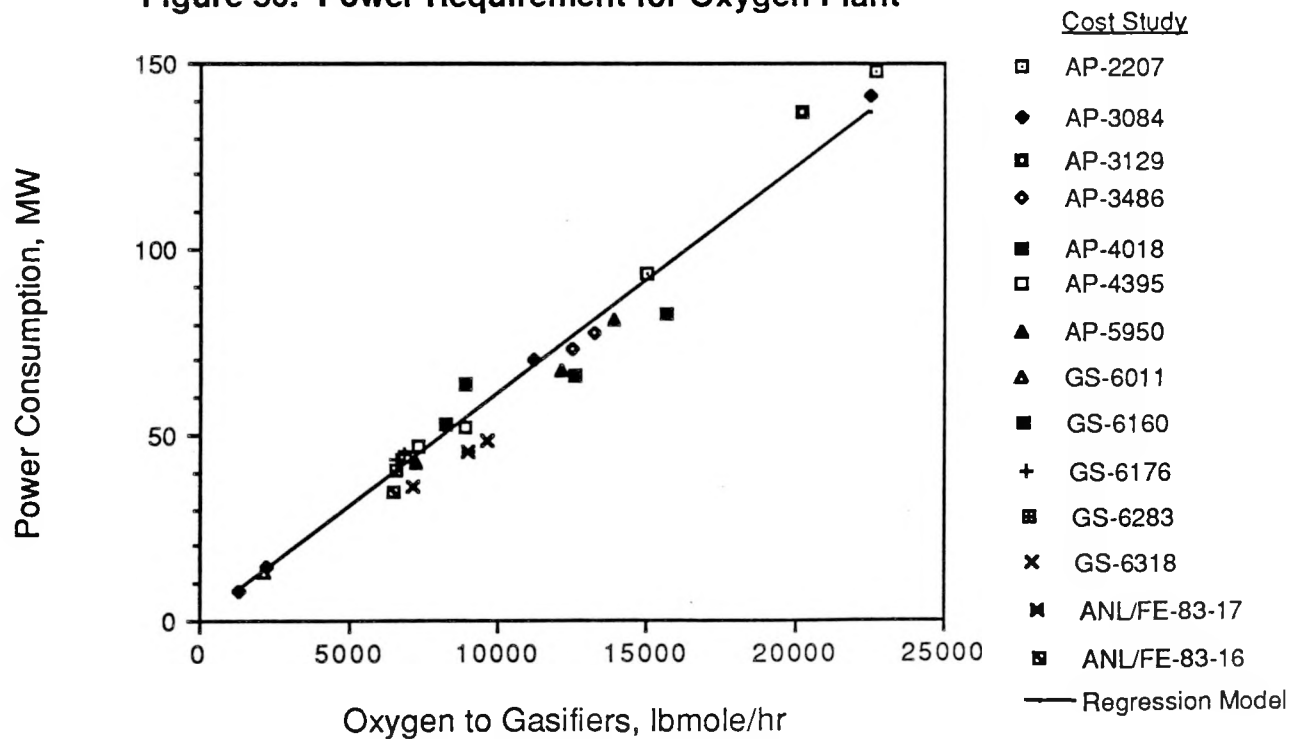


Figure 56. Power Requirement for Oxygen Plant



KRW Gasification (Cold Gas Cleanup)

$$W_{e,G} = 0.0476 M_{\text{syn,G,o}} \quad (R^2 = 0.894; n = 7) \quad (157)$$

where,

$$35,400 \leq M_{\text{syn,G,o}} \leq 120,600 \text{ lbmole/hr.}$$

The standard error of the estimate is 520 kW. The regression model is shown graphically in Figure 57.

Low Temperature Gas Cooling

$$W_{e,LT} = -5,600 + 0.0108 M_{\text{syn,LT,o}} + 14.19 P_{\text{syn,LT,o}} \quad (R^2 = 0.890; n = 10) \quad (158)$$

where,

$$32,300 \leq M_{\text{syn,LT,o}} \leq 74,400 \text{ lbmole/hr}$$

$$373 \leq P_{\text{syn,LT,o}} \leq 463 \text{ psia.}$$

For this plant section, a multivariate regression was found to yield significantly better results than a single-variate analysis. The standard error of the estimate is 240 kW. A comparison of reported and predicted power consumption is given in Figure 58.

Selexol Process

$$W_{e,S} = 348 + 0.478 \left(M_{\text{syn,S,i}} \right)^{0.839} \quad (R^2 = 0.881; n = 18) \quad (159)$$

where,

$$4,000 \leq M_{\text{syn,S,i}} \leq 74,500 \text{ lbmole/hr.}$$

The standard error of the estimate is 550 kW. The regression model is shown graphically in Figure 59.

Claus Plant

$$W_{e,C} = 0.021 m_{\text{s,C,o}} \quad (R^2 = 0.870; n = 20) \quad (160)$$

where,

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

The standard error of the estimate is 67 kW. The regression model is shown graphically in Figure 60.

**Figure 57. Power Requirement for the Gasification Section
(KRW System with Cold Gas Cleanup)**

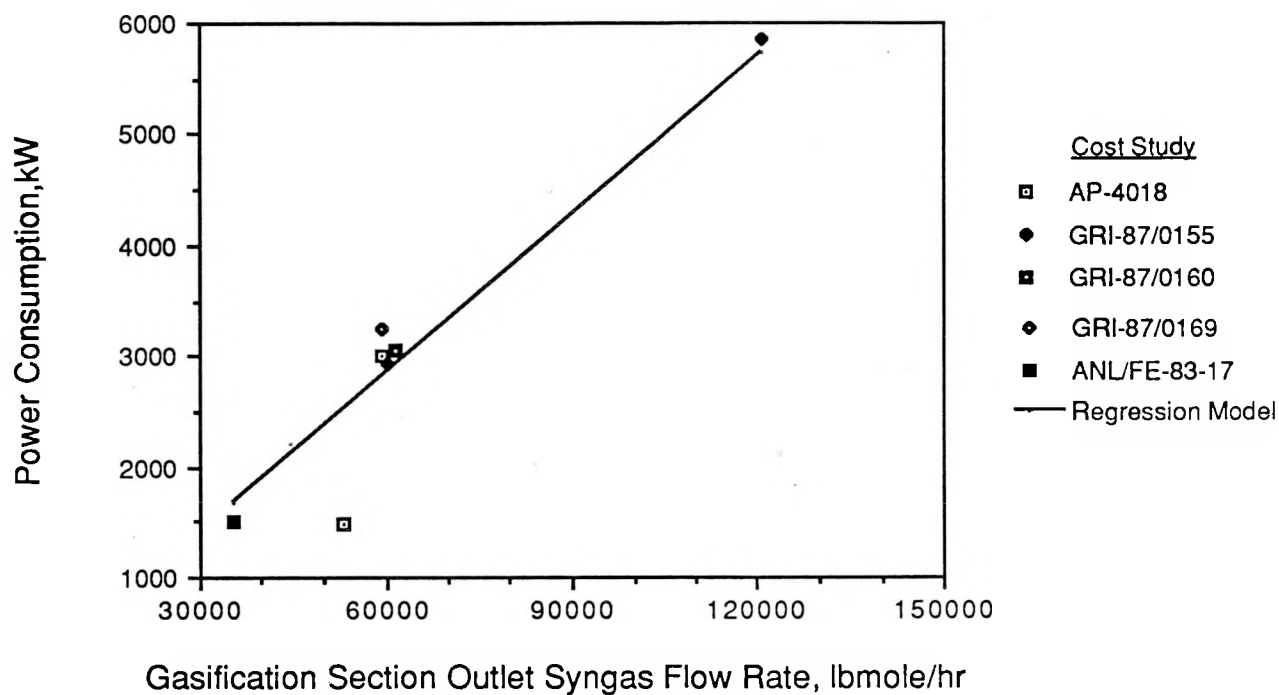


Figure 58. Power Consumption of Low Temperature Cooling Section

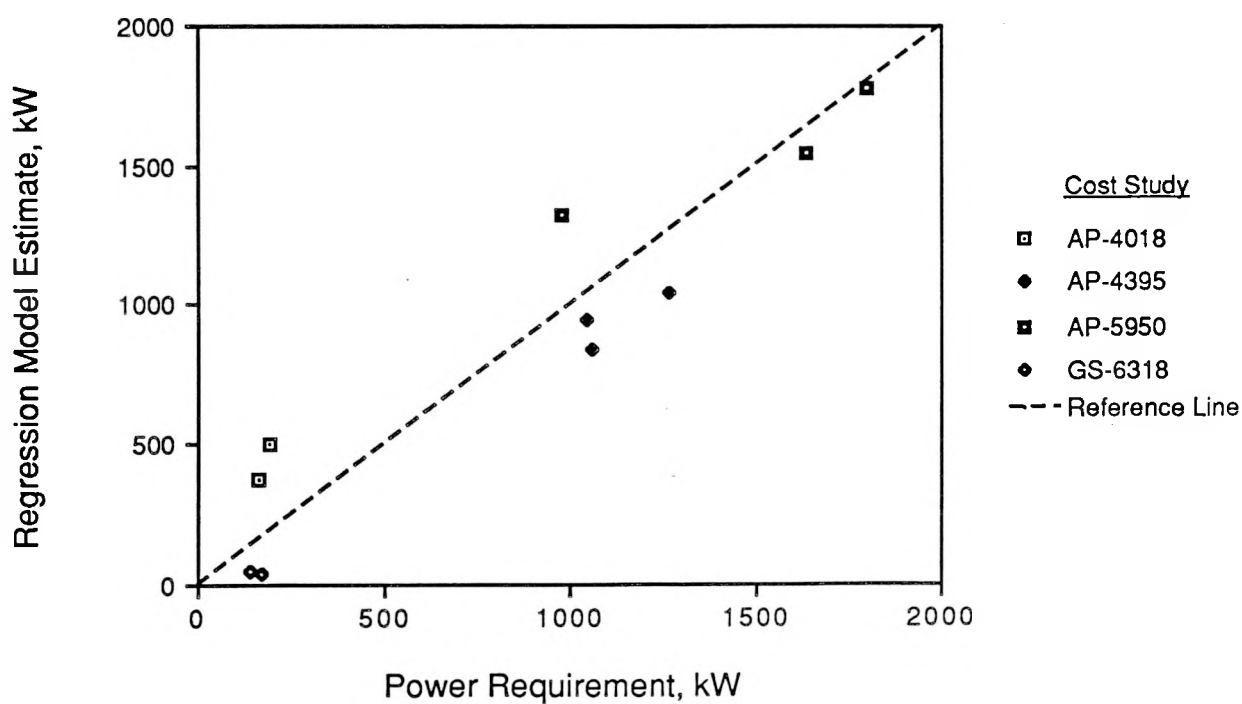


Figure 59. Power Requirement of the Selexol Units

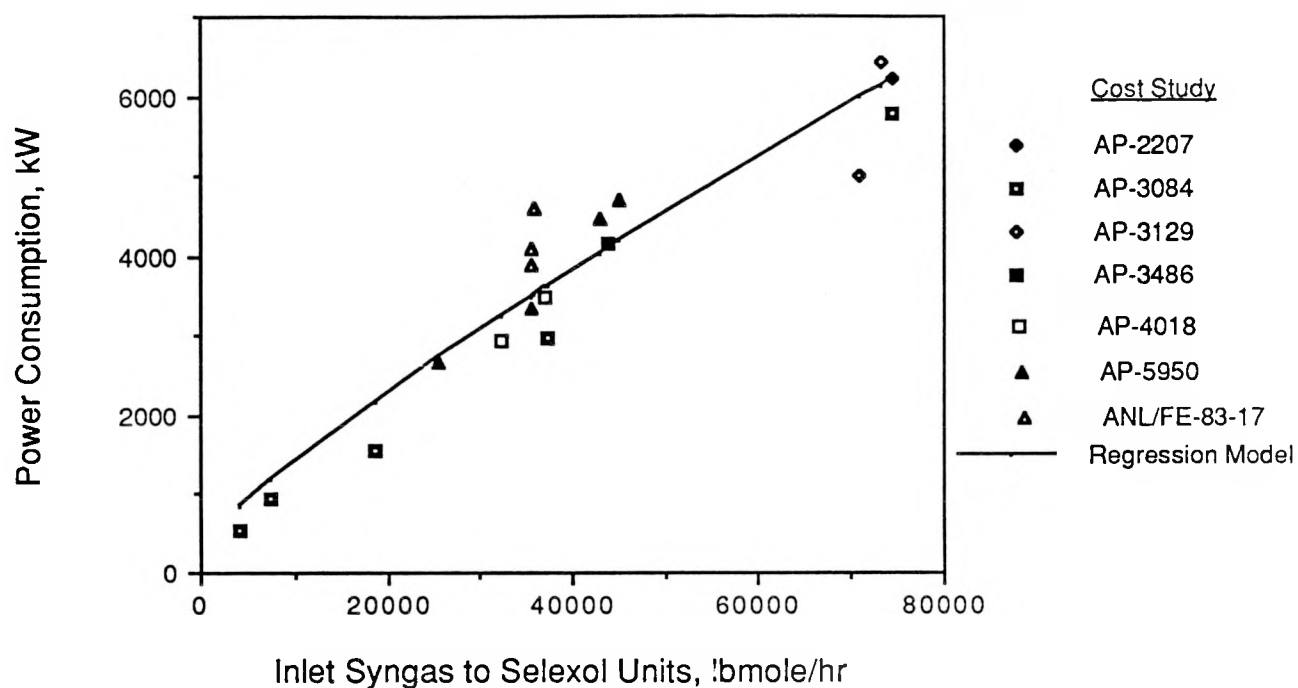
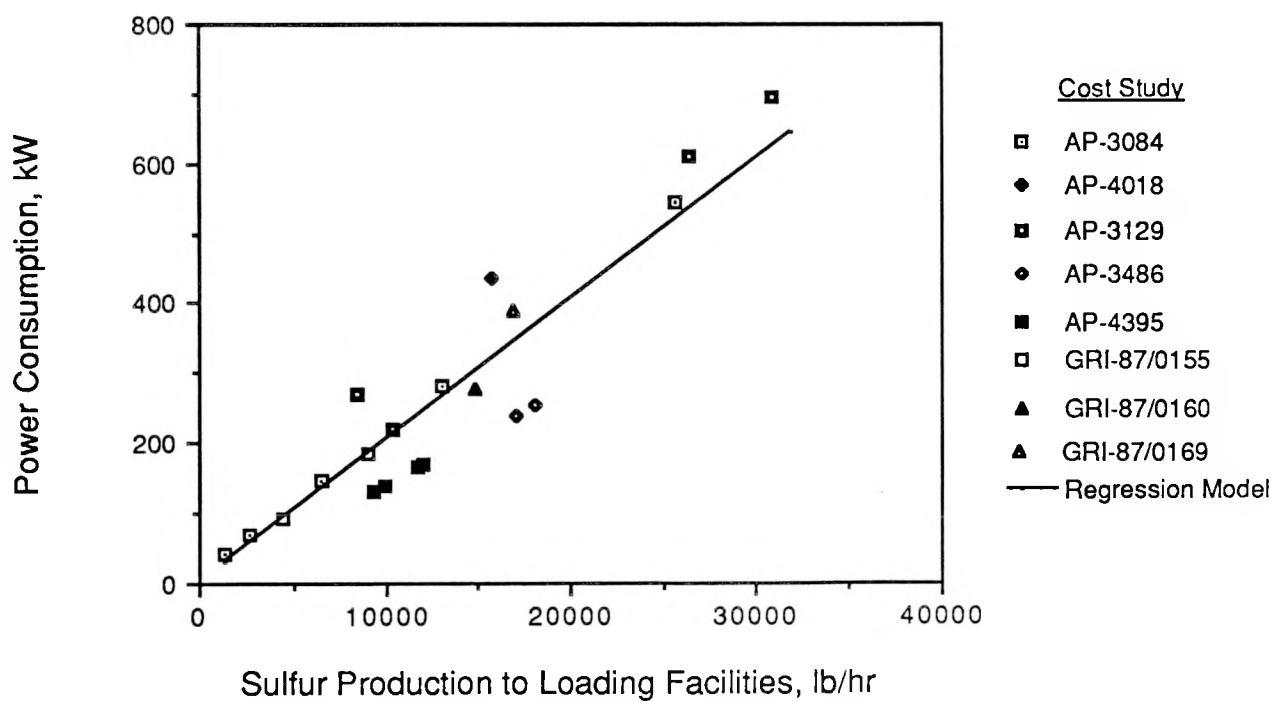


Figure 60. Power Requirement for Two-Stage Claus Plants



SCOT Process

$$W_{e,SC} = 2.4 + 0.0408 m_{s,C,o} \quad (R^2 = 1.00; n = 6) \quad (161)$$

where,

$$9,000 \leq m_{s,C,o} \leq 18,000 \text{ lb/hr}$$

The standard error of the estimate for the data points included in the analysis is negligible. Four outlier data points were excluded from the regression analysis. Two of the excluded points were from the least recent study (Fluor, 1983b). The other two outliers are from a separate study (Fluor, 1985). Although one of these points is close to the regression line, the other point is an extreme outlier. The regression model is shown graphically in Figure 61.

Beavon-Stretford Process

$$W_{e,BS} = 44.5 + 1.12 m_{s,BS,o} \quad (R^2 = 0.990; n = 7) \quad (162)$$

where,

$$100 \leq m_{s,BS,o} \leq 2,000 \text{ lb/hr}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure 62.

Boiler Feed Water Treating

$$W_{e,BF} = 20.8 + 2.13 \times 10^{-4} m_{pw} \quad (R^2 = 0.975; n = 14) \quad (163)$$

where,

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

The standard error of the estimate is 38 kW. The regression model is shown graphically in Figure 63.

Process Condensate Treatment

$$W_{e,PC} = 7.34 \times 10^{-4} m_{sbd} \quad (R^2 = 1.00; n = 3) \quad (164)$$

where,

$$196,000 \leq m_{sbd} \leq 237,000 \text{ lb/hr}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure 64.

Figure 61. Power Requirement for the SCOT Process

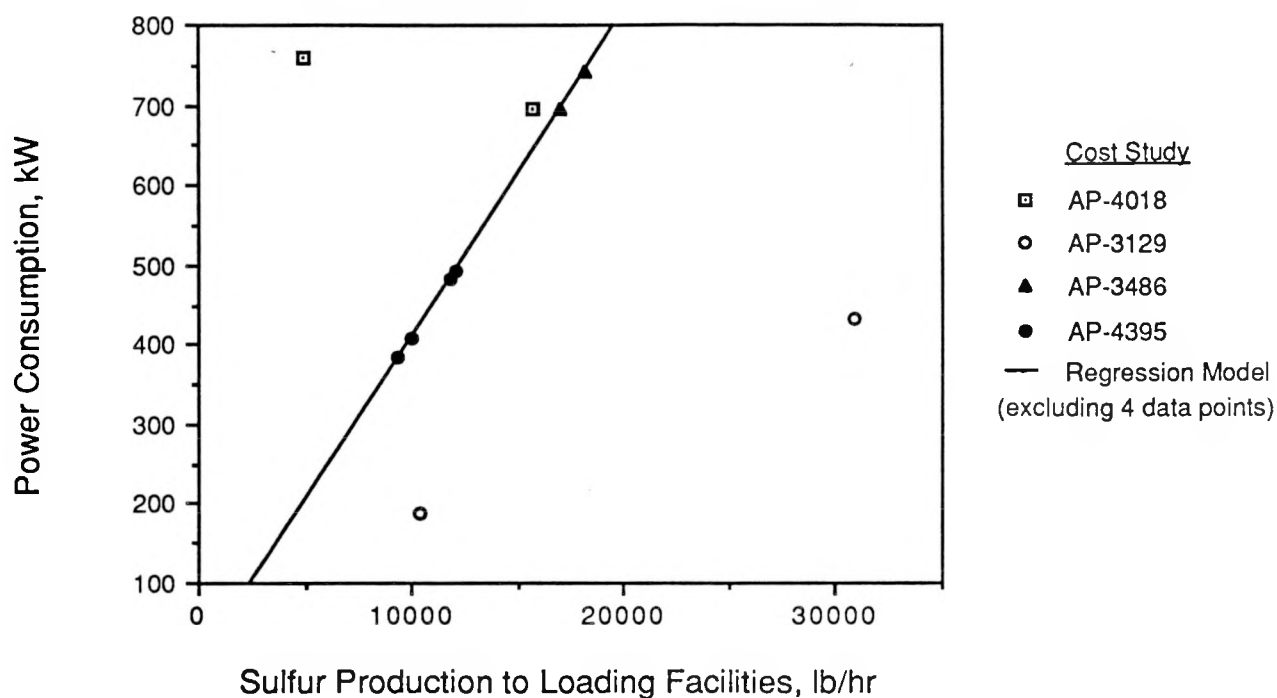


Figure 62. Power Requirement for the Beavon-Stretford Process

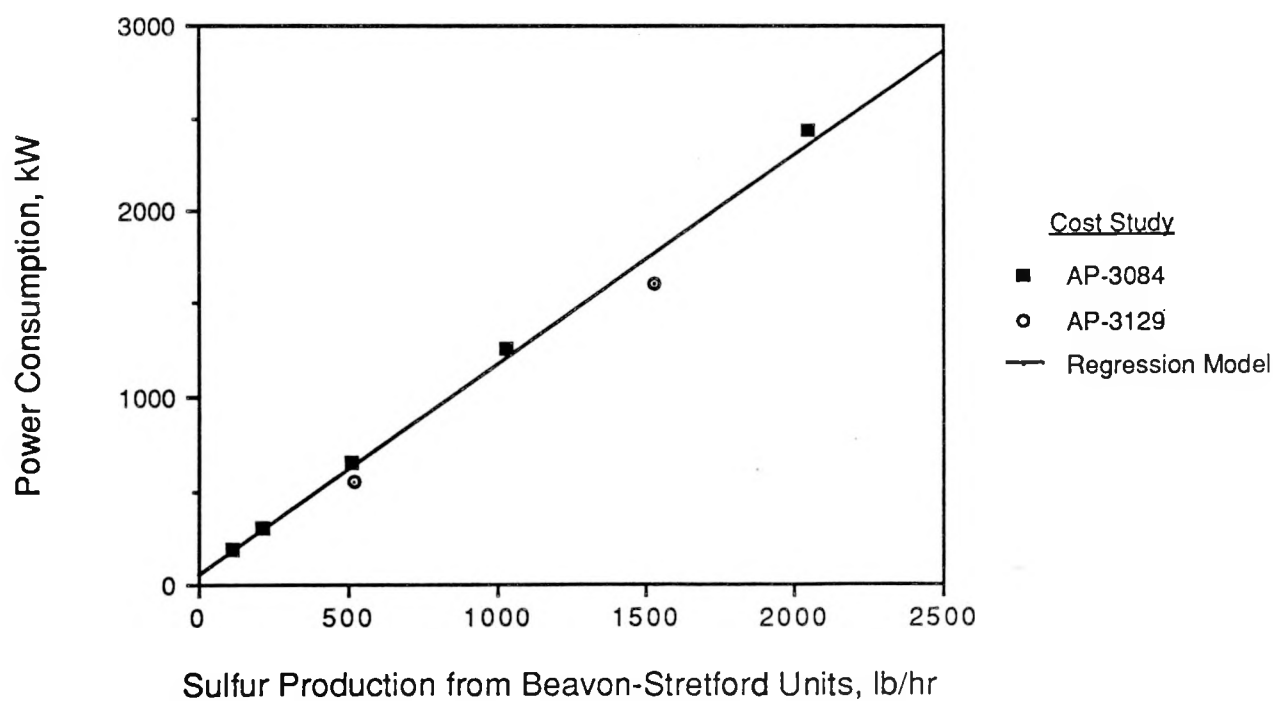


Figure 63. Power Requirement for Boiler Feed Water Treating

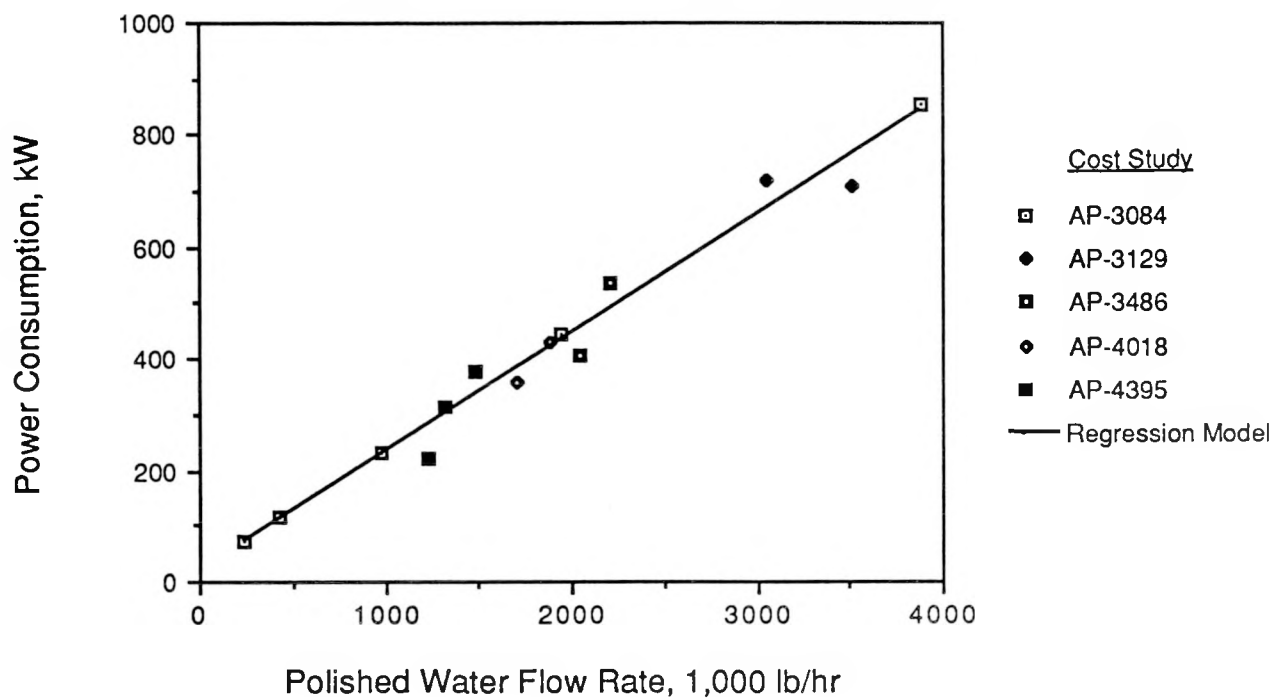
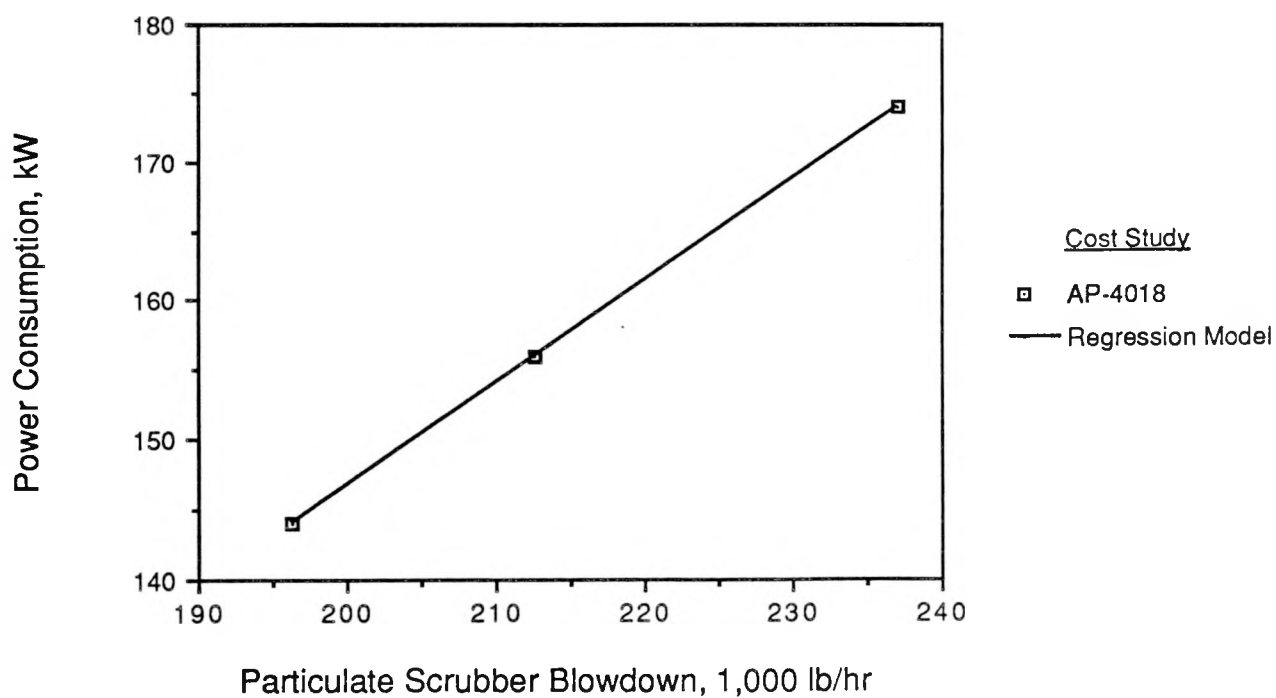


Figure 64. Power Requirement for Process Condensate Treatment



General Facilities

$$W_{e,GF} = 1,640 + 0.0877 \sum_{i \neq GF} W_{e,i} \quad (R^2 = 0.999; n = 4) \quad (165)$$

where,

$$58,000 \leq \sum_{i \neq GF} W_{e,i} \leq 82,000 \text{ kW.}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure 65.

Limestone Handling

$$W_{e,L} = -13.3 + 4.61 \times 10^{-3} m_{L,L,i} \quad (R^2 = 1.00; n = 3) \quad (166)$$

where,

$$5,000 \leq m_{L,L,i} \leq 104,000$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure 66.

KRW Gasification (Hot Gas Cleanup)

$$W_{e,G} = 58 + 8.3 \times 10^{-3} m_{cm,G,i} \quad (R^2 = 1.00; n = 3) \quad (167)$$

where,

$$30,000 \leq m_{cm,G,i} \leq 600,000 \text{ lb/hr.}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure 67.

Sulfation

$$W_{e,SF} = -3.7 + 0.0102 (m_{\text{fines}} + m_{\text{ash}}) \quad (R^2 = 1.00; n = 3) \quad (168)$$

where,

$$10,000 \leq (m_{\text{ash}} + m_{\text{fines}}) \leq 155,000 \text{ lb/hr.}$$

The standard error of the estimate is negligible. The regression model is shown graphically in Figure 68.

Lurgi Coal Handling

The power requirement for the Lurgi coal handling section consists of two components: coal preparation and coal receiving and storage. The coal preparation power requirement is associated with screening of the coal to separate fines prior to feed to the gasifier, as well as

Figure 65. General Facilities Power Requirement

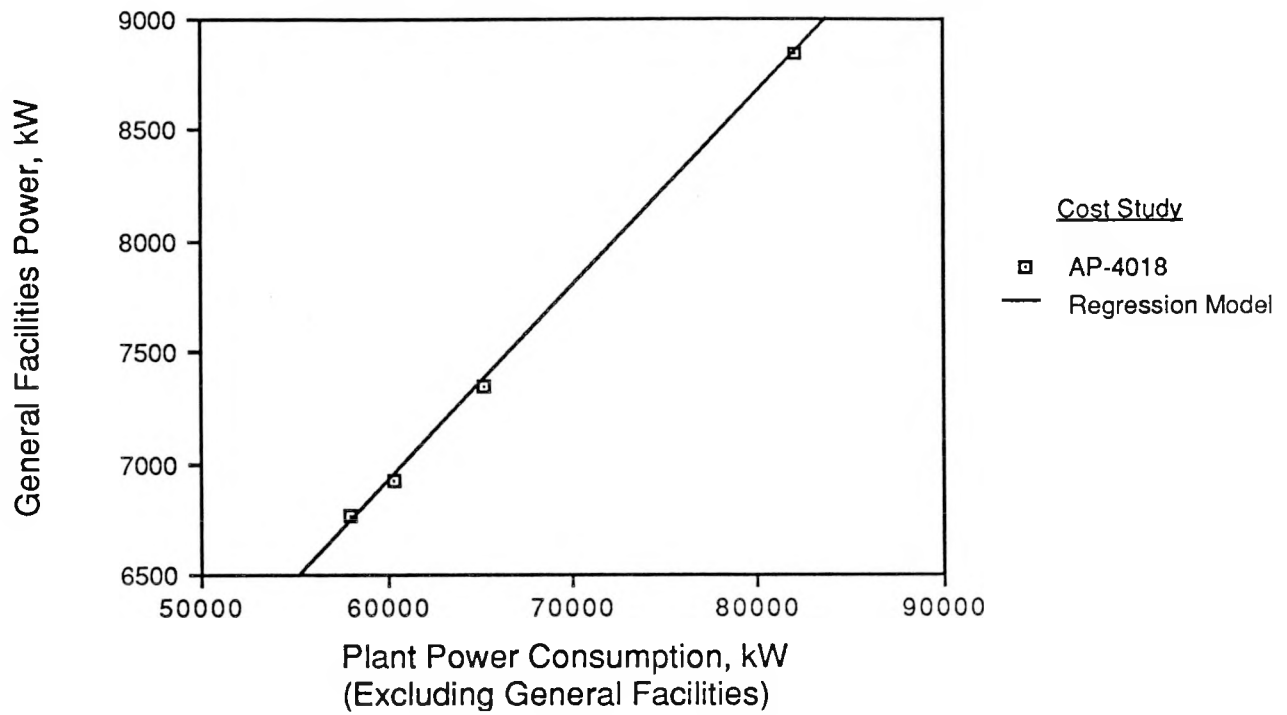


Figure 66. Power Requirement for Limestone Handling

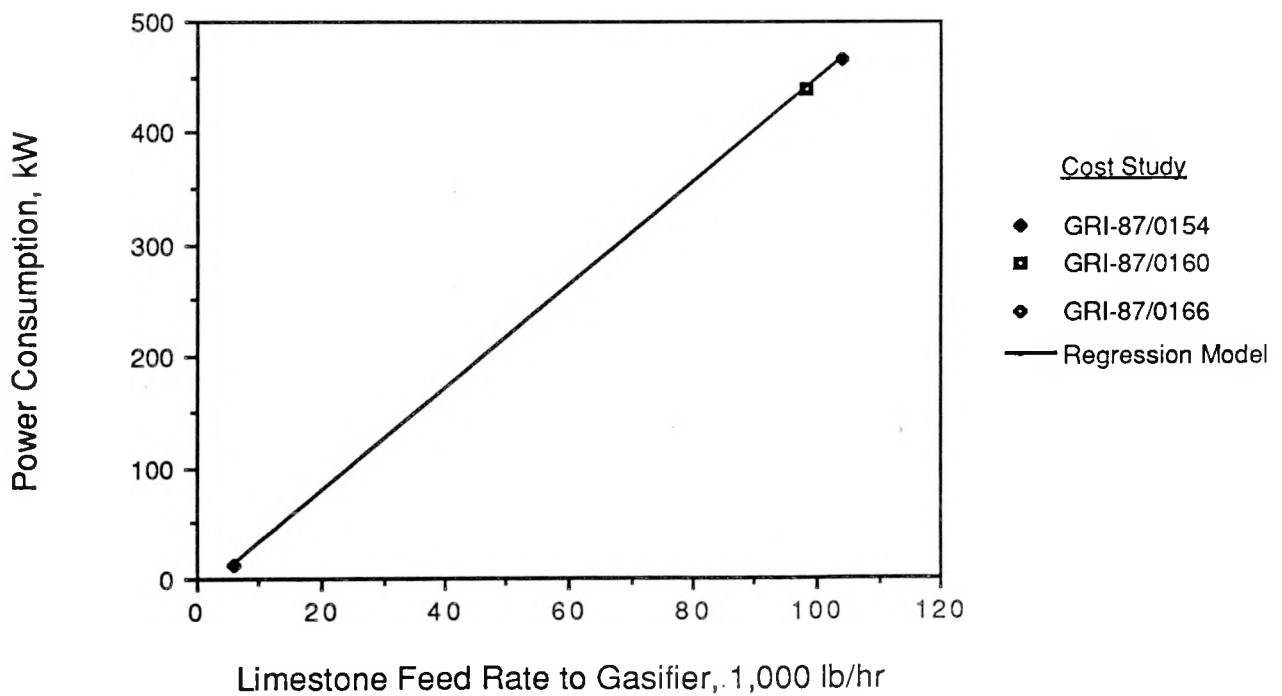


**Figure 67. Power Requirement for the Gasification Section
(KRW System with Hot Gas Cleanup)**

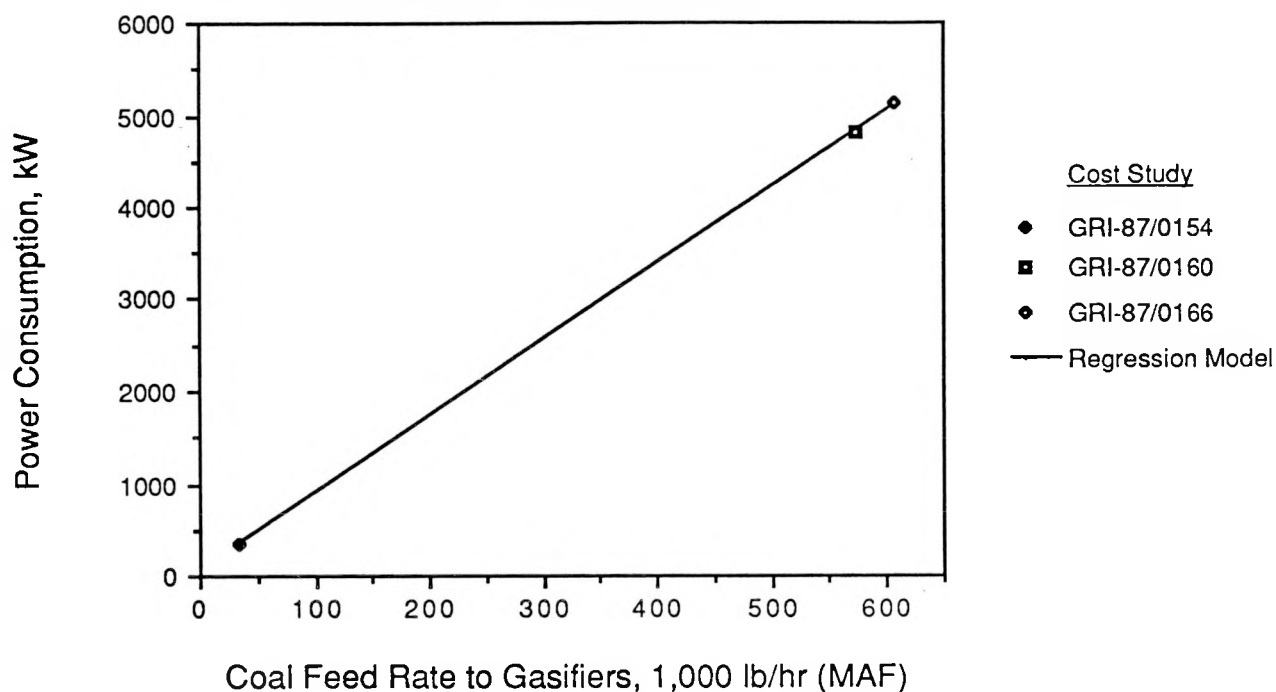
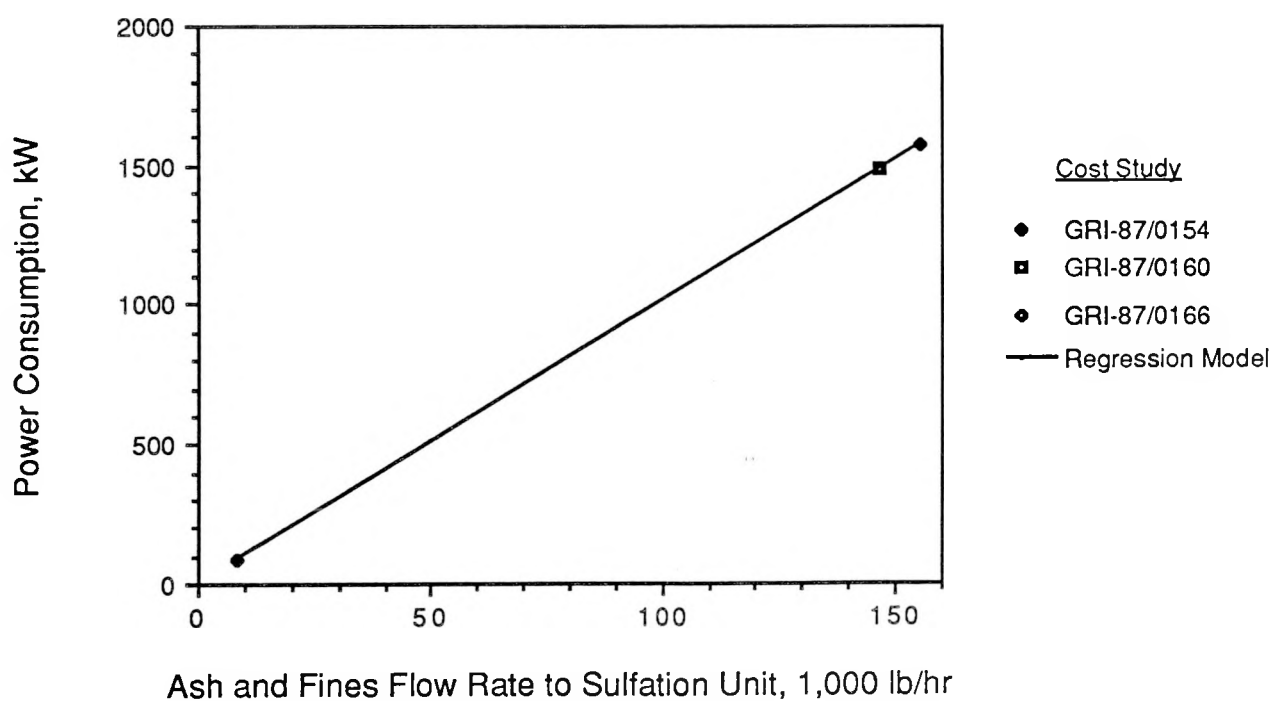


Figure 68. Power Requirement for the Sulfation Unit



transport of the coal from storage to the gasifier coal surge bins. The coal preparation power requirement is given by:

$$W_{e,CH,prep} = 0.00133 m_{cf,CH,i} \quad (R^2 = 0.989; n = 10) \quad (169)$$

where,

$$675,000 \leq m_{cf,CH,i} \leq 3,380,000 \text{ lb/hr.}$$

The standard error of the estimate is 120 kW. The regression model is shown graphically in Figure 69.

The power requirement for coal receiving and storage was assumed to be directly proportional to the coal feed rate. One data point was available (Smelser, 1986a). The coal receiving and storage power requirement is given by:

$$W_{e,CH,R\&S} = 5.6 \times 10^{-4} m_{cf,CH,i} \quad (170)$$

The total power requirement for the coal handling section is the sum of the coal preparation and coal receiving and storage power requirements, and is given by:

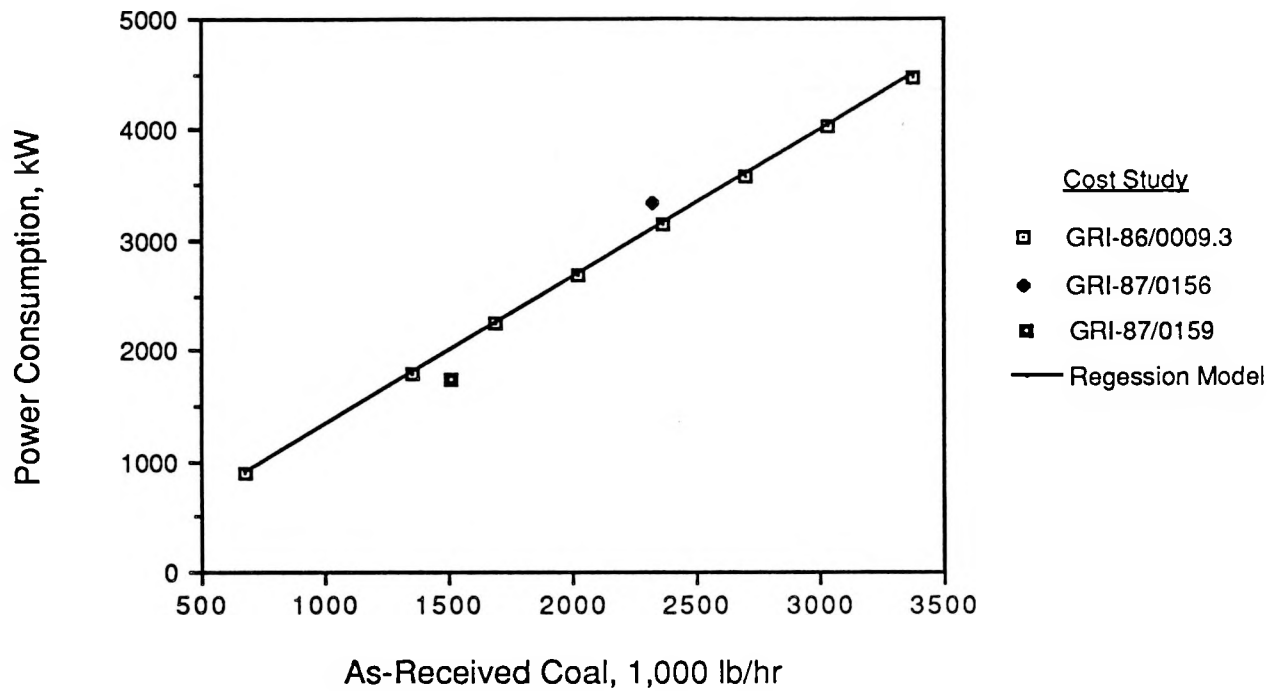
$$W_{e,CH} = W_{e,CH,prep} + W_{e,CH,R\&S} \quad (171)$$

In Section 4.1, it is noted that the study upon which the cost model of the Lurgi system is based assumes direct feed of all of the received coal, including fines, to the gasifier (Corman, 1986). Therefore, the power requirement given in Equation (171) may be an over-estimate. However, if problems occur as a result of fines in the gasifier, screening of the fines prior to coal feed would be required, and in that event Equation (171) is an appropriate model. The use of Equation (171) for the coal handling power requirement is therefore conservative.

Lurgi Gasification

The power requirement for Lurgi gasification was estimated by subtracting the coal receiving and storage power requirement estimated from Equation (170) and 50 percent of the coal preparation power requirement given in Equation (169) from the total power requirement for coal handling and gasification reported in Corman (1986). Only 50 percent of the coal preparation power estimate was used because the Corman study does not assume coal fines

Figure 69. Power Requirement for Lurgi Coal Preparation



separation. The gasification section power requirement was assumed to be directly proportional to the coal feed rate, and is given by:

$$W_{e,G} = 0.005 m_{cf,G,i} \quad (172)$$

Sulfuric Acid Plant

The power requirement for the sulfuric acid plant is assumed to be directly proportional to the inlet gas flow rate. Based on the estimate reported by SMC (1983), the power requirement is given by:

$$W_{e,SA} = 0.9 M_{SA,i} \quad (173)$$

6.5 Representing Uncertainty in Operating Cost Estimates

Unlike capital cost estimates, there is no widely accepted notion of contingency costs for operating costs. However, uncertainties in operating costs can have a significant impact on the overall uncertainty in the cost of a new technology process plant. These uncertainties may include: potential difficulties with operation and control of new processes, requiring additional operating labor or more highly trained (and paid) operators; problems with fouling, corrosion, or other material degradation requiring additional maintenance; variability or uncertainty in the unit costs of catalysts and chemicals, especially for new materials that are not yet commercially available (e.g., zinc ferrite sorbent); uncertainty regarding the requirement for catalysts and chemicals due to uncertainty in process performance (e.g., stoichiometric ratio for calcium-to-sulfur for in-bed desulfurization); uncertainty in the availability of ash disposal sites and the cost of ash disposal; uncertainty regarding the availability of byproduct markets and the price of byproducts; and uncertainty or variability in fuel prices. These types of uncertainties can be explicitly characterized and quantified using probability distributions and a Monte Carlo (or similar) simulation technique, as discussed previously in Section 1.7.

The cost of the zinc ferrite sorbent initial charge and annual makeup requirement is a case in point. This sorbent has not been mass produced for commercial use. Therefore, in preparing cost estimates, cost estimators typically rely on the judgment of catalyst suppliers regarding the likely commercial cost of the sorbent. The sorbent unit cost is projected to fall within a range of values, but cannot be predicted with certainty. One study indicates that a major catalyst manufacturer has estimated the sorbent unit cost at between \$3.00 and \$3.50 per pound, while cost estimators at METC had been using a value of \$2.50 per pound (Klett et al, 1986). If the unit cost of the sorbent is nominally estimated at \$3.00 per pound, then the

range of costs reported (from \$2.50 to \$3.50 per pound) represents 33 percent of the cost estimate. In another study (Banchik and Cover, 1988), the unit cost of zinc ferrite sorbent was assumed to vary from \$4.00 to \$10.00 per pound. Therefore, the reported range of unit costs from the open literature vary by a factor of four. If in addition the uncertainty in the *amount* of zinc ferrite sorbent required is considered, which results from uncertainty in the long term sulfur loading capability of the sorbent and the sorbent attrition rate, the overall uncertainty in the initial and annual total cost for zinc ferrite sorbent may be significantly greater than a factor of four. Yet these types of interactions between uncertainties in performance and cost parameters are not characterized in traditional deterministic cost estimates.

The unit cost of the zinc ferrite sorbent can be represented using a probability distribution. Similarly, the long term sorbent loading and the sorbent attrition rate can be represented using probability distributions. The probability distributions must be selected to represent judgments about the most likely values of these parameters, as well as the possible range in values that these parameters may assume. Using a probabilistic modeling approach, the effect of these uncertainties taken *simultaneously* on the total cost for zinc ferrite sorbent can be quantified. Also, a correlation analysis can be done to identify which uncertain model parameters are the most significant in determining the uncertainty in overall cost. Probabilistic analysis is therefore a valuable quantitative tool for identifying research priorities.

As for the capital cost models, an important application of the operating cost models for the selected IGCC systems will be the development of probabilistic estimates of IGCC system cost based on a disaggregated consideration of uncertainties in the parameters of the models. The development of the probabilistic parameter estimates, and application of these in specific case studies, is addressed in a later phase of the work described here.

7.0 TOTAL ANNUALIZED COST MODEL

The total annualized cost is the levelized annual revenue requirement required to cover all of the capital and operating costs for the economic life of the plant. For electric power plants, the total annualized cost is typically expressed as the cost of electricity. The total capital requirement, fixed operating cost, and variable operating cost are used to calculate the cost of producing electricity that is available for sale from the power plant, based on the net electrical output of the power plant. The net power output is the total power generated from the gas turbines and steam turbine less the total auxiliary power demand:

$$MW_{net} = MW_{GT} + MW_{ST} - 1,000 \sum_i W_{e,i} \quad (174)$$

The cost of electricity in mills (one-thousands of a dollar) per net kWh is given by:

$$C_{elec} = \frac{[1,000 f_{cr} TCR + f_{vclf} (FOC + VOC)] \left(\frac{1,000 \text{ mills}}{\text{dollar}} \right)}{MW_{net} \times 8,760 \times c_f} \quad (175)$$

where,

$$\begin{aligned} f_{cr} &= \text{capital recovery factor} \\ f_{vclf} &= \text{variable cost levelization factor} \end{aligned}$$

The numerator of Equation (175) is the total annual revenue requirement for the plant, and the denominator is the total net kilowatt-hours of electricity generated in a year. The total capital requirement from Equation (102) is in units of thousands of dollars. Therefore, a factor of 1,000 is used to convert the total capital requirement to units of dollars. The fixed operating cost from Equation (110) and the variable operating cost from Equation (149) are both in units of dollars. The annual revenue requirement shown in square brackets in the numerator is converted from dollars to mills, which is a more convenient unit for reporting the cost of electricity.

The capital recovery factor converts the capital cost into the equivalent levelized annual revenue required to provide a return to equity (stock) and debt (loan) financing sources, to pay for a portion of the principal, and to pay associated taxes and insurance (EPRI, 1987). The capital recovery factor, therefore, depends on the economic life of the plant, the type of financing used to supply the capital, and the applicable tax laws. A typical capital recovery factor for a 30 year life based on typical assumptions regarding financing and the federal Tax Reform Act of 1986 is 0.1034 (see EPRI, 1986). The variable cost levelization factor converts the annual costs into an equivalent levelized annual cost. If inflation and

escalation are assumed to be zero, the variable cost levelization factor has a value of 1.0. The capital recovery and variable cost levelization factors are typically calculated using the standard method described by EPRI (1986).

8.0 SUMMARY AND CONCLUSIONS

This chapter summarizes the cost models developed in this report. The models have been implemented as computer programs for integration with existing METC performance models of the selected IGCC systems. The limitations and applicability of the models are discussed, emphasizing that these models are intended to estimate costs *only* for the specific IGCC systems described in Sections 2, 3, and 4. Priorities for further cost model development are discussed. Finally, this section ends with a discussion of future work and conclusions.

8.1 Summary of Cost Models

This report documents the development of new cost models for selected IGCC systems. The cost models include detailed estimation of both capital and operating costs, based on accepted electric utility industry practice. These models were developed for a baseline "conventional" KRW-based IGCC system with cold gas cleanup and two alternative systems with advanced hot gas cleanup technology (one KRW-based and the other Lurgi-based). The cost models are intended for use with existing performance models, developed by DOE/METC in the ASPEN simulation environment, of these three IGCC systems. The cost models provide a consistent basis for comparative cost analysis between baseline and advanced technology options, as well as a basis for evaluation of each technology individually.

The capital cost models consist of two parts: a series of correlations for the direct cost of major plant sections; and a generic framework for estimating indirect and other capital costs. The direct cost correlations are based on key plant performance parameters. The parameters used in the direct cost correlations for the three selected IGCC technologies are summarized in Table 18. The total direct costs for each of the IGCC systems are estimated based on approximately 20 plant performance or design parameters for 10 to 12 major plant sections (excluding general facilities). The direct cost models therefore represent an intermediate level of detail, in contrast to either a detailed estimate for an actual plant (involving hundreds of separate equipment items and an equivalent number of performance and design parameters), or a simplified estimate in which the total direct capital cost is estimated based on scaling of a single parameter such as plant size. The direct cost models provide sufficient detail to evaluate the effect of changes in key performance and design parameters on the total capital cost. The direct cost models can also be modified as new or more detailed information on the cost of key plant sections becomes available.

The total capital cost for an IGCC plant is estimated based on the direct costs for each plant section and a number of other capital cost items. A summary of the key parameters in

Table 18. Summary of Parameters in the Direct Capital Cost Models

Description	Area No.	Direct Cost Model Parameter(s) ^{a,b}		
		KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Oxidant Feed	10	Ambient temperature Oxygen flow rate Oxidant oxygen purity	Compressor work Precooler heat transfer area	Compressor work Precooler heat transfer area
Coal Handling	20	Gasifier coal feed	Gasifier coal feed	AR coal feed
Limestone Handling	25	--	Limestone feed rate	--
Gasification	30	Oxidant feed Coal feed (MAF) Ash removal rate	Coal feed (MAF)	Number of units (based on coal throughput)
Coke, Fines, and Ash	31	--	--	Gasifier coal feed
Cyclones	32	--	--	Syngas pressure Syngas volume flow rate
Sulfation	35	--	Ash and fines removal rate	--
Gas Cooling	40	Syngas outlet temp. Syngas outlet pres. Syngas flow rate	--	--
Selexol	50	Syngas flow rate H ₂ S removal eff.	--	--
Zinc Ferrite	50	--	Sorbent loading Sulfur flow rate Sorbent bulk density Syngas volume flow Design pressure	Sorbent loading Sulfur flow rate Sorbent bulk density Syngas volume flow Design pressure
Claus Plant	60	Recovered sulfur flow rate	--	--
Sulfuric Acid Plant	60	--	--	Off-gas flow rate Concentration of sulfur dioxide Off-gas temp

(Continued on next page)

Table 18 (Continued). Summary of Parameters in the Direct Capital Cost Models

Description	Area No.	Direct Cost Model Parameter(s) ^{a,b}		
		KRW with Cold Gas Cleanup	KRW with Hot Gas Cleanup	Lurgi with Hot Gas Cleanup
Tail Gas	70			
SCOT Process		Recovered sulfur flow rate from Claus	--	--
Beavon-Stretford		Recovered sulfur flow rate from B-S	--	--
Boiler Feed Water	80	Raw water feed rate Polished water flow rate	Raw water feed rate Polished water flow rate	Raw water feed rate Polished water flow rate
Process Cond. Treat.	85	Particulate scrubber blowdown	--	--
Combined Cycle Gas Turbine	90			
	91	Number of units (power output)	Number of units (power output)	Number of units (power output)
HRSG	92	Pressure of high pressure steam High pressure steam flow rate	Pressure of high pressure steam High pressure steam flow rate	Pressure of high pressure steam High pressure steam flow rate
Steam Turbine	93	Power output	Power output	Power output
General Facilities	100	Fraction of other direct costs	Fraction of other direct costs	Fraction of other direct costs

^a Abbreviations are: temp. = temperature; pres. = pressure; eff. = efficiency; AR = as-received; and B-S = Beavon-Stretford; MAR = moisture- and ash-free.

^b The direct cost models are presented in detail for the three systems in Chapters 2, 3, and 4.

the total capital cost model is given in Table 19. In addition to the parameters required for the direct cost model, the total capital cost model requires specification of approximately 60 parameters, including contingency factors, and the various factors associated with the preproduction and initial catalyst and chemical costs. These include maintenance cost factors for each plant section, number of operating personnel for each plant section, and unit costs for labor, fuel, consumables, ash disposal, and byproduct credits.

The operating costs are estimated based on approximately 40 to 50 cost parameters (depending on the type of plant), which also influence the preproduction and initial catalyst and chemicals costs, as previously indicated. In addition, the material requirements for catalyst and chemicals, fuel, consumables, ash disposal, and byproduct sales are estimated based on key plant performance parameters. A summary of these parameters is given in Table 20. Therefore, the combined capital and operating cost models are based on approximately 100 performance, design, and cost parameters for each IGCC technology.

The total capital and operating costs are used in conjunction with an estimate of the net electric power production of the plant to estimate the cost of electricity produced by the plant. To accurately determine the net plant electrical output, estimates of the auxiliary power requirements of key plant sections were developed. The cost of electricity therefore depends on the parameters summarized in Tables 18, 19, and 20, plus an additional set of parameters, as discussed in Section 7. These additional parameters include the economic assumptions in the capital recovery factor and the performance parameters used to estimate the auxiliary power requirements. The auxiliary power requirements are discussed in Section 6.

In traditional deterministic cost estimates, the risk that a new technology may incur unforeseen costs is represented only in the capital cost estimates through process contingency factors. Typically, these factors are used to develop costs for a mature "fifth-of-a-kind" plant, as discussed in Section 5.1. However, in a probabilistic cost estimate, process contingency factors would not be used; instead, uncertainties in specific performance and cost parameters that contribute to technical and cost risks would be characterized explicitly and quantified using probability distributions. Furthermore, the characterization of uncertainty would not be limited to the capital cost estimate; uncertainty in operating costs can also be explicitly characterized. As indicated previously, the capital and operating cost models are based on approximately 100 parameters, each of which may be uncertain, especially for a new technology that has not been built on a commercial scale. Probabilistic analysis permits an evaluation of the simultaneous effects of uncertainty in multiple model parameters on key results such as the total capital cost, annual operating cost, and the cost of electricity. In a later

Table 19. Summary of Key Parameters in the Total Capital Cost Model

Indirect capital cost factor (default = 0.25)
Sales tax (default = 0.06)
Engineering and home office cost factor (default = 0.15)
Cost of environmental permits (default = \$1 million)
Process contingency factors (Table 12)
Project contingency factor (default = 0.20)
Interest for funds spent during construction (default = 0.10)
Inflation rate (default = 0, constant dollar basis)
Preproduction costs
 Depend on all parameters that affect operating costs
Initial catalyst and chemicals
 Depend on material requirements and unit costs for each material

Table 20. Summary of Key Parameters in the Operating Cost Model

Average labor rate (default = \$19.70/hour)
Shift factor (default = 4.75)
Number of operators per shift for each process area (Table 15)
Maintenance cost factor for each process area (Table 16)
Unit costs of fuel, consumables, ash disposal, and byproduct credits (Table 17)
Mass requirements for fuel, consumables, ash disposal, and byproducts

phase of the work described here, uncertainty in key performance and cost parameters used in the cost models will be quantified in a series of case studies.

8.2 Cost Model Implementation

As emphasized throughout this report, the cost models developed here are specifically intended to be coupled with an existing set of IGCC performance models developed by METC in the ASPEN simulation environment. The cost models for the IGCC systems have been coded into Fortran and implemented as subroutines along with the corresponding ASPEN performance models. The performance models determine the key material flow rates and process parameters required by the cost model to calculate capital and annual costs. Each IGCC cost model subroutine consists of a main program in which capital and annual costs are estimated based on the equations developed in the previous chapters. The cost model subroutine calls additional subroutines as needed for: (a) estimating the number of trains for selected process areas, (b) range-checking the predictive variables for the direct cost regression models, (c) estimating the fixed charge factor, and (d) estimating the variable cost levelization factor. These last two subroutines are required to estimate the total revenue requirement and the cost of electricity. Each subroutine contains extensive comments to document the code.

To implement the cost models with the ASPEN performance models, ASPEN Fortran blocks were developed to access the flowsheet performance variables required as inputs to the cost models. The Fortran blocks call the cost model subroutine and transfer the values for the performance variables via common blocks. The cost model subroutines record the values of the input variables and cost results to the simulation report file in a detailed summary format.

The new computer codes developed in this study are listed in Appendices A through E. Appendix A contains the computer code specific to the oxygen-blown KRW-based IGCC system with cold gas cleanup, including: (1) a listing of the new Fortran block used to access the cost model input performance variables in the ASPEN performance model; (2) a listing of the the new cost model subroutine used to calculate total and annual costs, and the cost of electricity; and (3) a sample output from an ASPEN simulation of the integrated performance and cost models . Appendix B contains the Fortran block, cost model subroutine, and sample output for the air-blown KRW-based IGCC system with hot gas cleanup. Appendix C contains the Fortran block, cost model subroutine, and sample output for the air-blown Lurgi-based IGCC system with hot gas cleanup.

A subroutine used to estimate the number of trains of equipment for many of the process area models in the oxygen-blown KRW-based system with cold gas cleanup is listed in Appendix D. Two subroutines were developed for range checking: one prints a warning when a variable exceeds the range of values used to develop the direct cost regression models. The other sets the value of a predictive variable to the upper or lower limit of the range if it is beyond the bounds of the values used to develop the regression model. A warning is printed when a predictive variable is extrapolated. The value of a variable is reset when extrapolation is not allowed and a warning is printed. These range-checking subroutines are listed in Appendix E. Finally, the subroutines used to estimate the fixed charge and variable cost levelization factors are listed in Appendix F. These subroutines were adapted to Fortran from BASIC programs listed in the EPRI Technical Assessment Guide (1986). These subroutines are required to estimate the cost of electricity.

8.3 Applicability and Limitations

The cost models developed here are intended to estimate the cost of the specific IGCC systems detailed in Sections 2, 3, and 4. *These models are not intended for application to any other type of system.* For example, these models should not be used to estimate costs for coal gasification fuel cell systems, coal-to-SNG systems, or general coal gasification refinery systems. The cost models for specific process areas should not be used to estimate costs for a process environment different from the specific IGCC systems described in this report. For example, the Claus plant cost model should not be used to estimate sulfur recovery costs for petroleum refining. The design basis for each IGCC system and each specific process area should be carefully reviewed when considering application of these models to estimating costs for anything other than the specific systems described here.

Because the available data used to develop the cost models are limited in the range of coal types and stream composition, use of these models for other than the base case designs using Illinois No. 6 coal must be carefully considered. While the data used to develop the cost model for the KRW-based IGCC system with cold gas cleanup includes a variety of coals (Illinois No. 6, Pittsburgh No. 8, Wyodak subbituminous, North Dakota lignite, and Texas lignite), the data used to develop the other two IGCC system cost models are based only on eastern coals (e.g., Illinois No. 6 or Pittsburgh No. 8). Thus, the range of applicability of the gasification section cost models for the two systems with hot gas cleanup should be limited to eastern coals.

The ASPEN performance models and most of the performance and cost studies are based on the GE Frame 7F gas turbine. As described in Section 2.10, this is a large (150 MW)

high firing temperature gas turbine. The cost of the HRSG process area, discussed in Section 2.11, is based on large gas turbines. Therefore, these models are limited to a combined cycle configuration that includes a large gas turbine such as the GE Frame 7F.

As discussed in Section 1.7 and throughout Sections 2, 3, and 4, many of the direct cost models have been developed using regression analysis. *In general, the regression models should not be extrapolated beyond the range of the data used to develop the models.* The cases where limited extrapolation may be reasonable are noted. The limits on the range of values for the predictive variables have implications for estimating the number of trains of process equipment in each process area. As noted throughout the text, the preferred alternative to extrapolating the direct cost models is to adjust the number of trains, where possible, so that the mass flow rate per train of the key predictive variable is within the bounds of the regression model. This approach to estimating the number of trains is used in the computer models.

The computer models are designed to print warning messages when extrapolation of a direct cost model occurs. When extrapolation occurs for a direct cost model that constitutes a small portion of the total direct cost, there generally will not be a significant effect on the accuracy of the total result. However, if a model is extrapolated significantly beyond its range of validity (say, 50 percent or more above the upper limit or below the lower limit), the results may be incorrect. In such cases, the user may wish to override the assumptions regarding the number of trains in the particular process area, or take other appropriate corrective action.

In the process of developing the direct cost models, the predicted values for direct cost have been compared with published values for the purpose of validating each process area model. The models developed to estimate the consumption rate of catalysts and chemicals have been similarly compared to published estimates. The statistical measures of R^2 and standard error given in the report for each model indicate the degree to which the models replicate the data of published studies. In general, there is excellent statistical agreement, leading to good replication of published results for identical input assumptions.

8.4 Directions for Cost Model Refinement

As discussed in Section 1.7, many of the direct cost models were developed based on regression analysis of performance and cost data taken from approximately 30 studies. With few exceptions, these regression models pass standard tests for significance and acceptability. The purpose of this section is to review the relative importance of the direct cost models and to identify priorities for further data collection or cost model development.

The research needs for the direct cost models are prioritized based on two considerations. The first consideration is the relative contribution of the direct cost for each process area to the total direct cost of the plant. Process areas which represent a significant portion of the total direct cost are candidates for the development of more detailed or robust cost models. The second consideration is the relative similarity or uniqueness of each process area when comparing one IGCC system to another. Because one of the key applications of the cost models developed here is for comparative analyses of alternative IGCC systems, process areas which differ between systems will be most important in distinguishing the cost of one system from another. For example, the cost of a cold gas cleanup system will differ from the cost of a hot gas cleanup system, while the cost of a heat recovery steam generator will be similar for IGCC systems with either hot or cold gas cleanup. Thus, it is more important to develop accurate cost models for the gas cleanup systems than for the heat recovery steam generators when comparing alternative systems.

To assess the relative importance of each process area cost to the total direct cost of each IGCC system, the contributions of the process area costs for each system have been ranked in Table 21 based on the sample cost model results in Appendices A, B, and C. The general facilities cost is given at the bottom of the list even though it may constitute a more significant portion of the total direct cost than many other process areas.

The results in Table 21 indicate that the gasification and gas turbine process areas are the two largest contributors to total direct cost for all three IGCC systems, accounting for approximately half the total direct cost. The oxidant feed section is a major contributor to the direct cost for the oxygen-blown system, but a small contributor for the two air-blown systems. The combined cycle system costs, including the gas turbines, steam turbine and heat recovery steam generators, contribute approximately one-third of the total direct cost for the three systems. The zinc ferrite process cost for the Lurgi-based system comprises about 18 percent of the total direct cost.

For the oxygen-blown KRW system with cold gas cleanup, the rankings of the cost models would imply that future cost model development should focus on the gasification, gas turbine, and oxidant feed process areas. Models for the KRW system with hot gas cleanup should focus on the gas turbine, gasification, and steam turbine process areas, while for the Lurgi system the focus would be on the gasification, gas turbine, and zinc ferrite systems. However, as previously discussed, prioritization of further model develop should also focus on those systems which are unique to a particular process. For example, the direct costs (in

Table 21. Illustrative Ranking of Process Areas Contributions to Total Direct Cost^a

KRW/Cold Gas Cleanup		KRW/Hot Gas Cleanup		Lurgi/Hot Gas Cleanup	
Process Area	% of TDC	Process Area	% of TDC	Process Area	% of TDC
Gasification	25.83	Gas Turbine	24.69	Gasification	26.32
Gas Turbine	18.22	Gasification	23.28	Gas Turbine	21.42
Oxidant Feed	11.63	Steam Turbine	10.53	Zinc Ferrite	17.67
Steam Turbine	7.86	HRSG	7.05	Steam Turbine	7.17
HRSG	5.63	Coal Handling	6.81	HRSG	6.03
Coal Handling	5.34	Sulfation	6.62	Coal Handling	4.68
Gas Cooling	4.71	Zinc Ferrite	3.65	Sulfuric Acid Plant	1.64
Selexol	3.13	Oxidant Feed	2.13	Oxidant Feed	1.04
Tail Gas	1.79	Boiler Feed Water	1.21	Boiler Feed Water	0.99
Claus Plant	1.38	Limestone Handling	0.98		
Boiler Feed Water	0.89				
Process Cond. Treat.	0.55				
General Facilities	13.04	General Facilities	13.04	General Facilities	13.04
Total Direct Cost	100.00		100.00		100.00

^a KRW/Cold Gas Cleanup = oxygen-blown KRW-based IGCC system with cold gas cleanup; KRW/Hot Gas Cleanup = air-blown KRW-based IGCC system with hot gas cleanup; Lurgi/Hot Gas Cleanup = air-blown Lurgi-based IGCC system with hot gas cleanup. All results are for illustrative case studies with Illinois No. 6 coal, and nominal 650 MW power plants. The detailed cost model outputs are in Appendix A.3, B.3, and C.3 for the respective systems.

dollars) of the gas turbine, heat recovery steam generator, and steam turbine are similar for all three systems. The research needs for specific cost models are elaborated below.

Gasification Section

The gasification model for the KRW system with cold gas cleanup is based on studies with a variety of assumptions regarding coal type and mass flow rates, as discussed in Section 2. To the extent that the model is used to estimate costs for systems similar to the conceptual design studies from which the data were taken, the model should provide reliable results. However, the current model is not sensitive to parameters such as gasifier pressure, which influences, for example, the coal throughput per gasifier and the gasifier vessel wall thickness. A more general model would have to be based on disaggregation of the gasification area direct costs to specific process equipment. However, insufficient data were available at this time to develop such a detailed model.

The gasification area cost model for the air-blown KRW system with hot gas cleanup is the least detailed of the three gasification cost models and should be a priority for further cost model development. The model was developed based on data for an oxygen-blown system with high temperature gas cooling using heat exchangers. However, the performance model to which the cost model is coupled assumes air-blown operation with raw gas quenching. Several assumptions were made to adjust the costs to represent an air-blown system with raw gas quench. An additional limitation of the model is that the data are only for Pittsburgh No. 8 coal. The development of a more detailed gasification model for the air-blown system is a complementary task to developing a more detailed model for the oxygen-blown system discussed above.

The gasification process area models for the Lurgi-based system are relatively robust, with the exception of the model used to estimate the coke, ash, and fines handling equipment cost. The current model can be applied only with Illinois No. 6 coal. If other coals are assumed, the model user should determine the coal throughput requirement and supercede the default value. The coal throughput per gasifier is used to determine the number of trains of equipment. Because the basic unit of this system is the Lurgi Mark IV gasifier, which is commercially available, the costs developed using this model are expected to be within the accuracy of preliminary cost estimates, if the model is properly applied.

Gas Turbine Section

The cost of the gas turbines for all systems are based on a unit cost for a single gas turbine. While this cost is believed to be representative of the cost for a GE Frame 7F gas turbine modified to fire medium-BTU coal gas from oxygen-blown systems, it may not be an

accurate cost for a gas turbine modified to fire low-BTU coal gas from air-blown systems, nor for a gas turbine with a high fuel inlet temperature. Thus, the current gas turbine cost model may not capture possible cost differences for applications in oxygen-blown systems with cold gas cleanup vis-a-vis air-blown systems with hot gas cleanup. The development of more detailed gas turbine costs should be considered for a future study.

Oxygen Plant Section

The oxygen plant costs developed for the oxygen-blown KRW system are believed to be reliable if the model is applied within the range of parameter values used to develop the regression model. Therefore, additional development effort for this model may not be warranted. However, as noted in Section 2, for high purity oxygen plants there is an opportunity to produce argon as a saleable byproduct, which may improve the economics of oxygen-blown systems. The potential economic advantage of argon byproduct recovery may be limited by the size of the argon gas market. Therefore, further investigation is required to determine whether the development of a byproduct air separation plant model is warranted.

Steam Turbine Section

The steam turbine is the third or fourth most costly process area for the three IGCC systems. The cost model for this process area is a simple linear model based on reheat steam turbine power output. The ASPEN flowsheets for all three IGCC systems assume the same reheat steam turbine design, and the only difference between the simulations is expected to be in terms of the steam flow rates and corresponding electrical power production. Therefore, the simple cost model suffices to differentiate costs on the basis of power output. However, the current model should not be used in comparative studies if the steam turbine design basis in the flowsheet for one system is different than another.

Zinc Ferrite Section

The zinc ferrite process area is the third mostly costly area for the Lurgi-based system, although it is only the seventh most costly area for the KRW-based system with hot gas cleanup. The design basis for this model is a fixed bed system. The current model is sensitive to the most significant parameters that determine zinc ferrite costs, including the syngas volume flow rate, inlet syngas hydrogen sulfide and carbonyl sulfide flow rates, design pressure, rail shippable vessel diameter, superficial gas velocity, and sorbent characteristics (e.g., long term sulfur loading, sorbent bulk density). This model is therefore fairly robust, and is suitable for estimating the costs of both the KRW-based and Lurgi-based systems with hot gas cleanup, which feature significantly different inlet sulfur loadings. However, GE has developed an alternative moving bed zinc ferrite system, which is claimed to have lower costs than the fixed bed system. The development of a cost model for a moving bed zinc ferrite

system is recommended, assuming the availability of data from which to develop or validate such a model.

Heat Recovery Steam Generator Section

The heat recovery steam generator is the fourth or fifth most costly process area for the three IGCC systems. The cost model for this process area is based on both the high pressure superheated steam flow rate to the steam turbine and the pressure level of this steam. The ASPEN flowsheets for all three IGCC systems assume the same heat recovery steam generator design, and the only difference between the simulations is expected to be the steam flow rates. Therefore, the cost model suffices to differentiate costs on the basis of steam flow rate. A more detailed model of the heat recovery steam generator is needed for comparative studies only if the design basis in the ASPEN flowsheet for one system is changed to differ from that of another system.

Coal Handling Section

The coal handling system costs are of minor interest when evaluating advanced processes. Coal handling systems are commercially available. The models developed here are simple and will differentiate the cost of alternative IGCC systems based on the coal flow rate. Thus, the models are expected to provide reasonable results for coal handling costs when comparing the two KRW-based systems. In developing the Lurgi coal handling cost model, there was considerable attention to the comparative costs with the KRW systems. The comparison of coal handling costs between the KRW and Lurgi systems is expected to reflect the differences between these systems. For a complete discussion of these differences, see Section 4.

Sulfation Section

The sulfation process area for the KRW-based system with hot gas cleanup poses both technical and cost risks for commercialization of this technology. Because very little data for this system were available (i.e., three data points), only a simple cost model was developed. This process area merits further investigation in terms of both environmental performance and cost.

Other Process Areas

While no other single process area contributes more than five percent of the total cost of any of the IGCC systems, the combined costs of the Selexol, Claus, and Beavon-Stretford process areas constitute over six percent of the direct cost for the KRW system with cold gas cleanup. These models are believed to provide good estimates of process costs. In particular, the Claus and Beavon-Stretford models almost exactly reproduce the costs reported in the

literature. As discussed in Section 2, this does not necessarily imply that these models are "correct," because the reported costs may have been developed using simple cost scaling assumptions. However, it also implies that the models are no worse than the approach used in the various cost studies. While the Selexol process cost model does not capture all of the variation in the data, it does reasonably encode the relationship between cost and syngas flow rate and hydrogen sulfide removal efficiency. Therefore, these cold gas cleanup system models are expected to provide good results for evaluation of the KRW-based system with cold gas cleanup and comparison of this system with the alternative systems with hot gas cleanup.

The low temperature gas cooling model for the system with cold gas cleanup closely reproduces the data used to develop the model, and it captures many of the key relationships between performance and cost. The boiler feed water treatment process area is common to all systems and represents only about one percent of the total direct cost. The model is an excellent fit to the data reported in the cost studies. Therefore, no further development of this model is required. Both the process condensate treatment and limestone handling cost models were developed based on very limited data. However, these process areas contribute less than one percent of the total direct cost. If data become available to improve these models, the models should be redeveloped.

The development of all cost models is limited, of course, by the availability of data. For the purpose of developing systematic cost models, estimates prepared for DOE were generally less useful than studies prepared for EPRI or GRI, mainly because DOE studies did not employ a consistent costing method, or report costs with a consistent level of detail. Thus, it is recommended that DOE develop cost reporting guidelines to increase the usefulness of future studies and allow for comparisons with other work. It is also important to report in detail the design basis used to develop performance and cost estimates.

8.5 Future Work

The cost models developed here can be used for traditional deterministic cost analysis, in which capital and operating costs are expressed as point estimates. However, an important aspect of developing credible cost estimates which accurately reflect the state of technological development is the characterization and quantification of uncertainty in key process and cost parameters. This task will be addressed in a later phase of this research in conjunction with existing DOE/METC process performance models and the newly developed stochastic modeling capability for the ASPEN simulator.

The cost models provide a means for disaggregated consideration of the uncertainty in specific performance, capital cost, and operating cost parameters, about which engineers are best able to make judgments. This approach contrasts with the traditional contingency factor approach, which is used to represent uncertainties only in capital costs, based on an expectation of unforeseen costs. Contingency factors implicitly lump technical and cost risks into a single factor, thereby obscuring the detailed basis (if any) for the uncertainty estimate. By disaggregating judgments about uncertainties in specific model parameters, and representing them by probability distributions, the most significant parameters affecting uncertainty in total costs can be identified using probabilistic analysis techniques. This also aids in prioritization for further research and evaluation.

In a later phase of this research, expert judgments regarding uncertainties in key performance and cost parameters will be elicited and incorporated into the performance and cost models for three IGCC systems. The performance and cost models can then be exercised to answer a number of questions. Of principal interest is the total cost of each power plant alternative. Typically, this is expressed as the cost of electricity, and is often the basis for comparing competing technologies. The types of questions that can be answered by probabilistic performance and cost models of IGCC power plants include:

- What is the uncertainty in the total capital and operating cost for each power plant alternative?
- Which uncertain model parameters most affect uncertainty in overall cost?
- How likely is one plant design to be better than another plant design, as measured by specific attributes (e.g., thermal efficiency, cost of electricity)?
- What is the likely pay-off from further research and development focused on specific parts of the process technology?

These questions reflect three types of analyses. One type of analysis is a characterization of the performance and cost of a single technology, which is appropriate for answering the first two questions. This type of analysis has important implications for identifying research priorities. For example, by determining the uncertain model parameters which contribute most to the uncertainty in a measure of total cost, research can be focused to reduce uncertainty in that particular parameter.

A second type of analysis is a direct comparison of alternative plant designs, in which probability distributions for the *differences* in particular attributes are estimated. A probability

distribution for the difference in the cost of electricity between two plants, for example, provides a quantitative estimate of the probability that one technology will be less expensive than the other. It also provides an indication of how much less expensive it will be. This type of analysis can be helpful in selecting the type of IGCC systems that should be considered for further evaluations.

A third type of analysis, implied by the last question above, is a comparison of the uncertainty in performance and cost of a plant based on existing information, compared to better information that would result from specific research programs. This type of analysis can be used to estimate the value of research.

8.6 Conclusions

Typically, the emphasis in comparative evaluation of power plant designs early in the research phase is on measures of performance, with less attention devoted to the development of reliable cost estimates. The cost models developed in this work provide a consistent basis for the evaluation of selected IGCC systems, with a level of detail appropriate for evaluation of technologies in the research phase. The cost models permit answers to questions regarding the economic feasibility of proposed systems of research interest to DOE/METC and, therefore, can be a valuable tool for identifying research priorities. Because the cost models for the three alternative IGCC systems included in this study were developed using a consistent modeling approach, comparative analysis of these systems can be performed.

The collection of direct capital cost and operating cost models presented here also provides a foundation for the development of cost models of other IGCC systems. For example, a Texaco-based IGCC system with cold gas cleanup could be modeled using most of the capital and operating cost correlations developed for the KRW-based IGCC system with cold gas cleanup. Therefore, the development time for cost models of additional IGCC systems is considerably less than the time required to develop the first set of cost models.

The new models developed in this study, however, are intended for application only to the specific IGCC systems described in this report. They are not intended to estimate the costs of any other type of IGCC system, or of any other type of coal gasification system. While all of the models can be used to estimate costs for systems using Illinois No 6 coal, the models are limited in their applicability to other coal types.

Future model development efforts should be prioritized to improve the robustness of the KRW gasification process area models and the gas turbine model. In addition, a cost model of a moving bed zinc ferrite system should be developed. DOE also should develop cost guidelines for contractors to encourage consistent reporting of design base and cost information at a level of detail that can assist the development of future cost models.

The cost models developed in this study have been implemented in Fortran for use with a set of existing IGCC performance models developed by METC. The model results have been compared to published studies of similar systems. These comparisons demonstrate that the new cost models provide reasonable estimates of capital and operating costs. However, as with most models, the results depend strongly on the input assumptions provided by the user.

With these constraints, the cost models can provide a basis for systematic economic evaluation of advanced IGCC systems of particular interest to DOE/METC. When applied in conjunction with a probabilistic modeling capability, these also models can be used to evaluate specific research options based on an explicit characterization of uncertainties.

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APPENDIX A: COST MODEL FOR KRW-BASED IGCC SYSTEM WITH COLD GAS CLEANUP

The cost model of a KRW-based IGCC system with cold gas cleanup is implemented in Fortran as a subroutine. This subroutine estimates direct capital costs based on the equations developed in Section 2 of the report, and estimates the total capital cost and annual costs based on the approach documented in Section 5 and 6, respectively. The values of the performance variables required for the cost model are taken from the ASPEN flowsheet of a KRW-based IGCC system with cold gas cleanup developed by Stone (1985) and modified by Craig (1988). 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 a KRW-based IGCC system with cold gas cleanup. The code includes the ASPEN Fortran block OKCCOST for inclusion in the ASPEN simulation model of the IGCC system and a subroutine USROKCC 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.

A.1 ASPEN Fortran Block OKCCOST

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.

FORTTRAN OKCCOST

```
;
;   Author:  H.C. Frey, Carnegie-Mellon University
;   Date:    January 29, 1990
;   Description: This block accesses the key variables required to
;   estimate capital and annual costs, and then calls subroutine
;   USROKCC (USer Oxygen-blown Krw-based igcc with Cold gas cleanup
;   Cost model).
;
F   REAL*8    CLCOMP, BSCMP1, BSCMP2, P1530, P565,
F   1   P415, P150, P65, P21, GTC1, GTC2, GTC3, GTT1,
F   2   GTT2, ST415, ST115, ST21, ST1, MCFGI, TA, MOXGI,
F   3   MOGI, PERH2O, PERASH, MSNLTO, TSNLTO, PSNLTO, MH2SSI, MSYNSI,
F   4   MH2SSO, MSCO, MSBSO, MRW, MPW1, MPW2, MPW3, MSBD,
F   5   PSHRO, MSHRO, HVCOAL, WWATER, MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT,
F   6   MNH3OUT, MCO2OT2
F   COMMON/USRCST1/ CLCOMP, BSCMP1, BSCMP2, P1530, P565,
```

```

F 1 P415, P150, P65, P21, GTC1, GTC2, GTC3, GTT1,
F 2 GTT2, ST415, ST115, ST21, ST1, MCFGI, TA, MOXGI,
F 3 MOGI, PERH2O, PERASH, MSNLTO, TSNLTO, PSNLTO, MH2SSI, MSYNSI,
F 4 MH2SSO, MSCO, MSBSO, MRW, MPW1, MPW2, MPW3, MSBD,
F 5 PHSHRO, MSHSHRO, HVCOAL, WWATER, MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT,
F 6 MNH3OUT, MCO2OT2
F COMMON /USRHOC/ HOC
DESC 'Sample Cost Model Inputs and Call Cost Subroutine'
IN-UNITS ENG POWER=WATT
DEFINE CLCOMP STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WCLCOMP
DEFINE BSCMP1 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WBSCOMP1
DEFINE BSCMP2 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WBSCOMP2
DEFINE P1530 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WP1530
DEFINE P565 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WP565
DEFINE P415 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WP415
DEFINE P150 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WP150
DEFINE P65 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WP65
DEFINE P21 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WP21
DEFINE GTC1 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WGT-C1
DEFINE GTC2 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WGT-C2
DEFINE GTC3 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WGT-C3
DEFINE GTT1 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WGT-T1
DEFINE GTT2 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WGT-T2
DEFINE ST415 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WT415
DEFINE ST115 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WT115
DEFINE ST21 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WT21
DEFINE ST1 STRM-ATTR-VAR ATTR=WORK VAR=POWER STREAM=WT1
DEFINE MCFGI SUBSTREAM-VAR STREAM=COALFEED SUBS=NC VAR=MASSFLOW
DEFINE TA SUBSTREAM-VAR STREAM=GTAIR SUBS=MIXED VAR=TEMP
DEFINE MOXGI SUBSTREAM-VAR STREAM=OXYGEN SUBS=MIXED VAR=MOLEFLOW
DEFINE MOGI MOLE-FLOW STREAM=OXYGEN SUBS=MIXED COMPONENT=O2
DEFINE PERH2O COMP-ATTR-VAR STREAM=COALFEED SUBS=NC COMP=COAL &
ATTR=PROXANAL ELEMENT=1
DEFINE PERASH COMP-ATTR-VAR STREAM=COALFEED SUBS=NC COMP=COAL &
ATTR=PROXANAL ELEMENT=4
DEFINE MSNLTO STREAM-VAR STREAM=TOSELEX VAR=MOLEFLOW
DEFINE TSNLTO STREAM-VAR STREAM=TOSELEX VAR=TEMP
DEFINE PSNLTO STREAM-VAR STREAM=TOSELEX VAR=PRES
DEFINE MH2SSI MOLE-FLOW STREAM=TOSELEX SUBS=MIXED COMP=H2S
DEFINE MSYNSI SUBSTREAM-VAR STREAM=TOSELEX SUBS=MIXED VAR=MOLEFLOW
DEFINE MH2SSO MOLE-FLOW STREAM=CLEANGAS SUBS=MIXED COMP=H2S
DEFINE MSCO MASS-FLOW STREAM=CLAUSULF SUBS=CISOLID COMP=SULFUR
DEFINE MSBSO MASS-FLOW STREAM=BSSULF SUBS=CISOLID COMP=SULFUR
DEFINE MRW SUBSTREAM-VAR STREAM=MAKEUP SUBS=MIXED VAR=MASSFLOW
DEFINE MPW1 SUBSTREAM-VAR STREAM=WATER1 SUBS=MIXED VAR=MASSFLOW
DEFINE MPW2 SUBSTREAM-VAR STREAM=E115COND SUBS=MIXED VAR=MASSFLOW
DEFINE MPW3 SUBSTREAM-VAR STREAM=NH3RCOND SUBS=MIXED VAR=MASSFLOW
DEFINE MSBD SUBSTREAM-VAR STREAM=PURGEH2O SUBS=MIXED VAR=MASSFLOW
DEFINE PHSHRO STREAM-VAR STREAM=SHSTEAM VAR=PRES
DEFINE MSHSHRO SUBSTREAM-VAR STREAM=SHSTEAM SUBS=MIXED VAR=MASSFLOW
DEFINE HVCOAL COMP-ATTR-VAR STREAM=COALFEED SUBS=NC COMP=COAL &
ATTR=COALMISC ELEMENT=1
DEFINE WWATER SUBSTREAM-VAR STREAM=WASTEH2O SUBS=MIXED VAR=MASSFLOW
;
; Environmental species of concern
;
DEFINE MCO2OUT MOLE-FLOW STREAM=STACKGAS SUBS=MIXED COMP=CO2
DEFINE MCOOUT MOLE-FLOW STREAM=STACKGAS SUBS=MIXED COMP=CO

```

```

DEFINE MSO2OUT MOLE-FLOW      STREAM=STACKGAS SUBS=MIXED  COMP=SO2
DEFINE MCOSOUT MOLE-FLOW      STREAM=STACKGAS SUBS=MIXED  COMP=COS
DEFINE MNH3OUT MOLE-FLOW      STREAM=NH3OUT   SUBS=MIXED  COMP=NH3
DEFINE MCO2OT2 MOLE-FLOW      STREAM=OFFGAS   SUBS=MIXED  COMP=CO2
;
F      CALL USROKCC
;
EXECUTE AFTER BSSEP;  (BSSEP IS THE LAST BLOCK IN THE SIMULATION)
;

```


A.2 Cost Model Subroutine USROKCC

The cost model subroutine calculates the capital and annual costs for the KRW-based IGCC system with cold gas cleanup discussed in Section 2. This subroutine calls subroutines NTRAINS, USRWCK, USRRCK, USRFCF, and USRVCLF. Subroutine NTRAINS is listed in Appendix D, and is used to calculate the number of trains for certain process areas in USROKCC. USRWCK and USRRCK are used for range-checking and are listed in Appendix E. USRFCF and USRVCLF are used to calculate the fixed charge and variable cost levelization factors required to estimate the cost of electricity. These are listed in Appendix F.

```

SUBROUTINE USROKCC
IMPLICIT DOUBLE PRECISION (A-H,O-Z)

C
C Revision 1.00: H.C. Frey                               Date: March 24, 1990
C Short Title: Cost Model for KRW-IGCC With Cold Gas Cleanup
C Description: Capital and annual cost model of KRW-based integrated
C gasification combined cycle system with cold gas cleanup. Based on
C ASPEN input file developed by K.R. Stone (1985) and modified by K.R.
C Craig (1988). Cost model documented in Frey and Rubin (1989).
C
C The cost basis for the model is January 1989 (Chemical Engineering
C Plant Cost Index = 351.5).
C
C This subroutine requires the use of ENGLISH units in ASPEN.
C
REAL*8  CLCOMP, BSCMP1, BSCMP2, P1530, P565,
1  P415, P150, P65, P21, GTC1, GTC2, GTC3, GTT1,
2  GTT2, ST415, ST115, ST21, ST1, MCFGI, TA, MOXGI,
3  MOGI, PERH2O, PERASH, MSNLTO, TSNLTO, PSNLTO, MH2SSI, MSYNSI,
4  MH2SSO, MSCO, MSBSO, MRW, MPW1, MPW2, MPW3, MSBD,
5  PHSHRO, MSHSHRO, HVCOAL, WWATER, MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT,
6  MNH3OUT, MCO2OT2
REAL*8  MSABFI, MSHBFI, MSPBFI, MHYBFI, MMOBFI, MSABFPI, MSHBFPI,
1  MCWI, MLMCWI, MSOCWI, MSACWI, MCICWI, MSUCWI, MCLCWI, MBIOCWI,
2  MSSSI, MCATCI, MCATBSI, MFOGTI, MADSGFI, MWATER, MLPGI, MSAC,
3  IC, MNAOH, MCOALIN, MH2OIN, MASHOUT, MWH2OOUT
REAL*8  INTRST, MASH, MCMGI, MPW, NSHIFTS, NOPERS, MSULFR,
1  DCCH, DCOF, DCG, DCLT, DCS, DCC, DCBS, DCBF, DCPC, DCGT,
2  DCHR, DCST, DCGF
REAL*8  PWRNET, GENEFF

C
C Common blocks for input performance variables:
C   USRCST1 passes simulation variables accessed in block OKCCOST
C   USER passes variables from the ASPEN simulation, including NRPT
C   USROK1 passes cost model parameters
C   USROK2 passes capital cost uncertainty factors
C   USROK3 passes maintenance cost factors
C   USROK4 passes regression model error terms
C   USROK5 passes miscellaneous other uncertainties, including unit
C   costs.
C
COMMON /USRCST1/ CLCOMP, BSCMP1, BSCMP2, P1530, P565,

```

```

1   P415,   P150,   P65,   P21,   GTC1,   GTC2,   GTC3,   GTT1,
2   GTT2,   ST415,   ST115,   ST21,   ST1,   MCFG1,   TA,   MOXGI,
3   MOGI, PERH2O, PERASH, MSNLTO, TSNLTO, PSNLTO, MH2SSI, MSYNSI,
4   MH2SSO, MSCO, MSBSO, MRW, MPW1, MPW2, MPW3, MSBD,
5   PHSHRO, MSHSHRO, HVCOAL, WWATER, MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT,
6   MNH3OUT, MCO2OT2
COMMON /USER/ RMISS, IMISS, NGBAL, IPASS, IRESTR, ICONVG, LMSG,
1   LPMMSG, KFLAG, NHSTRY, NRPT, NTRMNL
COMMON /USRRPT/ NREPT
COMMON /USROK1/ CF, EREAL, FBM, FEHO, FGF, FICC, FPJ,
1   INTRST, EINF, NOYEARS, IBKLIF, RTAX, ALABOR,
2   NSHIFTS
COMMON /USROK2/ FPCCH, FPCOF, FPCG, FPCLT, FPCS, FPCC, FPCBS,
1   FPCBF, FPCPC, FPCGT, FPCHR, FPCST, FPCGF
COMMON /USROK3/ FMCOF, FMCCH, FMCG, FMCLT, FMCS, FMCC, FMCBS,
1   FMCBF, FMCPC, FMCCC, FMCGF
COMMON /USROK4/ ERRCH, ERROF, ERRG, ERRLT, ERRS, ERRC, ERRBS,
1   ERRCF, ERRCPC, ERRCR, ERRST, ERRCWCH, ERRCWF, ERRCWG,
2   ERRCWT, ERRCWS
COMMON /USROK5/ UCLMST, UPACID, UPSULF, UCASHD
COMMON /USRHOC/ HOC
NREPT=NRPT

```

C
C
C

CONVERT UNITS FROM BTU/HR TO WATTS

```

CLCOMP=0.292875*CLCOMP
BSCMP1=0.292875*BSCMP1
BSCMP2=0.292875*BSCMP2
P1530=0.292875*P1530
P565=0.292875*P565
P415=0.292875*P415
P150=0.292875*P150
P65=0.292875*P65
P21=0.292875*P21
GTT1=0.292875*GTT1
GTT2=0.292875*GTT2
GTC1=0.292875*GTC1
GTC2=0.292875*GTC2
GTC3=0.292875*GTC3
ST415=0.292875*ST415
ST115=0.292875*ST115
ST21=0.292875*ST21
ST1=0.292875*ST1
HOC=0.000430210*HOC

```

C

XO=0.0

C

C

--- Print out the input performance parameters -----

C

```

WRITE (NRPT,100)
100 FORMAT(30X,'PERFORMANCE SUMMARY'
& /11X,
& 'Oxygen Blown KRW-Based IGCC System with Cold Gas Cleanup'/)
WRITE (NRPT,110) MCFG1,TA,MOXGI,MOGI,PERH2O,PERASH
110 FORMAT('COST MODEL INPUT PERFORMANCE PARAMETERS ',
& '-----',/
& 10X,' Description Value',/
& 10X,' -----',/

```

```

& 10X,'Mass flow of coal to gasifier',F14.0,' lb/hr',/
1 10X,'      Ambient temperature',F14.0,' F',/
2 10X,' Oxidant feedrate to gasifier',F14.2,' lbmole/hr',/
3 10X,'      Oxygen flow to gasifier',F14.2,' lbmole/hr',/
4 10X,'      Percent moisture in coal',F14.2,' percent',/
5 10X,'      Percent ash in coal',F14.2,' percent')
WRITE(NRPT,120) MSNLTO,TSNLTO,PSNLTO,MH2SSI,MSYNSI
120 FORMAT(
& 10X,' Molar flow of syngas to LTGC',F14.2,' lbmole/hr',/
1 10X,'      Syngas temperature in LTGC',F14.2,' F',/
2 10X,'      Syngas pressure in LTGC',F14.2,' psia',/
3 10X,'      H2S entering Selexol unit',F14.2,' lbmole/hr',/
4 10X,' Syngas entering Selexol unit',F14.2,' lbmole/hr')
WRITE(NRPT,130) MH2SSO,MSCO,MSBSO,MRW,MPW1,MPW2
130 FORMAT(
& 10X,'Molar flow of H2S out of Selx',F14.2,' lbmole/hr',/
1 9X,'Mass flow of sulfur from Claus',F14.2,' lb/hr',/
2 10X,' Mass flow of sulfur from B/S',F14.2,' lb/hr',/
3 10X,'      Mass flow of raw water',F14.2,' lb/hr',/
4 10X,'Mass flow of polished water 1',F14.2,' lb/hr',/
5 10X,'Mass flow of polished water 2',F14.2,' lb/hr')
WRITE(NRPT,140) MPW3,MSBD,GTT1,GTT2,GTC1,GTC2,GTC3
140 FORMAT(
1 10X,'Mass flow of polished water 3',F14.2,' lb/hr',/
2 9X,'Mass flow of Scrubber Blowdown',F14.2,' lb/hr',/
3 10X,'      Gas Turbine Power 1',E14.6,' Watts ',/
4 10X,'      Gas Turbine Power 2',E14.6,' Watts ',/
5 10X,'      Gas Turbine Compressor 1',E14.6,' Watts ',/
6 10X,'      Gas Turbine Compressor 2',E14.6,' Watts ',/
7 10X,'      Gas Turbine Compressor 3',E14.6,' Watts ')
WRITE(NRPT,150) PSHRO, MSHRO, ST415, ST115, ST21, ST1
150 FORMAT(
& 10X,' Pressure of HP steam (HRSG)',F14.2,' psia',/
1 10X,' Mass flow of HP steam (HRSG)',F14.2,' lb/hr',/
2 10X,'      Steam Turbine Power 1',E14.6,' Watts ',/
3 10X,'      Steam Turbine Power 2',E14.6,' Watts ',/
4 10X,'      Steam Turbine Power 3',E14.6,' Watts ',/
5 10X,'      Steam Turbine Power 4',E14.6,' Watts ')
WRITE(NRPT,160) HVCOAL,WWATER,P1530,P565,P415,P150,P65,P21
160 FORMAT(
& 10X,'      Heating value of coal',F14.2,' BTU/lb',/
& 10X,'      Waste water flow rate',F14.2,' lb/hr',/
2 10X,'      Steam Cycle Pump 1530 psia',E14.6,' Watts ',/
3 10X,'      Steam Cycle Pump 565 psia',E14.6,' Watts ',/
4 10X,'      Steam Cycle Pump 415 psia',E14.6,' Watts ',/
5 10X,'      Steam Cycle Pump 150 psia',E14.6,' Watts ',/
2 10X,'      Steam Cycle Pump 65 psia',E14.6,' Watts ',/
3 10X,'      Steam Cycle Pump 21 psia',E14.6,' Watts ')
WRITE(NRPT,165) MCO2OUT,MCOOUT,MSO2OUT,MCOSOUT,MNH3OUT,
& MCO2OT2,HOC
165 FORMAT(
& 10X,'      CO2 from gas turbine',F14.2,' lbmole/hr',/
& 10X,'      CO from gas turbine',F14.2,' lbmole/hr',/
& 10X,'      SO2 from gas turbine',F14.2,' lbmole/hr',/
& 10X,'      COS from gas turbine',F14.2,' lbmole/hr',/
& 10X,'      NH3 from gas scrubbing',F14.2,' lbmole/hr',/
& 10X,'      CO2 from Beavon-Stretford',F14.2,' lbmole/hr',/
& 10X,' Actual heating value of coal',F14.2,' BTU/lb')

```

```

C
C ----- INPUT VARIABLE DEFINITIONS -----
C
C   The variable notation in this subroutine is the same as in Frey and
C   Rubin (1990).
C
C   MCFG1 = Coal feed to gasifier, lb/hr
C   TA     = Ambient temperature, F
C   MOXGI  = Oxidant feed to gasifier, lbmole/hr
C   PERH2O = Percent moisture in the coal
C   PERASH = Percent ash in the coal
C   MSNLTO = Syngas flowrate exiting low temp. cooling unit, lbmole/hr
C   TSNLTO = Syngas temperature, F
C   PSNLTO = Syngas pressure, psia
C   MH2SSI = Molar flow of hydrogen sulfide into Selexol unit, lbmole/hr
C   MSYNSI = Molar flow of syngas into Selexol unit, lbmole/hr
C   MH2SSO = Molar flow of hydrogen sulfide out of Selexol unit, lbmole/hr
C   MSCO   = Mass flow of sulfur out of Claus plant, lb/hr
C   MSBSO  = Mass flow of sulfur out of Beavon-Stretford unit, lb/hr
C   MRW    = Raw water inlet flow rate, lb/hr
C   MPW1   = Polished water flowrate, lb/hr
C   MPW2   = Polished water flowrate, lb/hr
C   MPW3   = Polished water flowrate, lb/hr
C   MSBD   = Scrubber blowdown flowrate, lb/hr
C   GTT1   = Gas turbine power, first-stage turbine, W
C   GTT2   = Gas turbine power, second-stage turbine, W
C   GTC1   = Gas turbine compressor 1 power, W
C   GTC2   = Gas turbine compressor 2 power, W
C   GTC3   = Gas turbine compressor 3 power, W
C   PHSHRO = Pressure of high pressure steam from HRSG, psia
C   MSHSHRO = Mass flow of high pressure steam from HRSG, lb/hr
C   ST415  = HP Steam turbine power output, W
C   ST115  = MP Steam turbine power output, W
C   ST21   = IP Steam turbine power output, W
C   ST1    = LP Steam turbine power output, W
C   P1530  = Steam cycle pump 1530 psia power input, W
C   P565   = Steam cycle pump 565 psia power input, W
C   P415   = Steam cycle pump 415 psia power input, W
C   P150   = Steam cycle pump 150 psia power input, W
C   P65    = Steam cycle pump 65 psia power input, W
C   P21    = Steam cycle pump 21 psia power input, W
C   HVCOAL = Heating value of coal, BTU/lb
C   QCOOLW = Heat flow from condensor, BTU/hr
C   WWATER = Waste water mass flow, lb/hr
C
C ----- COST MODEL PARAMETERS -----
C
C   The following variables are cost model parameters.
C
C
C   Set the Chemical Engineering Plant Cost Index and Industrial
C   Chemicals Producer Price indicator to January of the cost
C   year NCSTYR
C
C   NCSTYR=1989
C
C   IF (NCSTYR.EQ.1989) THEN
C       CI=351.5
C       CICPPI=411.25
C   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=351.5
      CICPPI=347.4
    ELSEIF (NCSTYR.EQ.1983) THEN
      CI=315.5
      CICPPI=339.9
    ENDIF

```

C
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----- COST MODEL PARAMETER DEFINITIONS -----

```

      CF      = Annual plant capacity factor (decimal)
      EREAL   = Annual real escalation rate (decimal)
      FBM     = Fraction of byproduct sales to marketing costs (decimal)
      FEHO    = Engineering and home office fee factor (decimal)
      FGF     = General facilities cost, fraction of other directs (decimal)
      FICC    = Indirect Construction Cost factor (decimal)
      FPJ     = Project Contingency factor (decimal)
      INTRST  = Cost of capital during construction (decimal)
      EINFL   = Inflation rate (decimal)
      NOYEARS = Number of years of construction (years)
      IBKLIF  = Book life of project (years)
      RTAX    = Sales tax rate (decimal)
      ALABOR  = Average labor rate, including burdens ($/hour)
      NSHIFTS = Number of work shifts per day (including vacations)
      FCF     = Fixed charge factor (decimal)
      DISC    = Discount rate (decimal)
      VCLF    = Variable cost levelization factor (decimal)
      GENEFF  = Generator Efficiency

```

```

      CF=0.65
      EREAL=0.0
      FBM=0.10
      FEHO=0.15
      FGF=0.15
      FICC=0.25
      FPJ=0.20
      INTRST=0.05
      EINFL=0.0
      NOYEARS=4
      IBKLIF=30
      RTAX=0.06
      ALABOR=19.70
      NSHIFTS=4.75
      CALL USRFCF(IBKLIF,FCF,DISC)
      CALL USRVCLF(EINFL,EREAL,DISC,IBKLIF,VCLF)
      INTRST=DISC
      GENEFF=0.985

```

```

C ----- PROCESS CONTINGENCY FACTORS -----
C The following are process contingency factors for each process area.
C
C   FPCCH = 0.050
C   FPCOF = 0.000
C   FPCG  = 0.375
C   FPCLT = 0.000
C   FPCS  = 0.100
C   FPCC  = 0.050
C   FPCBS = 0.100
C   FPCBF = 0.000
C   FPCPC = 0.500
C   FPCGT = 0.030
C   FPCHR = 0.030
C   FPCST = 0.030
C   FPCGF = 0.100
C
C ----- REGRESSION MODEL ERROR TERMS -----
C The error terms for the direct cost regression models have "ERR" as the
C first three letters. The error terms for the auxiliary power consumption
C models have "ERRW" as the first four letters. Models with negligible
C error terms are not included.
C
C For the purpose of deterministic simulation, the error terms are set
C to the mean value of zero if the errors have a normal distribution or
C to the median value of one if the errors have a lognormal distribution.
C
C   ERRCH = 0.0
C   ERROF = 1.0
C   ERRG  = 0.0
C   ERRLT = 0.0
C   ERRS  = 0.0
C   ERRC  = 0.0
C   ERRBS = 0.0
C   ERRBF = 1.0
C   ERRPC = 0.0
C   ERRHR = 0.0
C   ERRST = 0.0
C   ERRWCH = 0.0
C   ERRWOF = 0.0
C   ERRWG  = 0.0
C   ERRWLT = 0.0
C   ERRWS  = 0.0
C
C ----- MAINTENANCE COST FACTORS -----
C Maintenance cost factors, based on plant section capital cost
C (Based on Table 16 in Frey and Rubin (1989))
C
C   FMCOF=0.02
C   FMCCCH=0.03
C   FMCG=0.045
C   FMCLT=0.03
C   FMCS=0.02
C   FMCC=0.02
C   FMCBS=0.02
C   FMCBF=0.015
C   FMCPCH=0.02

```

FMCCC=0.015
FMCGF=0.015

C
C ----- UNIT COSTS FOR FUEL, CHEMICALS, AND CATALYSTS -----
C The following are unit costs for fuels, chemicals, and catalysts
C based on data in Table 17 from Frey and Rubin (1989). The base
C year for these costs is January, 1989. The costs are adjusted to
C other years using the Chemical Engineering magazine industrial
C chemicals producers prices index.
C

UCCOAL=1.61*(CICPPI/411.25)
UCSAI=110*(CICPPI/411.25)
UCNAOH=220*(CICPPI/411.25)
UCNA2H=0.70*(CICPPI/411.25)
UCHYDR=3.20*(CICPPI/411.25)
UCMORP=1.30*(CICPPI/411.25)
UCLIME=80*(CICPPI/411.25)
UCSODA=160*(CICPPI/411.25)
UCCORI=1.90*(CICPPI/411.25)
UCSURF=1.25*(CICPPI/411.25)
UCCHLR=250*(CICPPI/411.25)
UCBIOC=3.60*(CICPPI/411.25)
UCSELS=1.80*(CICPPI/411.25)
UCCLSC=440*(CICPPI/411.25)
UCSACA=1.90*(CICPPI/411.25)
UCSCCA=230*(CICPPI/411.25)
UCSCCH=0.36*(CICPPI/411.25)
UCBSCT=170*(CICPPI/411.25)
UCZFSO=3.30*(CICPPI/411.25)
UCPLTA=2.80*(CICPPI/411.25)
UCFLRL=11.70*(CICPPI/411.25)
UCWWCH=840*(CICPPI/411.25)
UCFUEL=42*(CICPPI/411.25)
UCRAWW=0.73*(CICPPI/411.25)
UCLMST=18*(CICPPI/411.25)
UCBNTN=0.029*(CICPPI/411.25)
UCCOKE=0.021*(CICPPI/411.25)
UCASHD=10*(CICPPI/411.25)
UCSLGD=11.30*(CICPPI/411.25)

C
C ----- UNIT PRICES FOR BYPRODUCTS -----
C The following are unit PRICES for byproducts, and may differ from
C unit costs of the same commodities purchased for consumption in
C the plant
C
C UPSULF=125*(CICPPI/411.25)
C UPACID=40*(CICPPI/411.25)

C
C ----- UNIT COST OF LAND -----
C The following is the cost of land, \$/acre
C
C UCLAND=6500

C
C ----- DIRECT CAPITAL COSTS -----
C
C The direct capital costs for 13 process areas are estimated. These
C include: Coal Handling (CH) Boiler Feedwater System (BF)
C Oxidant Feed (OF) Process Condensate Treat. (PC)

```

C          Gasification (G)          Gas Turbine (GT)
C          Low Temp. Gas Cooling (LT) Heat Recovery Steam Gen. (HR)
C          Selexol (S)              Steam Turbine (ST)
C          Claus (C)                General Facilities (GF)
C          Beavon-Stretford (BS)
C
C Calculate the direct cost for coal handling (Area 20). Coal feed
C rate is lb/hr. Original regression based on tons/day. Coefficient
C adjusted accordingly.
C
NTCH=1
NOCH=1
CALL USRWCK (3.61D+5,MCFGI,1.917D+6,'MCFGI ','DCCH ')
DCCH = ((3300 + 0.04908*MCFGI)+ERRCH)*(CI/351.5)
C
C Calculate the direct cost for the oxidant feed section (Area 10).
C The default number of units is 2. Units are lbmole/hr for the oxidant,
C degrees fahrenheit for temperature, fraction for oxygen purity.
C The regression model error for oxidant feed has a lognormal
C distribution with a median of one. Range checks are included for
C the predictive variables of ambient temperature, oxygen purity,
C and oxygen mass flow rate per train. The number of trains are
C determined based on the total mass flow rate of oxygen. The minimum
C number of operating trains is two. There are no spare trains.
C
CALL NTRAINS(6.25D+2,MOGI,1.135D+4,NOOF)
IF(NOOF.LT.2) NOOF=2
NTOF=NOOF
ETAO2 = MOGI/MOXGI
CALL USRRCK (2.0D+1, TA,9.5D+1,'TA ','DCOF ')
CALL USRRCK (9.5D-1,ETAO2,9.8D-1,'ETAO2 ','DCOF ')
CALL USRWCK (6.25D+2,(MOGI/NOOF),1.135D+4,'MOGI/N','DCOF ')
DCOF = (14.35*NTOF*TA**(0.067)/(1-ETAO2)**0.073 *
& (MOGI/NOOF)**0.852*ERROF) * (CI/351.5)
C
C Direct capital cost of the Gasification Section (Area 30). The coal
C feed used in the model is on a moisture and ash free (MAF) basis in
C units of tons/day. ASPEN reports coal use in lb/hr. Ash is in units
C of tons/day, and it is assumed, as an approximation, that all of the
C ash in the coal is removed in the gasification section. The oxidant
C feed rate is in units of lbmole/hr. The number of trains for this
C process area is determined based on the coal mass flow rate and
C regression models from Frey and Rubin (1990). Warnings are printed
C if any of the predictive variables exceed the ranges used to
C develop the regression model.
C
PERH2O=PERH2O/100.
PERASH=PERASH/100.
MCMGI = MCFGI*(1-PERH2O-PERASH)*24/2000
Mash = MCFGI*PERASH*24/2000
NTG = INT(0.556+0.001049*MCMGI+0.5)
NOG = INT(0.557+0.000910*MCMGI+0.5)
IF(NOG.EQ.0) NOG=1
IF(NTG.EQ.0) NTG=1
IF(NOG.EQ.NTG) THEN
    IF(NOG.GT.3) NTG=NOG+1
ENDIF
DCG = (27700+0.0184*NTG*(Mcmgi/Nog)**0.867*(Moxgi/Nog)**0.987*

```



```

1 (Mash/Nog)**0.083 + ERRG) * (CI/351.5)
  DCG2 = (27700 + 0.0184*(NTG+1)*(Mcmgi/(Nog+1))**0.867*
1 (Moxgi/(Nog+1))**0.987*(Mash/(Nog+1))**0.083+ERRG)*(CI/351.5)
  IF (DCG2.LT.DCG).AND.((Mcmgi/(Nog+1)).GT.(8.00D+2))) THEN
    NOG=NOG+1
    NTG=NTG+1
    DCG=DCG2
  ENDIF
  CALL USRWCK ( 8.00D+2, (Mcmgi/Nog), 1.250D+3, 'MCMG/N', 'DCG  ')
  CALL USRWCK (1.450D+3, (Moxgi/Nog), 2.000D+3, 'MOXG/N', 'DCG  ')
  CALL USRWCK ( 5.0D+1, (Mash/Nog) , 2.00D+2, 'MASH/N', 'DCG  ')
C
C Direct Capital Cost of Low Temperature Gas Cooling Section (Area 40)
C Syngas conditions are based on outlet from gas cooling, or inlet to
C sulfur removal section. The temperature is in units of degrees F,
C pressure in units of psia, and molar flow in units of lbmole/hr.
C The number of operating trains are estimated based on the total syngas
C mass flow rate and the range of syngas flow rates per train used
C to develop the regression model. There are no spare trains. There
C are range checks on the predictive variables. For both the outlet
C syngas temperature and pressure, the predictive variables are not
C allowed to exceed the ranges used to develop the regression model.
C If they exceed the range, they are reset to the nearest limit and
C an error is printed. A warning is printed if the syngas flow rate
C per train exceeds the limits used to develop the regression model.
C
  CALL NTRAINS( 16.000D+3,MSNLTO, 37.200D+3,NOLT)
  NTLT=NOLT
  CALL USRRCK ( 8.5D+1,TSNLTO, 1.20D+2,'TSNLTO','DCLT  ')
  CALL USRRCK ( 3.85D+2,PSNLTO, 4.35D+2,'PSNLTO','DCLT  ')
  CALL USRWCK ( 16.000D+3, (MSNLTO/NOLT), 37.200D+3,
& 'MSN/NO','DCLT  ')
  DCLT = (8.78E-21 * NTLT * TSNLTO**2.38 * PSNLTO**6.06 *
1 (MSNLTO/NOLT)**0.79 + ERRLT) * (CI/351.5)
C
C Direct capital cost of Selexol sulfur removal (Area 50). Typically, 2
C total units, both operating, are used. The number of operating
C trains is calculated based on the syngas mass flow rate and the
C limits for syngas flow rate per train used to develop the regression
C model. There are no spare trains. A range check on the hydrogen
C sulfide removal efficiency resets this variable within the limits
C used to develop the regression model. If the syngas flow rate per
C train exceeds the limits used to develop the model, a warning is
C printed. Because most cost studies assume a minimum of two
C operating trains, the number of trains estimated here is at least
C two.
C
  CALL NTRAINS(3.000D+3,MSYNSI,67.300D+3,NOS)
  IF (NOS.LT.2) NOS=2
  NTS = NOS
  ETAH2S=(MH2SSI-MH2SSO)/MH2SSI
  CALL USRRCK ( 8.35D-1,ETAH2S, 9.97D-1,'ETAH2S','DCS  ')
  CALL USRWCK (3.000D+3, (MSYNSI/NOS), 67.300D+3, 'MSN/NO', 'DCS  ')
  DCS = (0.420*NTS/(1-ETAH2S))**0.059*(MSYNSI/NOS)**0.980 + ERRS)
1 * (CI/351.5)
C
C Direct capital cost of the Claus process (Area 60).
C The number of trains are estimated based on the recovered sulfur

```

mass flow rate and the limits of recovered sulfur mass flow rate per train used to develop the regression model. The number of total trains is the number of operating trains plus a spare train. A warning is printed if the recovered sulfur flow rate per train exceeds the limits used to develop the regression model. For a typical 500 MW IGCC plant, the size of a single train is between 5,000 lb/hr and 10,000 lb/hr of recovered sulfur.

```

CALL NTRAINS(5.0D+3,MSCO,1.0D+4,NOC)
NTC=NOC+1
CALL USRWCK (0.695D+3,(MSCO/NOC),18.100D+3,'MSCO/N','DCC  ')
DCC=(6.28*NTC*(MSCO/NOC)**0.668 + ERRRC) *(CI/351.5)
  
```

Direct capital cost of the Beavon-Stretford Process (Area 70)
 The number of trains for this process area is the same as the number of trains for the Claus plant process area. A warning message is printed if the mass flow rate per train of sulfur recovered in the Beavon-Stretford unit exceeds the range used to develop the regression model.

```

NOBS=NOC
NTBS=NTC
CALL USRWCK (0.075D+3,(MSBSO/NOBS),1.200D+3,'MSBS/N','DCBS  ')
DCBS=(57.5+66.2*NTBS*(MSBSO/NOBS)**0.645+ERRBS) *(CI/351.5)
  
```

Direct capital cost of the boiler feedwater system

Raw water is makeup water for the steam cycle
 Polished water includes steam turbine condensate and makeup water, (this is variable WATER1 in the KRW-IGCC Cold Gas flowsheet) which comprise the majority of the polished water stream flow, and condensate from miscellaneous process users such as waste water treatment. Warning messages are printed if the predictive variables exceed the ranges used to develop the regression model. The number of trains for this commercially available process area is one, with no spare. The regression model error has a lognormal distribution with a median of one.

```

NOBF=1
NTBF=1
MPW=MPW1+MPW2+MPW3
CALL USRWCK ( 24.000D+3,MRW,0.614000D+6,'MRW  ','DCBF  ')
CALL USRWCK (0.234000D+6,MPW,3.880000D+6,'MPW  ','DCBF  ')
DCBF=(0.145*MRW**0.307*MPW**0.435*ERRBF) *(CI/351.5)
  
```

Direct capital cost of process condensate treatment
 Only one train with no spares are assumed in the literature. An error is printed if the value of the predictive variable exceeds the range used to develop the regression model.

```

NOPC=1
NTPC=1
CALL USRWCK ( 52.000D+3,MSBD,0.237000D+6,'MSBD  ','DCPC  ')
DCPC=(2583+0.00213*MSBD + ERRPC) *(CI/351.5)
  
```

Direct capital cost of the gas turbines

The gross gas turbine electrical output is estimated in MW.

```

C   The number of gas turbines is estimated based on an assumption
C   of 150 MW output for each GE Frame 7F unit. The number of units
C   is rounded to the nearest integer. There are no spare gas turbines
C
WGTE=-GENEFF*(GTC1+GTC2+GTC3+GTT1+GTT2)/1000000
NOGT=INT(WGTE/150+0.50)
IF(NOGT.EQ.0) NOGT=1
NTGT=NOGT
DCGT=32000*NTGT*(CI/351.5)
C
C   Direct capital cost of heat recovery steam generator
C   The number of HRSGs is the same as the number of gas turbines.
C   If the pressure level assumed in the steam cycle exceeds the
C   limits used to develop the regression model, the cost is estimated
C   based on the nearest limit and a warning message is printed.
C   If the steam flow rate per train exceeds the limits of the model,
C   a warning message is printed and the model is extrapolated.
C
NOHR=NOGT
NTHR=NTGT
CALL USRRCK ( 0.650D+3,PHSHRO,1.545D+3,'PHSHRO','DCHR ')
CALL USRWCK ( 66.000D+3,(MHSRO/NOHR),0.640000D+6,
& 'MHSR/N','DCHR ')
DCHR=(-5364+0.00724*NTHR*PHSHRO**1.526*(MHSRO/NOHR)**0.242+ERRHR)
& *(CI/351.5)
C
C   Direct capital cost of the steam turbine
C
C   The gross electrical output of the steam turbine is estimated in MW.
C   There is only one steam turbine.
C
NOST=1
NTST=1
WSTE=-GENEFF*(ST415+ST115+ST21+ST1)/1000000
DCST=(158.7*(WSTE)+ERRST)*(CI/351.5)
C
C   Direct capital cost of general facilities
C
DCGF=FGF*(DCCH+DCOF+DCG+DCLT+DCS+DCC+DCBS+DCBF+
& DCPC+DCGT+DCHR+DCST)
C
----- CAPITAL COSTS: TOTAL PLANT INVESTMENT -----
C
C   The total plant investment is estimated in this section. The
C   total capital requirement, which includes items dependent on
C   operating costs, is not calculated until after the operating
C   costs are estimated.
C
C   Total Direct Cost (TDC)
C
TDC=DCCH+DCOF+DCG+DCLT+DCS+DCC+DCBS+DCBF+DCPC+DCGT+DCHR+DCST+DCGF
C
C   Total Indirect Cost (TIC)
C
CICC=FICC*TDC
CTAX=RTAX*(0.8*TDC+0.1*CICC)
CEHO=FEHO*(TDC+CICC+CTAX)
CEP=1000

```

```

TIC=CICC+CTAX+CEHO+CEP
C
C Process Contingency Costs
C
CPCCH = FPCCH*DCCH*(1+TIC/TDC)
CPCOF = FPCOF*DCOF*(1+TIC/TDC)
CPCG = FPCG*DCG*(1+TIC/TDC)
CPCLT = FPCLT*DCLT*(1+TIC/TDC)
CPCS = FPCS*DCS*(1+TIC/TDC)
CPCC = FPCC*DCC*(1+TIC/TDC)
CPCBS = FPCBS*DCBS*(1+TIC/TDC)
CPCBF = FPCBF*DCBF*(1+TIC/TDC)
CPCPC = FPCPC*DCPC*(1+TIC/TDC)
CPCGT = FPCGT*DCGT*(1+TIC/TDC)
CPCHR = FPCHR*DCHR*(1+TIC/TDC)
CPCST = FPCST*DCST*(1+TIC/TDC)
CPCGF = FPCGF*DCGF*(1+TIC/TDC)
C
CPC=CPCCH+CPCOF+CPCG+CPCLT+CPCS+CPCC+CPCBS+CPCBF+CPCPC+CPCGT+
& CPCHR+CPCST+CPCGF
C
C Project contingency cost
C
CPJ=FPJ*(TDC+TIC+CPC)
C
C Total Plant Cost (TPC)
C
TPC=TDC+TIC+CPC+CPJ
C
C Total plant investment (TPI)
C
Z=(1+INTRST)/(1+EREAL)
AF=(Z**NOYEARS-1)/(NOYEARS*(Z-1))
TPI=AF*TPC
AFDC=TPI-TPC
C
C ----- NET PLANT ELECTRICAL OUTPUT -----
C
C The net electrical output must be estimated 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 oxygen plant, the gasification section, low temperature gas cooling,
C Selexol, process condensate treatment, general facilities, and
C pump and compressor power consumption in the Claus, Beavon-Stretford,
C and steam cycle systems. The pump and compressor power consumption
C is estimated directly from the ASPEN simulation. The other areas
C are estimated from regression relationships taken from Frey and Rubin
C (1989).
C
C Gas Turbine Output:  WGTE          Steam Turbine Output:  WSTE
C
C The auxiliary loads have been converted to MW from Frey and Rubin (1989)
C
WCHE = 0.378 + 6.40D-6 * MCFGI + ERRWCH
WOFE = (0.9466+3.73D-4*TA+9.019D-6*TA**2)*(0.00609*MOGI + ERRWOF)
WGE = 4.76D-5*MSNLTO + ERRWG
WLTE = -5.6 + 1.08D-5*MSNLTO + 0.01419*PSNLTO + ERRWLT
WSE = 0.348 + 4.78D-4*MSYNSI**0.839 + ERRWS

```

WCE = 2.1D-5*MSCO
 WBSE = 0.0445 + 0.00112*MSBSO
 WPCE = 7.34D-7*MSBD
 WBFE = (P1530 + P565 + P415 + P150 + P65 + P21)/1000000.0
 WGFE = 1.64 + 0.0877*(WCHE+WOFE+WGE+WLTE+WSE+WCE+WBSE+WPCE+WBFE)
 WAUXE = WCHE+WOFE+WGE+WLTE+WSE+WCE+WBSE+WPCE+WBFE+WGFE
 WNETE = WGTE + WSTE - WAUXE

----- INITIAL CATALYSTS AND CHEMICALS -----

Selexol solvent

FH2S=MH2SSI/MSYNSI
 CALL USRWCK (0.050D+3,MH2SSI,0.900D+3 , 'MH2SSI', 'CHEMIS')
 CALL USRWCK (0.004D+0,FH2S , 0.013D+0, 'FH2S ', 'CHEMIS')
 CHEMIS=-25200+16.6*MH2SSI**0.935/FH2S**1.04

Claus Catalyst

CATIC=0.00503*MSCO

Beavon-Stretford Catalyst and Chemical: Chemical requirement is given
 in cost, rather than mass, units, because only cost data were
 available.

CATIBS = -1.3 + 0.641 * MSBSO
 CIBSCHEM=85.8*MSBSO*(CICPPI/337.7)

Initial startup fuel

FUELI= 80000*(WNETE/550)

Cost of initial makeup cooling water treatment chemicals

CCHICW=1000*(CICPPI/337.7)

Total cost of initial catalyst and chemicals
 (These costs are in dollars, and must be divided by 1,000 for
 conversion to thousands of dollars)

TCICC=(UCFUEL*FUELI+CCHICW+CIBSCHEM+UCBSCT*CATIBS+UCCLSC*CATIC+
 1 UCSELS*CHEMIS)/1000

----- FIXED OPERATING COSTS -----

Fixed Operating Costs: Operating Labor

NOPERS=21+2*NOG+4*NOGT
 OCL=ALABOR*2080*NSHIFTS*NOPERS

Fixed Operating Costs: Maintenance

OCMOF = FMCOF * (DCOF * (1+(TIC+CPJ)/TDC) + CPCOF)
 OCMCH = FMCCH * (DCCH * (1+(TIC+CPJ)/TDC) + CPCCH)
 OCMG = FMCG * (DCG * (1+(TIC+CPJ)/TDC) + CPCG)
 OCMLT = FMCLT * (DCLT * (1+(TIC+CPJ)/TDC) + CPCLT)
 OCMS = FMCS * (DCS * (1+(TIC+CPJ)/TDC) + CPCS)
 OCMC = FMCC * (DCC * (1+(TIC+CPJ)/TDC) + CPCC)

```

OCMBS = FMCBS * (DCBS * (1+(TIC+CPJ)/TDC) + CPCBS)
OCMBF = FMCBF * (DCBF * (1+(TIC+CPJ)/TDC) + CPCBF)
OCMPC = FMCPC * (DCPC * (1+(TIC+CPJ)/TDC) + CPCPC)
OCMCC = FMCCC * ((DCGT+DCHR+DCST) *
& (1+(TIC+CPJ)/TDC)+CPCGT+CPCHR + CPCST)
OCMGF = FMCGF * (DCGF * (1+(TIC+CPJ)/TDC) + CPCGF)

C
OCM = (OCMOF+OCMCH+OCMG+OCMLT+OCMS+OCMC+OCMBS+OCMBF+
& OCMPC+OCMCC+OCMGF)*1000
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 Cost for Coal (UCCOAL in $/MMBtu)
C
QCOAL=MCFGI*HVCOAL
OCFUEL=CF*8760*QCOAL*UCCOAL/1E+06

C
C Material requirements for consumables, and cost of consumables
C
C Raw and polished water treatment chemical requirements
C
MSABFI=CF*(47.0+0.00209*MRW)
MSHBFI=CF*(9.5+0.00042*MRW)
MSPBFI=CF*(115+0.00361*MRW)
MHYBFI=CF*(529+0.0174*MRW)
MMOBFI=CF*(420+0.0163*MRW)

C
CALL USRWCK (1.2D+6,MPW ,2.2D+6,'MPW ','MSABFP')
MSABFPI=CF*(15+5.4E-05*MPW)
MSHBFI=CF*(30+0.000107*MPW)

C
C The makeup cooling water requirement MCWI (gal/min) is estimated
C based on the plant electrical output. The ASPEN flowsheet does
C not provide an estimate of cooling water flow rate. The estimate
C for cooling water flow rate is based on analysis of data from
C several EPRI studies.
C
MCWI=9.3763*WNETE - 273.8
CALL USRWCK (0.415D+3,MCWI ,7.700D+3,'MCWI ','M'*CW')

C
C Cooling water treatment chemical requirements
C
MLMCWI=CF*(-26+0.143*MCWI)
MSOCWI=0.154*CF*MCWI
MSACWI=CF*(6+0.147*MCWI)
MCICWI=CF*(-7280+28.8*MCWI)
MSUCWI=MCICWI
MCLCWI=CF*(4.1+0.00374*MCWI)

```

```

C      MBIOCWI=CF*(4800+4.12*MCWI)
C
C      Makeup Selexol Solvent Requirement
C
C      CALL USRWCK (4.000D+3,MSYNSI ,74.500D+3,'MSYNSI','MSSSI ')
C      MSSSI=CF*(-350+1.58*MSYNSI)
C
C      Makeup Claus plant catalyst
C
C      MCATCI=0.000961*CF*MSCO
C
C      Makeup Beavon-Stretford Catalyst
C
C      MCATBSI=0.0856*CF*MSBSO
C
C      Makeup Beavon-Stretford Chemical COST (mass data not available)
C
C      CCHMBSI=170*CF*MSBSO*(CICPPI/411.25)
C
C      Other consumables: fuel oil, plant and instrument air adsorbent,
C      wastewater chemicals (COST, not mass basis), plant raw water
C      consumption, and liquified petroleum gas.
C
C      For wastewater treatment, the waste water flow rate is converted from
C      lb/hr to gpm, using 60 min/hr, 62.4 lb/ft3, and 7.48 gal/ft3.
C
C      MFOGTI=48000*CF*(WNETE/550)
C      MADSGFI=3600*CF*(WNETE/550)
C      CWW = CF * UCWWCH * WWATER * 1/(60.0 * 62.4) * 7.48
C
C      Only raw water input to the plant (process and deaerator makeup)
C      is included in water costs.
C      Cooling water is assumed to be taken from nearby body of water
C
C      MWATER=CF*8760*(MRW*1/62.4*7.48)/1000
C      MLPGI=4200*CF*(WNETE/550)
C
C      Estimate of consumable operating costs
C
C      MSAC = MSABFI + MSABFPI + MSACWI
C      OCSAI = UCSAI * MSAC
C      MNAOH = MSHBFI + MSHBFPI
C      OCNAOH = UCNAOH * MNAOH
C      OCNA2H = UCNA2H * MSPBFI
C      OCHYDR = UCHYDR * MHYBFI
C      OCMORP = UCMORP * MMOBFI
C      OCLIME = UCLIME * MLMCWI
C      OCSODA = UCSODA * MSQCWI
C      OCCORI = UCCORI * MCICWI
C      OCSURF = UCSURF * MSUCWI
C      OCCHLR = UCCHLR * MCLCWI
C      OCBIOC = UCBIOC * MBIOCWI
C      OCSELS = UCSELS * MSSSI
C      OCCLSC = UCCLSC * MCATCI
C      OCBST = UCBST * MCATBSI
C      OCFOIL = UCFUEL * MFOGTI
C      OCPLTA = UCPLTA * MADSGFI
C      OCRAWW = UCRAWW * MWATER

```

```

C      OCFLRL = UCFLRL * MLPGI
C
C      OCONNS = OCSAI + OCNAOH + OCNA2H + OCHYDR + OCMORP + OCLIME +
&          OCSODA + OCCORI + OCSURF + OCCHLR + OCBIOC + OCSELS +
&          OCCLSC + OCBSCT + CCHMBSI + OCFOIL + OCPLTA + OCRAWW +
&          CWW      + OCFLRL
C
C      Costs for ash disposal: MASH already calculated in the gasification
C      direct cost section. MASH is in units of tons/day.
C
C      OCASH = UCASHD * MASH * CF * 365.0
C
C      Cost credit for byproduct sales. Includes a factor for byproduct
C      marketing (FBM).
C
C      MSULFR=(MSCO+MSBSO)/2000.
C      OCBYP=(1-FBM)*MSULFR*CF*8760*UPSULF
C
C      TOTAL VARIABLE OPERATING COST
C
C      VOC = OCFUEL + OCONNS + OCASH - OCBYP
C
C      ----- PREPRODUCTION CAPITAL COSTS -----
C
C      PPFC = FOC/(12000)
C      PPOC = (0.083/CF)*(OCONNS+OCASH)/1000
C      PPFUEL=(0.021/CF)*OCFUEL/1000
C      PPC=PPFC+PPOC+PPFUEL+0.02*TPI
C
C      ----- Inventory Capital Costs -----
C
C      IC=(0.164/CF)*(OCFUEL+OCONNS)/1000
C
C      ----- Total Capital Requirement (TCR) -----
C
C      ALAND=-93+0.065*QCOAL/1000000
C      TCLAND=ALAND*UCLAND/1000
C      TCR=1.01*TPI+PPC+IC+TCICC+TCLAND
C
C      ----- Normalized Costs and Cost of Electricity -----
C
C      DPERKW = TCR/WNETE
C      FOCN = FOC/(WNETE*1000.)
C      VOCN = VOC/(WNETE*8760.*CF)
C      FUELN = OCFUEL/(WNETE*8760.*CF)
C      BYPN = OCBYP/(WNETE*8760.*CF)
C      VOCINC = VOCN-FUELN+BYPN
C      CELEC=(1000*FCF*TCR+VCLF*(FOC+VOC))/(WNETE*8760*CF)
C
C      Plant heat rate and efficiency
C
C      HEATRATE=HOC*MCFGI/(WNETE*1000.)
C      EFFICNCY=3414.98/HEATRATE
C
C      EMISSIONS SUMMARY
C
C      MCOALIN = MCFGI/(WNETE*1000.)
C      MH2OIN = MRW/(WNETE*1000.)

```



```

MASHOUT = MASH*2000.0/(24.0*WNETE*1000.)
MWH2OOUT= WWATER/(WNETE*1000.)
ECO2     = (MCO2OUT+MCO2OT2)*44.0/(WNETE*1000.)
ECO      = MCOOUT*28.0/(WNETE*1000.)
ESO2     = MSO2OUT*64.0/(HOC*MCFGI/1.0D+6)
ECOS     = MCOSOUT*60.0/(HOC*MCFGI/1.0D+6)
ENH3     = MNH3OUT*17.03/(WNETE*1000.)

```

```

C
C
C ----- COST SUMMARY REPORT -----
C

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```

C      WRITE(NRPT,200)
C      200 FORMAT(33X,'COST SUMMARY'
C            & /11X,
C            & 'Oxygen Blown KRW-Based IGCC System with Cold Gas Cleanup')
C

```

```

C ----- Report Cost Model Parameters -----
C

```

```

C      WRITE(NRPT,210)
C      210 FORMAT(/
C            & 'A. COST MODEL PARAMETERS -----',
C            & '-----')
C            WRITE(NRPT,220) CF,      NCSTYR, FGF, CI,      FICC, CICPPI,
C            &      RTAX,      EREAL, FEHO, INTRST, FPJ, NOYEARS,
C            &      NSHIFTS, FBM,      FCF, ALABOR, VCLF, IBKLIF
C      220 FORMAT('      Plant Capacity Factor: ',F6.2,
C            & '      Cost Year: January',I5,
C            & /'      General Facilities Factor: ',F6.2,
C            & '      Plant Cost Index: ',F6.1,
C            & /'      Indirect Construction: ',F6.2,
C            & '      Chemicals Cost Index: ',F6.1,
C            & /'      Sales Tax: ',F6.2,
C            & '      Escalation: ',F6.2,
C            & /'      Engr & Home Office Fee: ',F6.2,
C            & '      Interest: ',F6.2,
C            & /'      Project Contingency: ',F6.2,
C            & '      Years of construction: ',I6,
C            & /'      Number of Shifts: ',F6.2,
C            & '      Byproduct marketing: ',F6.2,
C            & /'      Fixed Charge Factor: ',F6.4,
C            & '      Average Labor Rate: ',F6.2,
C            & /'Variable Levelization Cost Factor: ',F6.4,
C            & '      Book Life (years): ',I6)
C

```

```

C      WRITE(NRPT,230)
C      230 FORMAT(/'B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS ',
C            & '-----',
C            & /26X'      Process      Maintenance',
C            & /26X'Plant Section      Contingency      Cost Factor',
C            & /26X'-----')
C            WRITE(NRPT,240) FPCCH, FMCCCH, FPCOF, FMCOF, FPCG, FMCG, FPCLT,
C            &      FMCLT,
C            &      FPCS, FMCS, FPCC, FMCC, FPCBS, FMCBS, FPCBF, FMCBF,
C            &      FPCPC, FMCPG, FPCGT, FMCCC, FPCHR, FMCCC, FPCST, FMCCC,
C            &      FPCGF, FMCGF
C      240 FORMAT(10X,'      Coal Handling ',10X,F6.3,10X,F6.3,
C            & /10X,'      Oxidant Feed ',10X,F6.3,10X,F6.3,
C            & /10X,'      Gasification ',10X,F6.3,10X,F6.3,

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& /10X,' Low Temperature Gas Cooling ',10X,F6.3,10X,F6.3,
& /10X,' Selexol ',10X,F6.3,10X,F6.3,
& /10X,' Claus Plant ',10X,F6.3,10X,F6.3,
& /10X,' Beavon-Stretford ',10X,F6.3,10X,F6.3,
& /10X,' Boiler Feedwater Treatment ',10X,F6.3,10X,F6.3,
& /10X,' Process Condensate Treatment ',10X,F6.3,10X,F6.3,
& /10X,' Gas Turbine ',10X,F6.3,10X,F6.3,
& /10X,'Heat Recovery Steam Generator ',10X,F6.3,10X,F6.3,
& /10X,' Steam Turbine ',10X,F6.3,10X,F6.3,
& /10X,' General Facilities ',10X,F6.3,10X,F6.3)

```

C
C
C

----- CAPITAL COSTS -----

```

WRITE(NRPT,250)
250 FORMAT('/C. DIRECT CAPITAL AND PROCESS ',
& 'CONTINGENCY COSTS ($1,000) ',
& '-----',
& /16X' Number of Units Direct Process',
& /16X'Plant Section Operating Total Capital Cost Contingency',
& /16X'-----' )
WRITE(NRPT,260) NOCH, NTCH, DCCH, CPCCH, NOOF, NTOF, DCOF, CPCOF,
& NOG, NTG, DCG, CPCG, NOLT, NTLT, DCLT, CPCLT,
& NOS, NTS, DCS, CPCS, NOC, NTC, DCC, CPCC,
& NOBS, NTBS, DCBS, CPCBS, NOBF, NTBF, DCBF, CPCBF,
& NOPC, NTPC, DCPC, CPCPC, NOGT, NTGT, DCGT, CPCGT,
& NOHR, NTHR, DCHR, CPCHR, NOST, NTST, DCST, CPCST,
& DCGF, CPCGF
260 FORMAT(
& /' Coal Handling ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Oxidant Feed ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Gasification ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Low Temperature Gas Cooling ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Selexol ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Claus Plant ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Beavon-Stretford ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Boiler Feedwater Treatment ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Process Condensate Treatment ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Gas Turbine ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'Heat Recovery Steam Generator ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Steam Turbine ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' General Facilities ',5X,' N/A',1X,' N/A',
& 3X,F10.0,3X,F10.0)
WRITE(NRPT,270) TDC, CICC, CTAX, CEHO, CEP, TIC, CPC, CPJ, TPC,
& AFDC, TPI, PPC, IC, TCICC, TCLAND, TCR
270 FORMAT('/D. TOTAL CAPITAL REQUIREMENT ($1,000) ----',
& '-----',
& /10X,' Description ',
& ,8X,'Annual Cost',
& /10X,' -----',
& ,8X,'-----',
& /10X,' Total Direct Cost ',8X,F10.0,
& /10X,' Indirect Construction Cost ',8X,F10.0,
& /10X,' Sales Tax ',8X,F10.0,
& /10X,' Engineering and Home Office Fees ',8X,F10.0,
& /10X,' Environmental Permitting ',8X,F10.0,
& /10X,' Total Indirect Costs ',8X,F10.0,
& /10X,' Total Process Contingencies ',8X,F10.0,
& /10X,' Project Contingency ',8X,F10.0,

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& /10X,'      Total Plant Cost                      ',8X,F10.0,
& /10X,'      AFDC                                ',8X,F10.0,
& /10X,'      Total Plant Investment                ',8X,F10.0,
& /10X,'      Preproduction (Startup) Costs        ',8X,F10.0,
& /10X,'      Inventory Capital                    ',8X,F10.0,
& /10X,'      Initial Catalysts and Chemicals      ',8X,F10.0,
& /10X,'      Land                                ',8X,F10.0,
& /10X,'TOTAL CAPITAL REQUIREMENT ($1,000) ----->',8X,F10.0)

C
C ----- FIXED OPERATING COSTS -----
C
      WRITE(NRPT,280) OCL, OCM, OCAS, FOC
280 FORMAT(/'E.  FIXED OPERATING COSTS ($/year) -----',
& '-----',
& /10X,'      Description                      ',15X,'Annual Cost',
& /10X,'      -----                      ',15X,'-----',
& /10X,'      Operating Labor                  ',11X,F11.0,
& /10X,'      Maintenance Costs                ',11X,F11.0,
& /10X,'      Administration and Supervision',11X,F11.0,
& /10X,'TOTAL FIXED OPERATING COST ',
& '($/year) ----->',F11.0)

C
C ----- Unit Costs for Fuel, Chemicals, and Catalysts -----
C
      WRITE(NRPT,290) UCSAI, MSAC, OCSAI,
& UCNAOH, MNAOH, OCNAOH, UCNA2H, MSPBFI, OCNA2H,
& UCHYDR, MHYBFI, OCHYDR, UCMORP, MMOBFI, OCMORP,
& UCLIME, MLMCWI, OCLIME, UCSODA, MSOCWI, OCSODA,
& UCCORI, MCICWI, OCCORI
290 FORMAT (/
& 'F.  VARIABLE OPERATING COSTS -----',
& '-----',
& /'      1.  CONSUMABLES ($/year)',
& /6X,'      ',
& '      Material                      Annual',
& /6X,'Description          Unit Cost      ',
& '      Requirement          Operating Cost',
& /6X,'-----          -----',
& '      -----          -----',
& /6X,' Sulfuric Acid: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'      NaOH: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'      Na2 HPO4: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Hydrazine: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Morpholine: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Lime: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'      Soda Ash: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'Corrosion Inh.: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0)
      WRITE(NRPT,300) UCSURF, MSUCWI, OCSURF, UCCHLR, MCLCWI, OCCHLR,
& UCBIOC, MBIOCWI, OCBIOC, UCSELS, MSSSI, OCSELS,
& UCCLSC, MCATCI, OCCLSC,
& UCSACA, XO, XO, UCSCCA, XO, XO, UCSCCH, XO, XO
300 FORMAT(
& 6X,'      Surfactant: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Chlorine: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'      Biocide: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,' Selexol Solv.: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'Claus Catalyst: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'Sul.. Acid Cat: ',F7.2,' $/liter    ',F10.1,' liter/yr    ',F14.0,

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& /6X,' SCOT Catalyst: ',F7.2,' $/ft3 ',F10.1,' ft3/yr ',F14.0,
& /6X,'SCOT Chemicals: ',F7.2,' $/ft3 ',F10.1,' ft3/yr ',F14.0)
WRITE(NRPT,310) UCBSCT, MCATBSI, OCBST, CCHMBSI,
& UCFUEL, MFOGTI, OCFOIL, UCPLTA, MADSGFI, OCPLTA,
& UCRAWW, MWATER, OCRAWW, UCWWCH, WWATER, CWW,
& UCFLRL, MLPGI, OCFLRL, OCCONS
310 FORMAT (
& 6X,' B/S Catalyst: ',F7.2,' $/ft3 ',F10.1,' ft3/yr ',F14.0,
& /6X,' B/S Chemicals: ',7X,' N/A ',10X,' N/A ',F14.0,
& /6X,' Fuel Oil: ',F7.2,' $/bbl ',F10.1,' bbl/yr ',F14.0,
& /6X,'Plant Air Ads.: ',F7.2,' $/lb ',F10.1,' lb/yr ',F14.0,
& /6X,' Raw Water: ',F7.2,' $/Kgal ',F10.1,' Kgal/yr ',F14.0,
& /6X,' Waste Water: ',F7.2,' $/gpm ww',F10.1,' lb/hr ',F14.0,
& /6X,' LPG - Flare: ',F7.2,' $/bbl ',F10.1,' bbl/yr ',F14.0,
& /6X,'TOTAL CONSUMABLES ($/year) ----->',F14.0)
WRITE(NRPT,320) UCCOAL, MCFGI, OCFUEL, UCASHD, MASH, OCASH,
& UPSULF, MSULFR, OCBYP, VOC
320 FORMAT (
& /' 2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)',
& /6X,' Coal: ',F7.2,' $/MMBtu ',F10.1,' lb/hr ',F14.0,
& /6X,' Ash Disposal: ',F7.2,' $/ton ',F10.1,' ton/day ',F14.0,
& /6X,'Byprod. Credit: ',F7.2,' $/ton ',F10.1,' ton/hr (' ,
& F10.0,')',//6X,
& 'TOTAL VARIABLE OPERATING COST ($/year) -----> '
& ,F11.0)
C
C ----- Normalized Costs and Cost of Electricity -----
C
WRITE(NRPT,330) WGTE, WCHE, WCE, WSTE, WOFE, WBSE, WAUXE, WGE,
& WPCE,WLTE, WBFE, WNETE, WSE, WGFE, DPERKW, FOCN, VOCINC,
& BYPN, FUELN, VOCN, CELEC
330 FORMAT (/
& 'G. COST OF ELECTRICITY -----',
& '-----',
& /' Power Summary (MWe) ',
& 'Auxiliary Loads (MWe)',
& /'-----',
& '-----',
& /'Gas Turbine Output ',F7.2,
& ' Coal Handling ',F6.2,' Claus ',F6.2,
& /'Steam Turbine Output ',F7.2,
& ' Oxidant Feed ',F6.2,' B/S ',F6.2,
& /'Total Auxiliary Loads ',F7.2,
& ' Gasification ',F6.2,' Proc. Cond ',F6.2,
& /'-----',
& ' Low T Cool. ',F6.2,' Steam Cycle ',F6.2,
& /'Net Electricity ',F7.2,
& ' Selexol ',F6.2,' General Fac ',F6.2,
& //27X,' Capital Cost: ',F10.2,' $/kW',
& /27X,' Fixed Operating Cost: ',F10.2,' $/(kW-yr)',
& /8X,'Incremental Variable Costs: ',F6.2,' mills/kWh',
& /8X,'Byproduct Credit: ',F6.2,' mills/kWh',
& /8X,'Fuel Cost: ',F6.2,' mills/kWh',
& /27X,'Variable Operating Cost: ',F10.2,' mills/kWh',
& /'COST OF ELECTRICITY ----->',
& F7.2,' mills/kWh'/)
WRITE(NRPT,340) HEATRATE, EFFICNCY
340 FORMAT(5X,'Heat Rate.is:',F7.0,

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& ' BTU/kWh. Efficiency is:',F7.4)
WRITE(NRPT,350) MCOALIN, MH2OIN, MASHOUT,MWH2OOUT,
& ECO2, ECO, ESO2, ECOS, ENH3
350 FORMAT(/
& 'H. ENVIRONMENTAL SUMMARY -----',
& '-----',
& /15X,'INPUTS: Coal ',F10.3,' lb/kWh',
& /15X,' Water ',F10.3,' lb/kWh',
& /15X,'OUTPUTS Ash ',F10.3,' lb/kWh',
& /15X,' Condens. ',F10.3,' lb/kWh',
& /15X,' CO2 ',F10.3,' lb/kWh',
& /15X,' CO ',F10.3,' lb/kWh',
& /15X,' SO2 ',F10.6,' lb/MMBtu',
& /15X,' COS ',F10.6,' lb/MMBtu',
& /15X,' NH3 ',F10.6,' lb/kWh'//)
RETURN
END

```

A.3 Sample Output

The following is a sample output from an ASPEN simulation of the oxygen-blown KRW-based IGCC system with cold gas cleanup 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.

PERFORMANCE SUMMARY		
Oxygen Blown KRW-Based IGCC System with Cold Gas Cleanup		
COST MODEL INPUT PERFORMANCE PARAMETERS	-----	
Description	Value	
-----	-----	
Mass flow of coal to gasifier	505682. lb/hr	
Ambient temperature	59. F	
Oxidant feedrate to gasifier	9556.80 lbmole/hr	
Oxygen flow to gasifier	9078.96 lbmole/hr	
Percent moisture in coal	12.00 percent	
Percent ash in coal	10.00 percent	
Molar flow of syngas to LTGC	40415.02 lbmole/hr	
Syngas temperature in LTGC	100.00 F	
Syngas pressure in LTGC	465.00 psia	
H2S entering Selexol unit	500.97 lbmole/hr	
Syngas entering Selexol unit	40415.02 lbmole/hr	
Molar flow of H2S out of Selx	28.96 lbmole/hr	
Mass flow of sulfur from Claus	14850.18 lb/hr	
Mass flow of sulfur from B/S	787.54 lb/hr	
Mass flow of raw water	530665.22 lb/hr	
Mass flow of polished water 1	2052589.87 lb/hr	
Mass flow of polished water 2	24370.43 lb/hr	
Mass flow of polished water 3	28494.77 lb/hr	
Mass flow of Scrubber Blowdown	146064.49 lb/hr	
Gas Turbine Power 1	-0.321804E+09 Watts	
Gas Turbine Power 2	-0.528127E+09 Watts	
Gas Turbine Compressor 1	0.169484E+09 Watts	
Gas Turbine Compressor 2	0.711544E+08 Watts	
Gas Turbine Compressor 3	0.155599E+09 Watts	
Pressure of HP steam (HRSG)	1465.00 psia	
Mass flow of HP steam (HRSG)	1457958.23 lb/hr	
Steam Turbine Power 1	-0.656835E+08 Watts	
Steam Turbine Power 2	-0.550237E+08 Watts	
Steam Turbine Power 3	-0.636999E+08 Watts	
Steam Turbine Power 4	-0.806036E+08 Watts	

Heating value of coal	12774.00 BTU/lb
Waste water flow rate	149403.82 lb/hr
Steam Cycle Pump 1530 psia	0.257140E+07 Watts
Steam Cycle Pump 565 psia	0.329344E+05 Watts
Steam Cycle Pump 415 psia	0.253146E+06 Watts
Steam Cycle Pump 150 psia	0.313233E+06 Watts
Steam Cycle Pump 65 psia	0.255991E+04 Watts
Steam Cycle Pump 21 psia	0.000000E+00 Watts
CO2 from gas turbine	24101.49 lbmole/hr
CO from gas turbine	90.03 lbmole/hr
SO2 from gas turbine	28.56 lbmole/hr
COS from gas turbine	13.89 lbmole/hr
NH3 from gas scrubbing	4.87 lbmole/hr
CO2 from Beavon-Stretford	899.59 lbmole/hr
Actual heating value of coal	11248.73 BTU/lb

COST VARIABLE RESET - Variable PSNLTO value of	465.000
in DCLT reset to the upper limit of	435.000

COST SUMMARY

Oxygen Blown KRW-Based IGCC System with Cold Gas Cleanup

A. COST MODEL PARAMETERS -----

Plant Capacity Factor:	0.65	Cost Year:	January 1989
General Facilities Factor:	0.15	Plant Cost Index:	351.5
Indirect Construction:	0.25	Chemicals Cost Index:	411.3
Sales Tax:	0.06	Escalation:	0.00
Engr & Home Office Fee:	0.15	Interest:	0.06
Project Contingency:	0.20	Years of construction:	4
Number of Shifts:	4.75	Byproduct marketing:	0.10
Fixed Charge Factor:	0.1034	Average Labor Rate:	19.70
Variable Levelization Cost Factor:	1.0000	Book Life (years):	30

B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS -----

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Oxidant Feed	0.000	0.020
Gasification	0.375	0.045
Low Temperature Gas Cooling	0.000	0.030
Selexol	0.100	0.020
Claus Plant	0.050	0.020
Beavon-Stretford	0.100	0.020
Boiler Feedwater Treatment	0.000	0.015
Process Condensate Treatment	0.500	0.020
Gas Turbine	0.030	0.015
Heat Recovery Steam Generator	0.030	0.015
Steam Turbine	0.030	0.015
General Facilities	0.100	0.015

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

Plant Section	Number of Units	Operating Total	Direct Capital Cost	Process Contingency
Coal Handling	1	1	28119.	2104.
Oxidant Feed	2	2	61274.	0.

Gasification	5	6	136089.	76362.
Low Temperature Gas Cooling	2	2	24838.	0.
Selexol	2	2	16471.	2465.
Claus Plant	2	3	7256.	543.
Beavon-Stretford	2	3	9431.	1411.
Boiler Feedwater Treatment	1	1	4672.	0.
Process Condensate Treatment	1	1	2894.	2165.
Gas Turbine	3	3	96000.	4309.
Heat Recovery Steam Generator	3	3	29638.	1330.
Steam Turbine	1	1	41426.	1860.
General Facilities	N/A	N/A	68716.	10282.

D. TOTAL CAPITAL REQUIREMENT (\$1,000) -----		Annual Cost
Description		-----
Total Direct Cost		526826.
Indirect Construction Cost		131706.
Sales Tax		26078.
Engineering and Home Office Fees		102691.
Environmental Permitting		1000.
Total Indirect Costs		261476.
Total Process Contingencies		102832.
Project Contingency		178227.
Total Plant Cost		1069360.
AFDC		102321.
Total Plant Investment		1171680.
Preproduction (Startup) Costs		29445.
Inventory Capital		15676.
Initial Catalysts and Chemicals		4966.
Land		2125.
TOTAL CAPITAL REQUIREMENT (\$1,000) ----->		1235608.

E. FIXED OPERATING COSTS (\$/year) -----		Annual Cost
Description		-----
Operating Labor		8369348.
Maintenance Costs		28236042.
Administration and Supervision		5899129.
TOTAL FIXED OPERATING COST (\$/year) ----->		42504519.

F. VARIABLE OPERATING COSTS -----			
1. CONSUMABLES (\$/year)			
Description	Unit Cost	Material Requirement	Annual Operating Cost
-----	-----	-----	-----
Sulfuric Acid:	110.00 \$/ton	1377.2 ton/yr	151492.
NaOH:	220.00 \$/ton	317.0 ton/yr	69736.
Na2 HPO4:	0.70 \$/lb	1320.0 lb/yr	924.
Hydrazine:	3.20 \$/lb	6345.7 lb/yr	20306.
Morpholine:	1.30 \$/lb	5895.4 lb/yr	7664.
Lime:	80.00 \$/ton	506.6 ton/yr	40531.
Soda Ash:	160.00 \$/ton	563.8 ton/yr	90210.
Corrosion Inh.:	1.90 \$/lb	100708.6 lb/yr	191346.
Surfactant:	1.25 \$/lb	100708.6 lb/yr	125886.
Chlorine:	250.00 \$/ton	16.4 ton/yr	4089.
Biocide:	3.60 \$/lb	18203.9 lb/yr	65534.
Selexol Solv.:	1.80 \$/lb	41278.7 lb/yr	74302.
Claus Catalyst:	440.00 \$/ton	9.3 ton/yr	4082.

Sul.. Acid Cat:	1.90 \$/liter	0.0 liter/yr	0.
SCOT Catalyst:	230.00 \$/ft3	0.0 ft3/yr	0.
SCOT Chemicals:	0.36 \$/ft3	0.0 ft3/yr	0.
B/S Catalyst:	170.00 \$/ft3	43.8 ft3/yr	7449.
B/S Chemicals:	N/A	N/A	87023.
Fuel Oil:	42.00 \$/bbl	35733.6 bbl/yr	1500811.
Plant Air Ads.:	2.80 \$/lb	2680.0 lb/yr	7504.
Raw Water:	0.73 \$/Kgal	362205.5 Kgal/yr	264410.
Waste Water:	840.00 \$/gpm ww	149403.8 lb/hr	162975.
LPG - Flare:	11.70 \$/bbl	3126.7 bbl/yr	36582.
TOTAL CONSUMABLES	(\$/year)	----->	2912855.

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

Coal:	1.61 \$/MMBtu	505681.7 lb/hr	59217154.
Ash Disposal:	10.00 \$/ton	606.8 ton/day	1439676.
Byprod. Credit:	125.00 \$/ton	7.8 ton/hr	(5008565.)

TOTAL VARIABLE OPERATING COST (\$/year) -----> 58561120.

G. COST OF ELECTRICITY -----

Power Summary (MWe)		Auxiliary Loads (MWe)			
Gas Turbine Output	446.89	Coal Handling	3.61	Claus	0.31
Steam Turbine Output	261.04	Oxidant Feed	55.29	B/S	0.93
Total Auxiliary Loads	78.00	Gasification	1.92	Proc. Cond	0.11
-----	-----	Low T Cool.	1.01	Steam Cycle	3.17
Net Electricity	629.92	Selexol	3.85	General Fac	7.80

	Capital Cost:	1961.53 \$/kW
	Fixed Operating Cost:	67.48 \$/(kW-yr)
Incremental Variable Costs:	1.21 mills/kWh	
Byproduct Credit:	1.40 mills/kWh	
Fuel Cost:	16.51 mills/kWh	
	Variable Operating Cost:	16.33 mills/kWh
COST OF ELECTRICITY	----->	63.80 mills/kWh

Heat Rate is: 9030. BTU/kWh. Efficiency is: 0.3782

H. ENVIRONMENTAL SUMMARY -----

INPUTS:	Coal	0.803 lb/kWh
	Water	0.842 lb/kWh
OUTPUTS	Ash	0.080 lb/kWh
	Condens.	0.237 lb/kWh
	CO2	1.746 lb/kWh
	CO	0.004 lb/kWh
	SO2	0.321306 lb/MMBtu
	COS	0.146473 lb/MMBtu
	NH3	0.000132 lb/kWh

APPENDIX B: COST MODEL FOR KRW-BASED IGCC SYSTEM WITH HOT GAS CLEANUP

The cost model of an air-blown KRW-based IGCC system with hot gas cleanup is implemented in Fortran as a subroutine. This subroutine estimates direct capital costs based on the equations developed in Section 3 of the report, and estimates the total capital cost and annual costs based on the approach documented in Section 5 and 6, respectively. The values of the performance variables required for the cost model are taken from the ASPEN flowsheet of a KRW-based IGCC system with hot gas cleanup developed by Craig (1988). 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 a KRW-based IGCC system with hot gas cleanup. The code includes the ASPEN Fortran block AKHCOST for inclusion in the ASPEN simulation model of the IGCC system and a subroutine USRAKHC 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.

B.1 ASPEN Fortran Block AKHCOST

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.

FORTTRAN AKHCOST

```
;
;   Author:  H.C. Frey, Carnegie-Mellon University
;   Date:    February 2, 1990
;   Description: This block accesses the key variables required to
;   estimate capital and annual costs, and then calls subroutine
;   USRAKHC (USer Air-blown Krw-based igcc with Hot gas cleanup
;   Cost model).
;
F   REAL*8          MCFGI,  SORB1,  SORB2,   WCME,  TAPC1I,  TAPC1O,
F   1  TWPC1I, TWPC1O,  QAPC1,  QW1PC1, QW2PC1, TAPC2O, TWPC2I, TWPC2O,
F   2  QAPC2, QW1PC2, QW2PC2, PERH2O, PERASH, PSNZFI, TSNZFI, MSNZFI,
F   3  MHSZFI, MCSZFI, MFINE1, MFINE2,   MRW,  MGBFWI, MCOND,  MAGTI,
F   4  PHSHRO, MSHRO,   TA,  HVCOAL, GTNETS, STNETS, STAUXE, MCO2OUT,
F   5  MCOOUT, MSO2OUT, MNH3IN
F   COMMON/USRCST2/  MCFGI,  SORB1,  SORB2,   WCME,  TAPC1I,  TAPC1O,
F   1  TWPC1I, TWPC1O,  QAPC1,  QW1PC1, QW2PC1, TAPC2O, TWPC2I, TWPC2O,
```

```

F 2 QAPC2, QW1PC2, QW2PC2, PERH2O, PERASH, PSNZFI, TSNZFI, MSNZFI,
F 3 MHSZFI, MCSZFI, MFINE1, MFINE2, MRW, MGBFWI, MCOND, MAGTI,
F 4 PSHRO, MSHRO, TA, HVCOAL, GTNETS, STNETS, STAUXE, MCO2OUT,
F 5 MCOOUT, MSO2OUT, MNH3IN
F COMMON/USRHOC/ HOC
;
DESC 'Sample Cost Model Inputs and Call Cost Subroutine'
IN-UNITS ENG
;
; VARIABLE DEFINITIONS -----
;
; MCFG1 = Mass flow of coal to gasifier, lb/hr
; SORB1 = Mass flow of limestone sorbent to gasifier (mixed), lb/hr
; SORB2 = Mass flow of limestone sorbent to gasifier (cisolid), lb/hr
; WCME = Work to boost air compressor, BTU/hr
; TAPC1I = Temperature of boost air at first precooling inlet, F
; TAPC1O = Temperature of boost air at first precooling outlet, F
; TWPC1I = Temperature of BFW at first precooling inlet, F
; TWPC1O = Temperature of BFW at first precooling outlet, F
; QAPC1 = Heat transferred from air in first precooling, BTU/hr
; QW1PC1 = Heat transferred to BFW from HRSG, BTU/hr
; QW2PC1 = Heat leaving from economizer, BTU/hr
; TAPC2O = Temperature of boost air at second precooling outlet, F
; TWPC2I = Temperature of BFW at second precooling inlet, F
; TWPC2O = Temperature of BFW at second precooling outlet, F
; QAPC2 = Heat transferred from air in second precooling, BTU/hr
; QW1PC2 = Heat transferred to BFW from HRSG, BTU/hr
; QW2PC2 = Heat leaving from economizer, BTU/hr
; PERH2O = Moisture content of coal feed, weight percent
; PERASH = Ash content of coal, weight percent
; PSNZFI = Pressure of syngas at zinc ferrite inlet, psia
; TSNZFI = Temperature of syngas at zinc ferrite inlet, F
; MSNZFI = Molar flow of syngas at zinc ferrite inlet, lbmole/hr
; MHSZFI = Molar flow of hydrogen sulfide (H2S) at zinc ferrite
; inlet, lbmole/hr
; MCSZFI = Molar flow of carbonyl sulfide (COS) at zinc ferrite
; inlet, lbmole/hr
; MFINE1 = Mass flow (NC) of ash and fines leaving gasifier, lb/hr
; MFINE2 = Mass flow (CISOLID) of ash and fines leaving gasifier, lb/hr
; MRW = Mass flow of raw water to power plant, lb/hr
; MGBFWI = Mass flow of quench BFW to gasifier outlet syngas, lb/hr
; MCOND = Mass flow of steam turbine condensation, lb/hr
; MAGTI = Mass flow of air to gas turbine inlet, lb/hr
; PSHRO = Pressure of high pressure steam at HRSG outlet, psia
; MSHRO = Mass flow of high pressure steam at HRSG outlet, lb/hr
; TA = Temperature of ambient air, F
; HVCOAL = Heating value of coal, BTU/lb
; GTNETS = Gas turbine net shaft work, BTU/hr
; STNETS = Steam turbines net shaft work, BTU/hr
; STAUXE = Steam cycle auxiliaries power consumption, BTU/hr
; MCO2OUT = CO2 molar flow rate in gas turbine exhaust, lbmole/hr
; MCOOUT = CO molar flow rate in gas turbine exhaust, lbmole/hr
; MSO2OUT = SO2 molar flow rate in gas turbine exhaust, lbmole/hr
; MNH3IN = NH3 molar flow rate into gas turbine combustor, lbmole/hr
;
; DEFINE MCFG1 SUBSTREAM-VAR STREAM=COALFEED SUBS=NC VAR=MASSFLOW
; DEFINE SORB1 SUBSTREAM-VAR STREAM=CALSORB SUBS=MIXED VAR=MASSFLOW
; DEFINE SORB2 SUBSTREAM-VAR STREAM=CALSORB SUBS=CISOLID VAR=MASSFLOW

```

```

DEFINE WCME      STRM-ATTR-VAR ATTR=WORK      VAR=POWER      STREAM=WBOOST
DEFINE TAPC1I    SUBSTREAM-VAR STREAM=TOBOOST  SUBS=MIXED      VAR=TEMP
DEFINE TAPC1O    SUBSTREAM-VAR STREAM=BOOST1    SUBS=MIXED      VAR=TEMP
DEFINE TWPC1I    SUBSTREAM-VAR STREAM=VHPBFW    SUBS=MIXED      VAR=TEMP
DEFINE TWPC1O    SUBSTREAM-VAR STREAM=HOTHBFW    SUBS=MIXED      VAR=TEMP
DEFINE QAPC1     STRM-ATTR-VAR ATTR=HEAT        VAR=Q           STREAM=QAIRHX1
DEFINE QW1PC1    STRM-ATTR-VAR ATTR=HEAT        VAR=Q           STREAM=QHRSG2
DEFINE QW2PC1    STRM-ATTR-VAR ATTR=HEAT        VAR=Q           STREAM=QLPBOIL
DEFINE TAPC2O    SUBSTREAM-VAR STREAM=BOOSTAIR  SUBS=MIXED      VAR=TEMP
DEFINE TWPC2I    SUBSTREAM-VAR STREAM=ECONBFW    SUBS=MIXED      VAR=TEMP
DEFINE TWPC2O    SUBSTREAM-VAR STREAM=HOTBFW     SUBS=MIXED      VAR=TEMP
DEFINE QAPC2     STRM-ATTR-VAR ATTR=HEAT        VAR=Q           STREAM=QAIRHX2
DEFINE QW1PC2    STRM-ATTR-VAR ATTR=HEAT        VAR=Q           STREAM=QHRSG3
DEFINE QW2PC2    STRM-ATTR-VAR ATTR=HEAT        VAR=Q           STREAM=QLPECXS
DEFINE PERH2O    COMP-ATTR-VAR STREAM=COALFEED  SUBS=NC         COMP=COAL &
                  ATTR=PROXANAL ELEMENT=1
DEFINE PERASH    COMP-ATTR-VAR STREAM=COALFEED  SUBS=NC         COMP=COAL &
                  ATTR=PROXANAL ELEMENT=4
DEFINE PSNZFI    STREAM-VAR      STREAM=RAWGAS          VAR=PRES
DEFINE TSNZFI    STREAM-VAR      STREAM=RAWGAS          VAR=TEMP
DEFINE MSNZFI    STREAM-VAR      STREAM=RAWGAS          VAR=MOLEFLOW
DEFINE MHSZFI    MOLE-FLOW       STREAM=RAWGAS          SUBS=MIXED      COMP=H2S
DEFINE MCSZFI    MOLE-FLOW       STREAM=RAWGAS          SUBS=MIXED      COMP=COS
DEFINE MFINE1    SUBSTREAM-VAR   STREAM=SOLIDS          SUBS=NC         VAR=MASSFLOW
DEFINE MFINE2    SUBSTREAM-VAR   STREAM=SOLIDS          SUBS=CISOLID    VAR=MASSFLOW
DEFINE MRW       SUBSTREAM-VAR   STREAM=MAKEUP          SUBS=MIXED      VAR=MASSFLOW
DEFINE MGBFWI    SUBSTREAM-VAR   STREAM=GASCBFW         SUBS=MIXED      VAR=MASSFLOW
DEFINE MCOND     SUBSTREAM-VAR   STREAM=WATER25         SUBS=MIXED      VAR=MASSFLOW
DEFINE MAGTI     SUBSTREAM-VAR   STREAM=GTAIR           SUBS=MIXED      VAR=MASSFLOW
DEFINE PSHRO     STREAM-VAR      STREAM=SHSTEAM         VAR=PRES
DEFINE MSHRO     SUBSTREAM-VAR   STREAM=SHSTEAM         SUBS=MIXED      VAR=MASSFLOW
DEFINE TA        SUBSTREAM-VAR   STREAM=GTAIR           SUBS=MIXED      VAR=TEMP
DEFINE HVCOAL    COMP-ATTR-VAR   STREAM=COALFEED        SUBS=NC         COMP=COAL &
                  ATTR=COALMISC ELEMENT=1
DEFINE GTNETS    STRM-ATTR-VAR   ATTR=WORK              VAR=POWER      STREAM=WGASTURB
DEFINE STNETS    STRM-ATTR-VAR   ATTR=WORK              VAR=POWER      STREAM=WSTMTURB
DEFINE STAUXE    STRM-ATTR-VAR   ATTR=WORK              VAR=POWER      STREAM=WSTMISC
;
; Collect values for an environmental summary
;
DEFINE MCO2OUT   MOLE-FLOW       STREAM=OUTGAS          SUBS=MIXED      COMP=CO2
DEFINE MCOOUT    MOLE-FLOW       STREAM=OUTGAS          SUBS=MIXED      COMP=CO
DEFINE MSO2OUT   MOLE-FLOW       STREAM=OUTGAS          SUBS=MIXED      COMP=SO2
DEFINE MNH3IN    MOLE-FLOW       STREAM=BURNED          SUBS=MIXED      COMP=NH3
;
F      CALL USRAKHC
;
; EXECUTE AFTER ST-MISC; (ST-MISC IS THE LAST BLOCK IN THE SIMULATION)
;

```

B.2 Cost Model Subroutine USRAKHC

The cost model subroutine calculates the capital and annual costs for the KRW-based IGCC system with hot gas cleanup discussed in Section 3. This subroutine calls subroutines USRWCK, USRRCK, USRFCF, and USRVCLF. USRWCK and USRRCK are used for range-checking and are listed in Appendix E. USRFCF and USRVCLF are used to calculate the fixed charge and variable cost levelization factors required to estimate the cost of electricity. These are listed in Appendix F.

```
SUBROUTINE USRAKHC
IMPLICIT DOUBLE PRECISION (A-H,O-Z)

C
C   Revision 1.00: H.C. Frey                               Date: February 25, 1990
C   Short Title: Cost Model for Air-blown KRW-IGCC With Hot Gas Cleanup
C   Description: Capital and annual cost model of KRW-based integrated
C   gasification combined cycle system with hot gas cleanup. Based on
C   ASPEN input file developed by K.R. Craig (1988).
C   Cost model documented in Frey and Rubin (1990).
C
C   The cost basis for the model is January 1989 (Chemical Engineering
C   Plant Cost Index = 351.5).
C
C   This subroutine requires the use of ENGLISH units in ASPEN.
C
C   REAL*8          MCFGI,  SORB1,  SORB2,  WCME,  TAPC1I,  TAPC1O,
1  TWPC1I, TWPC1O,  QAPC1,  QW1PC1, QW2PC1, TAPC2O, TWPC2I, TWPC2O,
2  QAPC2, QW1PC2, QW2PC2, PERH2O, PERASH, PSNZFI, TSNZFI, MSNZFI,
3  MHSZFI, MCSZFI, MFINE1, MFINE2,  MRW,  MGBFWI, MCOND,  MAGTI,
4  PHSHRO, MSHSHRO,  TA,  HVCOAL, GTNETS, STNETS, STAUXE, MCO2OUT,
5  MCOOUT, MSO2OUT, MNH3IN
C
C   REAL*8  MSABFI, MSHBFI, MSPBFI, MHYBFI, MMOBFI, MSABFPI, MSHBFPI,
1  MCWI, MLMCWI, MSOCWI, MSACWI, MCICWI, MSUCWI, MCLCWI, MBIOCWI,
2  MSSSI, MCATCI, MCATBSI, MFOGTI, MADSGFI, MWATER, MLPGI, MSAC,
3  IC, MNAOH, MLMANN, MSOZFI, MCOALIN, MH2OIN, MLMSTIN, MASHOUT
C   REAL*8  INTRST, MASH, MCMGI, MPW, NSHIFTS, NOPERS, MSULFR,
1  DCCH, DCL, DCCM, DCPC, DCOF, DCG, DCZF, DCS, DCBF, DCGT,
2  DCHR, DCST, DCGF
C   REAL*8  GENEFF
C
C   Common blocks for input performance variables:
C   USRCST2 passes simulation variables accessed in block AKHCOST
C   USER passes variables from the ASPEN simulation, including NRPT
C   USRAK1 passes cost model parameters
C   USRAK2 passes capital cost uncertainty factors
C   USRAK3 passes maintenance cost factors
C   USRAK4 passes regression model error terms
C   USRAK5 passes miscellaneous other uncertainties, including unit
costs.
C
C   COMMON/USRCST2/  MCFGI,  SORB1,  SORB2,  WCME,  TAPC1I,  TAPC1O,
1  TWPC1I, TWPC1O,  QAPC1,  QW1PC1, QW2PC1, TAPC2O, TWPC2I, TWPC2O,
2  QAPC2, QW1PC2, QW2PC2, PERH2O, PERASH, PSNZFI, TSNZFI, MSNZFI,
```

```

3  MHSZFI, MCSZFI, MFINE1, MFINE2,      MRW, MGBFWI, MCOND, MAGTI,
4  PHSHRO, MSHSHRO,      TA, HVCOAL, GTNETS, STNETS, STAUXE, MCO2OUT,
5  MCOOUT, MSO2OUT, MNH3IN
COMMON/USRHOC/ HOC
COMMON /USER/  RMISS, IMISS, NGBAL, IPASS, IRESTR, ICONVG, LMSG,
1  LPMMSG, KFLAG, NHSTRY, NRPT, NTRMNL
COMMON /USRRPT/ NREPT
COMMON /USRAK1/ CF,      EREAL, FBM,      FEHO, FGF,      FICC, FPJ,
1  INTRST,EINFL,  NOYEARS,IBKLIF,RTAX,  ALABOR,
2  NSHIFTS
COMMON /USRAK2/ FPCCH, FPCL,  FPCOF, FPCG,  FPCS,  FPCZF,
1  FPCBF, FPCGT, FPCHR, FPCST, FPCGF
COMMON/ USRAK3/ FMCOF, FMCCH, FMCL,  FMCG,  FMCS,  FMCZF,
1  FMCBF, FMCCC, FMCGF
COMMON /USRAK4/ ERRCH,  ERROF,
1  ERBBF, ERRHR,  ERRST,  ERRWCH
COMMON /USRAK5/ UCLMST, UPACID, UPSULF, UCASHD, UFG1, UFG2
1  UCZFSO, SLZF,  VSZF,  TAZF
NREPT=NRPT

```

```

C
C  CONVERT UNITS FROM BTU/HR TO WATTS
C

```

```

WCME   = 0.292875*WCME
GTNETS = 0.292875*GTNETS
STNETS = 0.292875*STNETS
STAUXE = 0.292875*STAUXE
HOC    = 0.000430210*HOC

```

```

C  XO=0.0

```

```

C  --- Print out the input performance parameters -----
C

```

```

WRITE(NRPT,100) MCFG1,SORB1, SORB2,  WCME,TAPC1I,TAPC1O,
& TWPC1I,TWPC1O
WRITE(NRPT,110) QAPC1,QW1PC1,QW2PC1,TAPC2O,TWPC2I,TWPC2O,
& QAPC2,QW1PC2
WRITE(NRPT,120) QW2PC2,PERH2O,PERASH,PSNZFI,TSNZFI,MSNZFI,
& MHSZFI,MCSZFI
WRITE(NRPT,130) MFINE1, MFINE2,  MRW, MGBFWI,MCOND, MAGTI,
& PHSHRO,MSHSHRO,TA, HVCOAL,GTNETS,STNETS,STAUXE
WRITE(NRPT,135) MCO2OUT,MCOOUT,MSO2OUT,MNH3IN,HOC

```

```

C  100 FORMAT('      *** SUMMARY OF COST MODEL INPUT ',
& 'PERFORMANCE PARAMETERS ***'//
& 'PLANT SECTION  ', 'INPUT PERFORMANCE PARAMETERS',
& 8X, 'VALUE', ' UNITS', /
& '-----', '-----', /
& 8X, '----', '----', /
& 'Coal Feed      ',
& 'Mass flow of coal to gasifier      ',F12.2,' lb/hr ',/
& 'Limestone Feed ',
& 'Mass flow of limestone (mixed)      ',F12.2,' lb/hr ',/
& 16X,'Mass flow of limestone (cisolid) ',F12.2,' lb/hr ',/
& 'Oxidant Feed   ',
& 'Work to boost air compressor      ',E12.6,' Watts',/
& 16X,'First precoolers inlet air temp. ',F12.2,' F ',/
& 16X,'First precoolers outlet air temp. ',F12.2,' F ',/
& 16X,'First precoolers inlet BFW temp. ',F12.2,' F ',/

```

```

& 16X,'First precoolers outlet BFW temp. ',F12.2,' F ')
110 FORMAT(
& 16X,'Heat transfer in first precoolers ',E12.4,' BTU/hr',/
& 16X,'Heat trans. to BFW from HRSG ',E12.4,' BTU/hr',/
& 16X,'Heat leaving from economizer ',E12.4,' BTU/hr',/
& 16X,'Second precoolers outlet air temp. ',F12.2,' F ',/
& 16X,'Second precoolers inlet BFW temp. ',F12.2,' F ',/
& 16X,'Second precoolers outlet BFW temp. ',F12.2,' F ',/
& 16X,'Heat transfer in second precoolers ',E12.4,' BTU/hr',/
& 16X,'Heat trans. to BFW from HRSG ',E12.4,' BTU/hr')
120 FORMAT(
& 16X,'Heat leaving from economizer ',E12.4,' BTU/hr',/
& 'Gasification ',
& 'Moisture in coal feed ',F12.2,' wt %',/
& 16X,'Ash in coal ',F12.2,' wt %',/
& 'Zinc Ferrite ',
& 'Zinc ferrite inlet syngas pres. ',F12.2,' psia ',/
& 16X,'Zinc ferrite inlet syngas temp. ',F12.2,' F',/
& 16X,'Zinc ferrite inlet syngas flow ',F12.2,' lbmole/hr',/
& 16X,'Zinc ferrite inlet H2S flow ',F12.2,' lbmole/hr',/
& 16X,'Zinc ferrite inlet COS flow ',F12.2,' lbmole/hr')
130 FORMAT(
& 'Sulfation ',
& 'Gasifier fines and ash (NC) ',F12.2,' lb/hr',/
& 16X,'Gasifier fines and ash (CISOLID) ',F12.2,' lb/hr',/
& 'BFW Treating ',
& 'Raw water to power plant ',F12.2,' lb/hr',/
& 16X,'Quench water to hot syngas ',F12.2,' lb/hr',/
& 16X,'Steam turbine condensate ',F12.2,' lb/hr',/
& 'Gas Turbine ',
& 'Gas turbine inlet air flow ',F12.2,' lb/hr',/
& 'Steam Cycle ',
& 'HRSG outlet HP steam pres. ',F12.2,' psia ',/
& 16X,'HRSG outlet HP steam flow ',F12.2,' lb/hr',/
& 16X,'Temp. of ambient air ',F12.2,' F',/
& 16X,'Heating value of coal ',F12.2,' BTU/lb',/
& 16X,'Gas turbine net shaft work ',E12.6,' Watts',/
& 16X,'Steam turbine net shaft work ',E12.6,' Watts',/
& 16X,'Steam cycle auxiliary power ',E12.6,' Watts')
135 FORMAT(
& 'Emissions ',
& 'CO2 emissions from gas turbine ',F12.2,' lbmole/hr ',/
& 16X,'CO emissions from gas turbine ',F12.2,' lbmole/hr',/
& 16X,'SO2 emissions from gas turbine ',F12.2,' lbmole/hr',/
& 16X,'NH3 entering gas turbine ',F12.2,' lbmole/hr',/
& 16X,'Coal Heating Value ',F12.2,' BTU/lb')

```

```

C
C ----- INPUT VARIABLE DEFINITIONS -----
C The variable notation in this subroutine is the same as in Frey and
C Rubin (1989).
C
C MCFG1 = Mass flow of coal to gasifier, lb/hr
C SORB1 = Mass flow of limestone sorbent to gasifier (mixed), lb/hr
C SORB2 = Mass flow of limestone sorbent to gasifier (cisolid), lb/hr
C WCME = Work to boost air compressor, BTU/hr
C TAPC1I = Temperature of boost air at first precoolers inlet, F
C TAPC1O = Temperature of boost air at first precoolers outlet, F
C TWPC1I = Temperature of BFW at first precoolers inlet, F

```

C TWPC10 = Temperature of BFW at first precoolers outlet, F
 C QAPC1 = Heat transferred from air in first precoolers, BTU/hr
 C QW1PC1 = Heat transferred to BFW from HRSG, BTU/hr
 C QW2PC1 = Heat leaving from economizer, BTU/hr
 C TAPC20 = Temperature of boost air at second precoolers outlet, F
 C TWPC2I = Temperature of BFW at second precoolers inlet, F
 C TWPC2O = Temperature of BFW at second precoolers outlet, F
 C QAPC2 = Heat transferred from air in second precoolers, BTU/hr
 C QW1PC2 = Heat transferred to BFW from HRSG, BTU/hr
 C QW2PC2 = Heat leaving from economizer, BTU/hr
 C PERH2O = Moisture content of coal feed, weight percent
 C PERASH = Ash content of coal, weight percent
 C PSNZFI = Pressure of syngas at zinc ferrite inlet, psia
 C TSNZFI = Temperature of syngas at zinc ferrite inlet, F
 C MSNZFI = Molar flow of syngas at zinc ferrite inlet, lbmole/hr
 C MHSZFI = Molar flow of hydrogen sulfide (H₂S) at zinc ferrite
 C inlet, lbmole/hr
 C MCSZFI = Molar flow of carbonyl sulfide (COS) at zinc ferrite
 C inlet, lbmole/hr
 C MFINE1 = Mass flow of ash and fines leaving gasifier (NC), lb/hr
 C MFINE2 = Mass flow of ash and fines leaving gasifier (CISOLID), lb/hr
 C MRW = Mass flow of raw water to power plant, lb/hr
 C MGBFWI = Mass flow of quench water to hot syngas, lb/hr
 C MCOND = Mass flow of steam turbine condensation, lb/hr
 C MAGTI = Mass flow of air to gas turbine inlet, lb/hr
 C PHSHRO = Pressure of high pressure steam at HRSG outlet, psia
 C MHSHRO = Mass flow of high pressure steam at HRSG outlet, lb/hr
 C TA = Temperature of ambient air, F
 C HVCOAL = Heating value of coal, BTU/lb
 C GTNETS = Gas turbine net shaft work, BTU/hr
 C STNETS = Steam turbines net shaft work, BTU/hr
 C STAUXE = Steam cycle auxiliaries power consumption, BTU/hr

C ----- COST MODEL PARAMETERS -----
 C The following variables are cost model parameters.

C Set the Chemical Engineering Plant Cost Index and Industrial
 C Chemicals Producer Price indicator to January of the cost
 C year NCSTYR

C NCSTYR=1989

C IF (NCSTYR.EQ.1989) THEN
 C CI=351.5
 C CICPPI=411.25
 C ELSEIF (NCSTYR.EQ.1988) THEN
 C CI=336.3
 C CICPPI=349.9
 C ELSEIF (NCSTYR.EQ.1987) THEN
 C CI=318.3
 C CICPPI=323.9
 C ELSEIF (NCSTYR.EQ.1986) THEN
 C CI=323.5
 C CICPPI=341.7
 C ELSEIF (NCSTYR.EQ.1985) THEN
 C CI=324.7
 C CICPPI=337.7

FPCGT = 0.030
 FPCHR = 0.030
 FPCST = 0.030
 FPCGF = 0.100

C
 C ----- REGRESSION MODEL ERROR TERMS -----
 C The error terms for the direct cost regression models have "ERR" as the
 C first three letters. The error terms for the auxiliary power consumption
 C models have "ERRW" as the first four letters. Models with negligible
 error
 C terms are not included.
 C

ERRCH = 0.0
 ERROF = 0.0
 ERBBF = 0.0
 ERRHR = 0.0
 ERRST = 0.0
 ERRWCH = 0.0

C
 C ----- MAINTENANCE COST FACTORS -----
 C Maintenance cost factors, based on plant section capital cost
 C (Based on Table 16 in Frey and Rubin (1989))
 C

FMCCCH = 0.02
 FMCL = 0.03
 FMCOF = 0.02
 FMCG = 0.045
 FMCS = 0.04
 FMCZF = 0.045
 FMCBF = 0.015
 FMCCC = 0.015
 FMCGF = 0.015

C
 C ----- UNIT COSTS FOR FUEL, CHEMICALS, AND CATALYSTS -----
 C The following are unit costs for fuels, chemicals, and catalysts
 C based on data in Table 17 from Frey and Rubin (1989). The base
 C year for these costs is January, 1989. The costs are adjusted to
 C other years using the Chemical Engineering magazine industrial
 C chemicals producers prices index.
 C

UCCOAL=1.61*(CICPPI/411.25)
 UCSAI=110*(CICPPI/411.25)
 UCNAOH=220*(CICPPI/411.25)
 UCNA2H=0.70*(CICPPI/411.25)
 UCHYDR=3.20*(CICPPI/411.25)
 UCMORP=1.30*(CICPPI/411.25)
 UCLIME=80*(CICPPI/411.25)
 UCSODA=160*(CICPPI/411.25)
 UCCORI=1.90*(CICPPI/411.25)
 UCSURF=1.25*(CICPPI/411.25)
 UCCHLR=250*(CICPPI/411.25)
 UCBIOC=3.60*(CICPPI/411.25)
 UCSELS=1.80*(CICPPI/411.25)
 UCCLSC=440*(CICPPI/411.25)
 UCSACA=1.90*(CICPPI/411.25)
 UCSCCA=230*(CICPPI/411.25)
 UCSCCH=0.36*(CICPPI/411.25)
 UCBSCT=170*(CICPPI/411.25)

UCZFSO=3.30*(CICPPI/411.25)
 UCPLTA=2.80*(CICPPI/411.25)
 UCFLRL=11.70*(CICPPI/411.25)
 UCWWCH=840*(CICPPI/411.25)
 UCFUEL=42*(CICPPI/411.25)
 UCRAWW=0.73*(CICPPI/411.25)
 UCLMST=18*(CICPPI/411.25)
 UCBNTN=0.029*(CICPPI/411.25)
 UCCOKE=0.021*(CICPPI/411.25)
 UCASHD=10*(CICPPI/411.25)
 UCSLGD=11.30*(CICPPI/411.25)

----- UNIT PRICES FOR BYPRODUCTS -----

The following are unit PRICES for byproducts, and may differ from unit costs of the same commodities purchased for consumption in the plant

UPSULF=125*(CICPPI/411.25)
 UPACID=40*(CICPPI/411.25)

----- UNIT COST OF LAND -----

The following is the cost of land, \$/acre

UCLAND=6500

----- DIRECT CAPITAL COSTS -----

The direct capital costs for 11 process areas are estimated. These include:

Coal Handling (CH)	Boiler Feedwater System (BF)
Limestone Handling (L)	Gas Turbine (GT)
Oxidant Feed (OF)	Heat Recovery Steam Gen. (HR)
Gasification (G)	Steam Turbine (ST)
Zinc Ferrite (ZF)	General Facilities (GF)
Sulfation (S)	

Calculate the direct cost for coal handling (Area 20). Coal feed rate is lb/hr. Original regression based on tons/day. Coefficient adjusted accordingly. The coal handling process area is considered a single train. An error is printed if the coal feed rate exceeds the range of values used to develop the regression model, although extrapolation is permitted.

NTCH=1

NOCH=1

CALL USRWCK (3.61D+5,MCFGI,1.917D+6,'MCFGI ','DCCH '

DCCH = (3300 + 0.04908*MCFGI + ERRCH)*(CI/351.5)

Calculate the direct cost for limestone handling (Area 25). Limestone feed rate MLGI is in lb/hr. The limestone handling process area is considered a single train.

NTL=1

NOL=1

MLGI = SORB1 + SORB2

DCL = (1160. + 0.026 * MLGI)*(CI/351.5)

Calculate the direct cost for the oxidant feed section (Area 10). The cost includes a boost compressor and a compressor precooler.

```

C
C First estimate the number of gas turbines based on the inlet air
C flow rate to the gas turbine compressor. The number of gas turbines
C is rounded to the nearest integer. The number of boost air
C compressors is equal to the number of gas turbines. There are no
C spare units. A warning is printed if the compressor power exceeds the
C limits used to develop the regression model. The user should
C carefully review the results if the model is extrapolated.
C
NOGT = INT(MAGTI/3000000.0+0.5)
IF(NOGT.EQ.0) NOGT=1
NTGT=NOGT

C
NOOF=NOGT
NTOF=NTGT
WCME=WCME/1000.
CALL USRWCK (1.9D+3, (WCME/NOOF), 8.2D+3, 'WCME/N', 'DCCM ')
DCCM= (-98.0 + 26. * NTOF * (WCME/NOOF)**0.485)

C
C UPC = Universal heat transfer coefficient for precoolers, Btu/(ft2-hr-F)
C DTLMP = Log mean temperature difference for precoolers, F
C APC = Heat transfer area of precoolers, ft2
C
C A warning message is printed if the heat transfer area exceeds
C the range of values used to develop the cost model for the
C precoolers.
C
UPC=40.0
TWPC1O = TWPC1I + (TWPC1O-TWPC1I)*(QAPC1/(QAPC1+QW1PC1-QW2PC1))
TLMPC1 = ((TAPC1I-TWPC1O)-(TAPC1O-TWPC1I))/
& LOG((TAPC1I-TWPC1O)/(TAPC1O-TWPC1I))
APC1= QAPC1/(UPC*TLMPC1*NOOF)
CALL USRWCK (0.0D+0, APC1, 1.0D+4, 'APC1/N', 'DCPC1 ')
DCPC1 = (24.+0.067*APC1-5.5E-6*APC1**2+2.8E-10*APC1**3)*NTOF

C
TWPC2O = TWPC2I + (TWPC2O-TWPC2I)*(QAPC2/(QAPC2+QW1PC2-QW2PC2))
TLMPC2 = ((TAPC2I-TWPC2O)-(TAPC2O-TWPC2I))/
& LOG((TAPC2I-TWPC2O)/(TAPC2O-TWPC2I))
APC2= QAPC2/(UPC*TLMPC2*NOOF)
CALL USRWCK (0.0D+0, APC2, 1.0D+4, 'APC2/N', 'DCPC2 ')
DCPC2 = (24.+0.067*APC2-5.5E-6*APC2**2+2.8E-10*APC2**3)*NTOF

C
DCOF= (DCCM + DCPC1 + DCPC2 + ERROF)*(CI/351.5)

C
C Direct capital cost of the Gasification Section (Area 30). The coal
C feed used in the model is on a moisture and ash free (MAF) basis in
C units of tons/day. ASPEN reports coal use in lb/hr. Ash is in units
C of tons/day, and it is assumed, as an approximation, that all of the
C ash in the coal is removed in the gasification section. The oxidant
C feed rate is in units of lbmole/hr.
C As an approximation, it is assumed that the number of operating
C and spare gasification trains is the same as would be estimated
C for an oxygen-blown system with the same coal feed rate.
C
C
UFG1=0.75
UFG2=1.2
PERH2O=PERH2O/100.

```

```

PERASH=PERASH/100.
MCMGI = MCFGI*(1-PERH2O-PERASH)
Mash = MCFGI*PERASH*24/2000
NTG = INT(0.556+0.001049*MCMGI*24/2000.+0.5)
NOG = INT(0.557+0.000910*MCMGI*24/2000.+0.5)
IF (NOG.EQ.0) NOG=1
IF (NTG.EQ.0) NTG=1
IF (NOG.EQ.NTG) THEN
  IF (NOG.GT.3) NTG=NOG+1
ENDIF
DCG = 16.5*MCMGI**0.68*UFG1*UFG2*(CI/351.5)

```

```

C
C Direct capital cost of zinc ferrite desulfurization
C

```

```

C This model estimates the cost of a fixed-bed system. The number of
C trains is adjusted so that each reactor vessel is rail-shippable.
C For each operating vessel, assumed to be in absorption mode, there
C is a vessel in regeneration mode. As a base case, it is assumed that
C there are two absorbing vessels per gasifier. The number of vessels
C is adjusted to conform to rail-shippable sizes and to keep the
C superficial velocity below VSZF.
C If the vessel diameter exceeds the rail-shippable size or is less
C than the minimum value required to keep the superficial velocity
C below the specified value (see variable VSZF), then a warning
C message is printed.
C

```

```

C NMZF = Number of operating vessels per operating gasifier
C DMAX = Maximum rail-shippable diameter, ft
C TAZF = Time of absorption cycle in zinc ferrite section, hours
C SLZF = Sorbent loading, long term, in zinc ferrite, wt fraction
C DBZF = Bulk density of zinc ferrite sorbent, lb/ft2
C VSZF = Superficial velocity, ft/sec
C

```

```

C NMZF = 1
C ILOOP = 0
C JLOOP = 0
C DMAX = 12.5
C TAZF = 168.0
C SLZF = 0.10
C DBZF = 82.0
C VSZF = 2.0
C MSZFI = (MHSZFI+ MCSZFI)*TAZF
C SCZF = (32-10.666667*SLZF)*MSZFI/SLZF

```

```

800 CONTINUE

```

```

IF (NMZF.LT.1) NMZF = 1
NOZF = NMZF*NOG
NTZF = 2*NMZF*NTG
VCZF = SCZF/(DBZF*NOZF)
DZF = (VCZF/3.141592654)**(1/3.0)
DZFMIN = 1.128*( 10.73*MSNZFI*(TSNZFI+460)
& / (3600.*VSZF*PSNZFI*NOZF) )**0.5
IF (DZF.LT.DZFMIN) THEN
  ILOOP = ILOOP+1
  NMZF = NMZF-1
  IF (NMZF.LE.1) THEN
    NMZF=1
    JLOOP=9
  
```

```

      ENDIF
      IF (ILOOP.LT.10) GOTO 800
    ELSEIF (DZF.GT.DMAX) THEN
      JLOOP = JLOOP+1
      NMZF = NMZF+1
      IF (JLOOP.LT.10) GOTO 800
    ENDIF
    CALL USRWCK (DZFMIN,DZF,DMAX,'DZF  ','DCZF  ')
    HZF  = (4*VCZF)/(3.141592654*DZF**2)
    AZF  = 4.1002-4.3468*DZF+0.95848*DZF**2-0.039117*DZF**3
    BZF  = (0.76181+0.5505*DZF-0.059669*DZF**2+0.00258088*DZF**3)/
&      (3.28084)
    FPIZF = 2.5 + 0.093*(PSNZFI**0.64)
    DCZF  = 1477.*FPIZF*(AZF+BZF*HZF)*NTZF*(CI/351.5)/1000.

C
C   Direct capital cost of the sulfation section
C   One operating and no spare trains are assumed.
C
NOS=1
NTS=1
DCS  = 13.0*(MFINE1+MFINE2)**0.639*(CI/351.5)

C
C   Direct capital cost of the boiler feedwater system
C
C   Raw water is makeup water for the steam cycle
C   Polished water includes steam turbine condensate and makeup water.
C   The boiler feedwater system is considered a single train.
C
NOBF=1
NTBF=1
MRW=MRW+MGBFWI
MPW=MRW+MCOND
CALL USRWCK ( 24.000D+3,MRW,0.614000D+6,'MRW  ','DCBF  ')
CALL USRWCK (0.234000D+6,MPW,3.880000D+6,'MPW  ','DCBF  ')
DCBF=(0.145*MRW**0.307*MPW**0.435+ERRBF)*(CI/351.5)

C
C   Direct capital cost of the gas turbines
C
C   The gross gas turbine electrical output is estimated in MW.
C   The number of units is estimated in the oxidant feed section,
C   because the number of boost air compressors is equal to the
C   number of gas turbines.
C
WGTE=-GENEFF*GTNETS/1000000
DCGT=32000*NTGT*(CI/351.5)

C
C   Direct capital cost of heat recovery steam generator
C   The number of total and operating units is the same as for
C   the gas turbines.
C
NOHR=NOGT
NTHR=NTGT
CALL USRWCK ( 0.650D+3,PHSHRO,1.545D+3,'PHSHRO','DCHR  ')
CALL USRWCK ( 66.000D+3,(MHSRO/NOHR),0.640000D+6,
&   'MHSR/N','DCHR  ')
DCHR=(-5364+0.00724*NTHR*PHSHRO**1.526*(MHSRO/NOHR)**0.242+ERRHR)
&   *(CI/351.5)

```

```

C   Direct capital cost of the steam turbine
C
C   The gross electrical output of the steam turbine is estimated in MW.
C   Only one steam turbine is assumed.
C
C       NOST=1
C       NTST=1
C       WSTE=-GENEFF*STNETS/1000000
C       DCST=(158.7*(WSTE)+ERRST)*(CI/351.5)
C
C   Direct capital cost of general facilities
C
C       DCGF=FGF*(DCCH+DCL+DCOF+DCG+DCZF+DCS+DCBF+DCGT+DCHR+DCST)
C
C   ----- CAPITAL COSTS:  TOTAL PLANT INVESTMENT -----
C
C   The total plant investment is estimated in this section.  The
C   total capital requirement, which includes items dependent on
C   operating costs, is not calculated until after the operating
C   costs are estimated.
C
C   Total Direct Cost (TDC)
C
C       TDC=DCCH+DCL+DCOF+DCG+DCZF+DCS+DCBF+DCGT+DCHR+DCST+DCGF
C
C   Total Indirect Cost (TIC)
C
C       CICC=FICC*TDC
C       CTAX=RTAX*(0.8*TDC+0.1*CICC)
C       CEHO=FEHO*(TDC+CICC+CTAX)
C       CEP=1000
C       TIC=CICC+CTAX+CEHO+CEP
C
C   Process Contingency Costs
C
C       CPCCH = FPCCH*DCCH*(1.+TIC/TDC)
C       CPCL  = FPCL *DCL *(1.+TIC/TDC)
C       CPCOF = FPCOF*DCOF*(1.+TIC/TDC)
C       CPCG  = FPCG *DCG *(1.+TIC/TDC)
C       CPCZF = FPCZF*DCZF*(1.+TIC/TDC)
C       CPCS  = FPCS *DCS *(1.+TIC/TDC)
C       CPCBF = FPCBF*DCBF*(1.+TIC/TDC)
C       CPCGT = FPCGT*DCGT*(1.+TIC/TDC)
C       CPCHR = FPCHR*DCHR*(1.+TIC/TDC)
C       CPCST = FPCST*DCST*(1.+TIC/TDC)
C       CPCGF = FPCGF*DCGF*(1.+TIC/TDC)
C
C       CPC=CPCCH+CPCL+CPCOF+CPCG+CPCS+CPCZF+CPCBF+CPCGT
C       & +CPCHR+CPCST+CPCGF
C
C   Project contingency cost
C
C       CPJ=FPJ*(TDC+TIC+CPC)
C
C   Total Plant Cost (TPC)
C
C       TPC=TDC+TIC+CPC+CPJ
C

```

```

C   Total plant investment (TPI)
C
C   Z=(1+INTRST)/(1+EREAL)
C   AF=(Z**NOYEARS-1)/(NOYEARS*(Z-1))
C   TPI=AF*TPC
C   AFDC=TPI-TPC
C
C   ----- NET PLANT ELECTRICAL OUTPUT -----
C
C   The net electrical output must be estimated 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   oxygen plant, the gasification section, limestone handling,
C   sulfation, general facilities, and power consumption in the
C   steam cycle. The pump and compressor power consumption
C   is estimated directly from the ASPEN simulation. The other areas
C   are estimated from regression relationships taken from Frey and Rubin
C   (1989).
C
C   Gas Turbine Output:  WGTE           Steam Turbine Output:  WSTE
C
C   The auxiliary loads have been converted to MW from Frey and Rubin (1989)
C
C   WCHE = 0.378 + 6.40D-6 * MCFGI + ERRWCH
C   WLE  = (-13.3 + 0.00461*MLGI)/1000.
C   WOFE = WCME/1000.
C   WGE  = (58. + 0.0083*MCMGI)/1000.
C   WSE  = (-3.7 + 0.0102*(MFINE1+MFINE2))/1000.
C   WBFE = STAUXE/1000000.0
C   WGFE = 1.64 + 0.0877*(WCHE+WLE+WOFE+WGE+WSE+WBFE)
C   WAUXE = WCHE+WLE+WOFE+WGE+WSE+WBFE+WGFE
C   WNETE = WGTE+WSTE-WAUXE
C
C   ----- INITIAL CATALYSTS AND CHEMICALS -----
C
C   Zinc Ferrite Sorbent: the total sorbent charge is:
C
C   TSCZF = (REAL(NTZF)/REAL(NOZF)) * SCZF
C
C   Initial startup fuel
C
C   FUELI= 80000*(WNETE/550.)
C
C   Cost of initial makeup cooling water treatment chemicals
C
C   CCHICW=1000*(CICPPI/337.7)
C
C   Total cost of initial catalyst and chemicals
C   (These costs are in dollars, and must be divided by 1,000 for
C   conversion to thousands of dollars)
C
C   TCICC=(UCFUEL*FUELI+CCHICW+UCZF*SO*TSCZF)/1000.
C
C   ----- FIXED OPERATING COSTS -----
C
C   Fixed Operating Costs: Operating Labor
C
C   NOPERS=16+2*NOG+4*NOGT

```



```

OCL=ALAB0R*2080*NSHIFTS*NOPERS
C
C Fixed Operating Costs: Maintenance
C
OCMOF = FMCOF * (DCOF * (1+(TIC+CPJ)/TDC) + CPCOF)
OCMCH = FMCH * (DCH * (1+(TIC+CPJ)/TDC) + CPCCH)
OCML = FMCL * (DCL * (1+(TIC+CPJ)/TDC) + CPCL)
OCMG = FMCG * (DCG * (1+(TIC+CPJ)/TDC) + CPCG)
OCMS = FMCS * (DCS * (1+(TIC+CPJ)/TDC) + CPCS)
OCMZ = FMZF * (DCZF * (1+(TIC+CPJ)/TDC) + CPCZF)
OCMBF = FMCBF * (DCBF * (1+(TIC+CPJ)/TDC) + CPCBF)
OCMCC = FMCCC * ((DCGT+DCHR+DCST) *
& (1+(TIC+CPJ)/TDC)+CPCGT+CPCHR + CPCST)
OCMGF = FMCGF * (DCGF * (1+(TIC+CPJ)/TDC) + PCGF)
C
OCM = (OCMOF+OCMCH+OCML+OCMG+OCMS+OCMZ+OCMBF+
& OCMCC+OCMGF)*1000
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
----- VARIABLE OPERATING COSTS -----
C
C Fuel Cost for Coal (UCCOAL in $/MMBtu)
C
QCOAL=MCFGI*HVCOAL
OCFUEL=CF*8760*QCOAL*UCCOAL/1E+06
C
C Material requirements for consumables, and cost of consumables
C
C Raw and polished water treatment chemical requirements
C
MSABFI=CF*(47.0+0.00209*MRW)
MSHBFI=CF*(9.5+0.00042*MRW)
MSPBFI=CF*(115+0.00361*MRW)
MHYBFI=CF*(529+0.0174*MRW)
MMOBFI=CF*(420+0.0163*MRW)
C
CALL USRWCK (1.2D+6,MPW ,2.2D+6,'MPW ','MSABFP')
MSABFPI=CF*(15+5.4E-05*MPW)
MSHBFI=CF*(30+0.000107*MPW)
C
C The makeup cooling water requirement MCWI (gal/min) is estimated
C based on the plant electrical output
C
MCWI=9.3763*WNETE - 273.8
CALL USRWCK (0.415D+3,MCWI ,7.700D+3,'MCWI ','M**CWI')
C
C Cooling water treatment chemical requirements
C
MLMCWI=CF*(-26+0.143*MCWI)

```

$MSOCWI = 0.154 * CF * MCWI$
 $MSACWI = CF * (6 + 0.147 * MCWI)$
 $MCICWI = CF * (-7280 + 28.8 * MCWI)$
 $MSUCWI = MCICWI$
 $MCLCWI = CF * (4.1 + 0.00374 * MCWI)$
 $MBIOCWI = CF * (4800 + 4.12 * MCWI)$

Makeup Zinc Ferrite Sorbent Requirement

$FATTZF = 0.20$
 $MSOZFI = 8760 * CF * FATTZF * SCZF / (2 * TAZF * 80)$

Annual Limestone Requirement (TON/YR)

$MLMANN = MLGI * 8760 * CF / 2000.$

Other consumables: fuel oil, plant and instrument air adsorbent, wastewater chemicals (COST, not mass basis), plant raw water consumption, and liquified petroleum gas.

For wastewater treatment, the waste water flow rate is converted from lb/hr to gpm, using 60 min/hr, 62.4 lb/ft³, and 7.48 gal/ft³.

$MFOGTI = 48000 * CF * (WNETE / 550)$
 $MADSGFI = 3600 * CF * (WNETE / 550)$
 $WWATER = 0.0$
 $CWW = CF * UCWWCH * WWATER * 1 / (60.0 * 62.4) * 7.48$

Only raw water input to the plant (process and deaerator makeup) is included in water costs.

Cooling water is assumed to be taken from nearby body of water

$MWATER = CF * 8760 * (MRW * 1 / 62.4 * 7.48) / 1000$
 $MLPGI = 4200 * CF * (WNETE / 550)$

Estimate of consumable operating costs

$MSAC = MSABFI + MSABFPI + MSACWI$
 $OCSAI = UCSAI * MSAC$
 $MNAOH = MSHBFI + MSHBFPI$
 $OCNAOH = UCNAOH * MNAOH$
 $OCNA2H = UCNA2H * MSPBFI$
 $OCHYDR = UCHYDR * MHYBFI$
 $OCMORP = UCMORP * MMOBFI$
 $OCLIME = UCLIME * MLMCWI$
 $OCSODA = UCSODA * MSOCWI$
 $OCCORI = UCCORI * MCICWI$
 $OCSURF = UCSURF * MSUCWI$
 $OCCHLR = UCCHLR * MCLCWI$
 $OCBIOC = UCBIOC * MBIOCWI$
 $OCZFSO = UCZFSO * MSOZFI$
 $OCLMST = UCLMST * MLMANN$
 $OCFOIL = UCFUEL * MFOGTI$
 $OCPLTA = UCPLTA * MADSGFI$
 $OCRAWW = UCRAWW * MWATER$
 $OCFLRL = UCFLRL * MLPGI$

$OCCONS = OCSAI + OCNAOH + OCNA2H + OCHYDR + OCMORP + OCLIME +$

```

&      OCSODA + OCCORI + OCSURF + OCCHLR + OCBIOC + OCZFSO +
&      OCLMST + OCFOIL + OCPLTA + OCRAWW +
&      CWW      + OCFLRL
C
C      Costs for ash disposal:
C
      OCBYP = 0.0
      OCASH = UCASHD * ((MFINE1+MFINE2)*24./2000.) * CF * 365.0
C
C      TOTAL VARIABLE OPERATING COST (No byproduct sales)
C
      VOC = OCFUEL + OCCONS + OCASH + OCBYP
C
C      ----- PREPRODUCTION CAPITAL COSTS -----
C
      PPFC = FOC/(12000)
      PPOC = (0.083/CF)*(OCCONS+OCASH)/1000
      PPFUEL=(0.021/CF)*OCFUEL/1000
      PPC=PPFC+PPOC+PPFUEL+0.02*TPI
C
C      ----- Inventory Capital Costs -----
C
      IC=(0.164/CF)*(OCFUEL+OCCONS)/1000
C
C      ----- Total Capital Requirement (TCR) -----
C
      ALAND=-93+0.065*QCOAL/1000000
      TCLAND=ALAND*UCLAND/1000
      TCR=1.01*TPI+PPC+IC+TCICC+TCLAND
C
C      ----- Normalized Costs and Cost of Electricity -----
C
      DPERKW = TCR/WNETE
      FOCN = FOC/(WNETE*1000.)
      VOCN = VOC/(WNETE*8760.*CF)
      FUELN = OCFUEL/(WNETE*8760.*CF)
      BYPN = OCBYP/(WNETE*8760.*CF)
      VOCINC = VOCN-FUELN+BYPN
      CELEC=(1000*FCF*TCR+VCLF*(FOC+VOC))/(WNETE*8760*CF)
C
C      Plant Heat Rate and Efficiency
C
      HEATRATE=HOC*MCFGI/(WNETE*1000.0)
      EFFICNCY=3414.98/HEATRATE
C
C      Emissions Summary
C      NOTE: The NOx emission estimate is based only on fuel-bound NOx
C      derived from ammonia in the fuel gas, and does not include other
C      fuel bound NOx or thermal NOx.
C      CO and CO2 emissions are in units of lb/kWh.
C      SO2 and NOx emissions are in units of lb/MMBtu coal.
C
      MCOALIN = MCFGI/(WNETE*1000.)
      MH2OIN = MRW/(WNETE*1000.)
      MLMSTIN = MLGI/(WNETE*1000.)
      MASHOUT = (MFINE1+MFINE2)/(WNETE*1000.)
      ECO2 = MCO2OUT*44.0/(WNETE*1000.)
      ECO = MCOOUT*28.0/(WNETE*1000.)

```

```

ESO2 = MSO2OUT*64.0/(HOC*MCFGI/1.0D+6)
ENOX = 0.7*MNH3IN*30.8/(HOC*MCFGI/1.0D+6)

```

```

----- COST SUMMARY REPORT -----

```

```

WRITE(NRPT,200)
200 FORMAT(33X,'COST SUMMARY'
& /11X,
& 'Air Blown KRW-Based IGCC System with Hot Gas Cleanup')

```

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----- Report Cost Model Parameters -----

```

```

WRITE(NRPT,210)
210 FORMAT(/
& 'A. COST MODEL PARAMETERS -----',
& '-----')
WRITE(NRPT,220) CF, NCSTYR, FGF, CI, FICC, CICPPI,
& RTAX, EREAL, FEHO, INTRST, FPJ, NOYEARS,
& NSHIFTS, FBM, FCF, ALABOR, VCLF, IBKLIF
220 FORMAT(' Plant Capacity Factor: ',F6.2,
& ' Cost Year: January',I5,
& /' General Facilities Factor: ',F6.2,
& ' Plant Cost Index: ',F6.1,
& /' Indirect Construction: ',F6.2,
& ' Chemicals Cost Index: ',F6.1,
& /' Sales Tax: ',F6.2,
& ' Escalation: ',F6.2,
& /' Engr & Home Office Fee: ',F6.2,
& ' Interest: ',F6.2,
& /' Project Contingency: ',F6.2,
& ' Years of construction: ',I6,
& /' Number of Shifts: ',F6.2,
& ' Byproduct marketing: ',F6.2,
& /' Fixed Charge Factor: ',F6.4,
& ' Average Labor Rate: ',F6.2,
& /'Variable Levelization Cost Factor: ',F6.4,
& ' Book Life (years): ',I6)

```

```

WRITE(NRPT,230)
230 FORMAT('/B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS ',
& '-----',
& /26X' Process Maintenance',
& /26X'Plant Section Contingency Cost Factor',
& /26X'-----')
WRITE(NRPT,240) FPCCH, FMCCH, FPCL, FMCL, FPCOF, FMCOF,
& FPCG, FMCG, FPCZF, FMCZF, FPCS, FMCS,
& FPCBF, FMCBF, FPCGT, FMCCC, FPCHR, FMCCC,
& FPCST, FMCCC, FPCGF, FMCGF
240 FORMAT(10X,' Coal Handling ',10X,F6.3,10X,F6.3,
& /10X,' Limestone Handling ',10X,F6.3,10X,F6.3,
& /10X,' Oxidant Feed ',10X,F6.3,10X,F6.3,
& /10X,' Gasification ',10X,F6.3,10X,F6.3,
& /10X,' Zinc Ferrite ',10X,F6.3,10X,F6.3,
& /10X,' Sulfation ',10X,F6.3,10X,F6.3,
& /10X,' Boiler Feedwater Treatment ',10X,F6.3,10X,F6.3,
& /10X,' Gas Turbine ',10X,F6.3,10X,F6.3,

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&      /10X,'Heat Recovery Steam Generator ',10X,F6.3,10X,F6.3,
&      /10X,'                        Steam Turbine ',10X,F6.3,10X,F6.3,
&      /10X,'                        General Facilities ',10X,F6.3,10X,F6.3)
C
C ----- CAPITAL COSTS -----
C
      WRITE(NRPT,250)
250 FORMAT(/'C. DIRECT CAPITAL AND PROCESS ',
& 'CONTINGENCY COSTS ($1,000) ',
& '-----',
& /16X'          Number of Units          Direct          Process',
& /16X'Plant Section Operating Total Capital Cost Contingency',
& /16X'-----' )
      WRITE(NRPT,260) NOCH, NTCH, DCCH, CPCCH, NOL, NTL, DCL, CPCL,
&      NOOF, NTOF, DCOF, CPCOF, NOG, NTG, DCG, CPCG,
&      NOZF, NTZF, DCZF, CPCZF, NOS, NTS, DCS, CPCs,
&      NOBF, NTBFB, DCBFB, PCBFB, NOGT, NTGT, DCGT, PCPGT,
&      NOHR, NTHR, DCHR, PCCHR, NOST, NTST, DCST, PCST,
&      DCGF, PCGF
260 FORMAT(
& /'          Coal Handling ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          Limestone Handling ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          Oxidant Feed ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          Gasification ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          Zinc Ferrite ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          Sulfation ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /' Boiler Feedwater Treatment ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          Gas Turbine ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'Heat Recovery Steam Generator ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          Steam Turbine ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
& /'          General Facilities ',5X,' N/A',1X,' N/A',
& 3X,F10.0,3X,F10.0)
      WRITE(NRPT,270) TDC, CICC, CTAX, CEHO, CEP, TIC, CPC, CPJ, TPC,
& AFDC, TPI, PPC, IC, TCICC, TCLAND, TCR
270 FORMAT(/'D. TOTAL CAPITAL REQUIREMENT ($1,000) ----',
& '-----',
& /10X,'          Description          '
& ,8X,'Annual Cost',
& /10X,'          -----'
& ,8X,'-----',
& /10X,'          Total Direct Cost          ',8X,F10.0,
& /10X,'          Indirect Construction Cost          ',8X,F10.0,
& /10X,'          Sales Tax          ',8X,F10.0,
& /10X,'          Engineering and Home Office Fees          ',8X,F10.0,
& /10X,'          Environmental Permitting          ',8X,F10.0,
& /10X,'          Total Indirect Costs          ',8X,F10.0,
& /10X,'          Total Process Contingencies          ',8X,F10.0,
& /10X,'          Project Contingency          ',8X,F10.0,
& /10X,'          Total Plant Cost          ',8X,F10.0,
& /10X,'          AFDC          ',8X,F10.0,
& /10X,'          Total Plant Investment          ',8X,F10.0,
& /10X,'          Preproduction (Startup) Costs          ',8X,F10.0,
& /10X,'          Inventory Capital          ',8X,F10.0,
& /10X,'          Initial Catalysts and Chemicals          ',8X,F10.0,
& /10X,'          Land          ',8X,F10.0,
& /10X,'TOTAL CAPITAL REQUIREMENT ($1,000) ----->',8X,F10.0)
C
C ----- FIXED OPERATING COSTS -----

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C
WRITE(NRPT,280) OCL, OCM, OCAS, FOC
280 FORMAT(/'E.  FIXED OPERATING COSTS ($/year)  -----',
& '-----',
& /10X,'          Description',          ',15X,'Annual Cost',
& /10X,'          -----',          ',15X,'-----',
& /10X,'          Operating Labor',          ',11X,F11.0,
& /10X,'          Maintenance Costs',          ',11X,F11.0,
& /10X,'          Administration and Supervision',11X,F11.0,
& /10X,'TOTAL FIXED OPERATING COST ',
& ' ($/year) ----->',F11.0)

C
C ----- Unit Costs for Fuel, Chemicals, and Catalysts -----
C
WRITE(NRPT,290) UCSAI, MSAC, OCSAI,
& UCNAOH, MNAOH, OCNAOH, UCNA2H, MSPBFI, OCNA2H,
& UCHYDR, MHYBFI, OCHYDR, UCMORP, MMOBFI, OCMORP,
& UCLIME, MLMCWI, OCLIME, UCSODA, MSOCWI, OCSODA,
& UCCORI, MCICWI, OCCORI
290 FORMAT (/
& 'F.  VARIABLE OPERATING COSTS  -----',
& '-----',
& /'      1.  CONSUMABLES ($/year)',
& /6X,'          ',
& ' Material          Annual',
& /6X,'Description          Unit Cost',
& ' Requirement          Operating Cost',
& /6X,'-----',
& '-----',
& /6X,' Sulfuric Acid: ',F7.2,' $/ton',F10.1,' ton/yr',F14.0,
& /6X,'      NaOH: ',F7.2,' $/ton',F10.1,' ton/yr',F14.0,
& /6X,'      Na2 HPO4: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0,
& /6X,'      Hydrazine: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0,
& /6X,'      Morpholine: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0,
& /6X,'      Lime: ',F7.2,' $/ton',F10.1,' ton/yr',F14.0,
& /6X,'      Soda Ash: ',F7.2,' $/ton',F10.1,' ton/yr',F14.0,
& /6X,'Corrosion Inh.: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0)
WRITE(NRPT,300) UCSURF, MSUCWI, OCSURF, UCCHLR, MCLCWI, OCCHLR,
& UCBIOC, MBIOCWI, OCBIOC, UCZFSO, MSOZFI, OCZFSO,
& UCLMST, MLMANN, OCLMST
300 FORMAT (
& 6X,'      Surfactant: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0,
& /6X,'      Chlorine: ',F7.2,' $/ton',F10.1,' ton/yr',F14.0,
& /6X,'      Biocide: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0,
& /6X,'      Zinc Fer Sorb: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0,
& /6X,'      Limestone: ',F7.2,' $/ton',F10.1,' ton/yr',F14.0)
WRITE(NRPT,310) UCFUEL, MFOGTI, OCFOIL, UCPLTA, MADSGFI, OCPLTA,
& UCRAWW, MWATER, OCRAWW, UCWWCH, WWATER, CWW,
& UCFLRL, MLPGI, OCFLRL, OCCONS
310 FORMAT (
& 6X,'      Fuel Oil: ',F7.2,' $/bbl',F10.1,' bbl/yr',F14.0,
& /6X,'Plant Air Ads.: ',F7.2,' $/lb',F10.1,' lb/yr',F14.0,
& /6X,'      Raw Water: ',F7.2,' $/Kgal',F10.1,' Kgal/yr',F14.0,
& /6X,'      Waste Water: ',F7.2,' $/gpm ww',F10.1,' lb/hr',F14.0,
& /6X,'      LPG - Flare: ',F7.2,' $/bbl',F10.1,' bbl/yr',F14.0,
& /6X,'TOTAL CONSUMABLES ($/year) ----->',F14.0)
WRITE(NRPT,320) UCCOAL, MCFGI, OCFUEL, UCASHD,
& ((MFINE1+MFINE2)*24./2000.),OCASH, VOC

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320 FORMAT(
& /'      2.  FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)',
& /6X,'      Coal: ',F7.2,' $/MMBtu ',F10.1,' lb/hr      ',F14.0,
& /6X,'      Ash Disposal: ',F7.2,' $/ton      ',F10.1,' ton/day ',F14.0,
& //6X,'TOTAL VARIABLE OPERATING COST ($/year) -----> ',
& ,F11.0)

C
C ----- Normalized Costs and Cost of Electricity -----
C
      WRITE(NRPT,330) WGTE, WCHE, WSE, WSTE, WLE, WBFE, WAUXE, WOFE,
&      WGFE,WGE, WNETE, XO, DPERKW, FOCN, VOCINC,
&      BYPN, FUELN, VOCN, CELEC
330 FORMAT (/
&      'G.  COST OF ELECTRICITY -----',
&      '-----',
&      /'      Power Summary (MWe)      ',
&      'Auxiliary Loads (MWe)',
&      /'-----',
&      '-----',
&      /'Gas Turbine Output      ',F7.2,
&      '      Coal Handling ',F6.2,' Sulfation      ',F6.2,
&      /'Steam Turbine Output ',F7.2,
&      '      Limestne Hdlg ',F6.2,' Steam Cycle ',F6.2,
&      /'Total Auxiliary Loads ',F7.2,
&      '      Oxidant Feed ',F6.2,' General Fac ',F6.2,
&      /'-----',
&      '      Gasification ',F6.2,
&      /'Net Electricity      ',F7.2,
&      '      Zinc Ferrite ',F6.2,
&      //27X,'      Capital Cost: ',F10.2,' $/kW',
&      /27X,'      Fixed Operating Cost: ',F10.2,' $/(kW-yr)',
&      /8X,'Incremental Variable Costs: ',F6.2,' mills/kWh',
&      /8X,'Byproduct Credit:      ',F6.2,' mills/kWh',
&      /8X,'Fuel Cost:      ',F6.2,' mills/kWh',
&      /27X,'Variable Operating Cost: ',F10.2,' mills/kWh',
&      /'COST OF ELECTRICITY ----->',
&      F7.2,' mills/kWh'/)
      WRITE(NRPT,340) HEATRATE, EFFICNCY
340 FORMAT(5X,'(Heat Rate is:',F7.0,' Efficiency is:',F7.4,')')
      WRITE(NRPT,350) MCOALIN, MH2OIN, MLMSTIN,MASHOUT,
&      ECO2, ECO, ESO2, ENOX
350 FORMAT(/
&      'H.  ENVIRONMENTAL SUMMARY -----',
&      '-----',
&      /15X,'INPUTS:      Coal      ',F10.3,' lb/kWh',
&      /15X,'      Water      ',F10.3,' lb/kWh',
&      /15X,'      Limestone',F10.3,' lb/kWh',
&      /15X,'OUTPUTS      Ash      ',F10.3,' lb/kWh',
&      /15X,'      CO2      ',F10.3,' lb/kWh',
&      /15X,'      CO      ',F10.3,' lb/kWh',
&      /15X,'      SO2      ',F10.4,' lb/MMBtu',
&      /15X,'      NOx      ',F10.4,' lb/MMBtu'//)
      RETURN
      END

```

B.3 Sample Output

The following is a sample output from an ASPEN simulation of the air-blown KRW-based IGCC system with hot gas cleanup 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.

*** SUMMARY OF COST MODEL INPUT PERFORMANCE PARAMETERS ***

PLANT SECTION	INPUT PERFORMANCE PARAMETERS	VALUE	UNITS
Coal Feed	Mass flow of coal to gasifier	471973.49	lb/hr
Limestone Feed	Mass flow of limestone (mixed)	2042.85	lb/hr
	Mass flow of limestone (cisolid)	100099.51	lb/hr
Oxidant Feed	Work to boost air compressor	0.329687E+08	Watts
	First precoolers inlet air temp.	719.75	F
	First precoolers outlet air temp.	550.00	F
	First precoolers inlet BFW temp.	505.57	F
	First precoolers outlet BFW temp.	605.29	F
	Heat transfer in first precoolers	0.6721E+08	BTU/hr
	Heat trans. to BFW from HRSG	0.5106E+09	BTU/hr
	Heat leaving from economizer	0.4300E+09	BTU/hr
	Second precoolers outlet air temp.	414.12	F
	Second precoolers inlet BFW temp.	241.30	F
	Second precoolers outlet BFW temp.	501.47	F
	Heat transfer in second precoolers	0.5299E+08	BTU/hr
	Heat trans. to BFW from HRSG	0.6525E+09	BTU/hr
	Heat leaving from economizer	0.2198E+09	BTU/hr
Gasification	Moisture in coal feed	12.00	wt %
	Ash in coal	10.00	wt %
Zinc Ferrite	Zinc ferrite inlet syngas pres.	465.00	psia
	Zinc ferrite inlet syngas temp.	1140.08	F
	Zinc ferrite inlet syngas flow	102538.44	lbmole/hr
	Zinc ferrite inlet H ₂ S flow	52.39	lbmole/hr
	Zinc ferrite inlet COS flow	2.51	lbmole/hr
Sulfation	Gasifier fines and ash (NC)	48172.28	lb/hr
	Gasifier fines and ash (CISOLID)	96168.34	lb/hr
BFW Treating	Raw water to power plant	258255.85	lb/hr
	Quench water to hot syngas	311697.71	lb/hr
	Steam turbine condensate	1458707.80	lb/hr
Gas Turbine	Gas turbine inlet air flow	9000000.04	lb/hr
Steam Cycle	HRSG outlet HP steam pres.	1465.00	psia
	HRSG outlet HP steam flow	1108391.15	lb/hr

	Temp. of ambient air	59.00 F
	Heating value of coal	12774.00 BTU/lb
	Gas turbine net shaft work	-.444333E+09 Watts
	Steam turbine net shaft work	-.261962E+09 Watts
	Steam cycle auxiliary power	0.298344E+07 Watts
Emissions	CO2 emissions from gas turbine	24605.42 lbmole/hr
	CO emissions from gas turbine	0.00 lbmole/hr
	SO2 emissions from gas turbine	1.08 lbmole/hr
	NH3 entering gas turbine	23.24 lbmole/hr
	Coal Heating Value	11248.73 BTU/lb
COST VAR WARNING ---- Variable WCME/N value of 10989.551		
	in DCCM above the upper limit of	8200.000

COST SUMMARY
Air Blown KRW-Based IGCC System with Hot Gas Cleanup

A. COST MODEL PARAMETERS -----

Plant Capacity Factor:	0.65	Cost Year:	January 1989
General Facilities Factor:	0.15	Plant Cost Index:	351.5
Indirect Construction:	0.25	Chemicals Cost Index:	411.3
Sales Tax:	0.06	Escalation:	0.00
Engr & Home Office Fee:	0.15	Interest:	0.06
Project Contingency:	0.20	Years of construction:	4
Number of Shifts:	4.75	Byproduct marketing:	0.10
Fixed Charge Factor:	0.1034	Average Labor Rate:	19.70
Variable Levelization Cost Factor:	1.0000	Book Life (years):	30

B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS -----

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.020
Limestone Handling	0.000	0.030
Oxidant Feed	0.050	0.020
Gasification	0.375	0.045
Zinc Ferrite	0.400	0.045
Sulfation	0.500	0.040
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.030	0.015
Heat Recovery Steam Generator	0.030	0.015
Steam Turbine	0.030	0.015
General Facilities	0.100	0.015

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

Plant Section	Number of Units		Direct Capital Cost	Process Contingency
	Operating	Total		
Coal Handling	1	1	26464.	1981.
Limestone Handling	1	1	3816.	0.
Oxidant Feed	3	3	8293.	621.
Gasification	5	6	90493.	50800.
Zinc Ferrite	10	24	14200.	8503.
Sulfation	1	1	25751.	19275.
Boiler Feedwater Treatment	1	1	4699.	0.
Gas Turbine	3	3	96000.	4311.
Heat Recovery Steam Generator	3	3	27391.	1230.

Steam Turbine	1	1	40950.	1839.
General Facilities	N/A	N/A	50709.	7591.

D. TOTAL CAPITAL REQUIREMENT (\$1,000) -----		
Description		Annual Cost
Total Direct Cost		388766.
Indirect Construction Cost		97192.
Sales Tax		19244.
Engineering and Home Office Fees		75780.
Environmental Permitting		1000.
Total Indirect Costs		193216.
Total Process Contingencies		96151.
Project Contingency		135627.
Total Plant Cost		813760.
AFDC		77864.
Total Plant Investment		891624.
Preproduction (Startup) Costs		24046.
Inventory Capital		16037.
Initial Catalysts and Chemicals		26539.
Land		1943.
TOTAL CAPITAL REQUIREMENT (\$1,000) ----->		969106.

E. FIXED OPERATING COSTS (\$/year) -----		
Description		Annual Cost
Operating Labor		7396168.
Maintenance Costs		21892612.
Administration and Supervision		4845964.
TOTAL FIXED OPERATING COST (\$/year) ----->		34134744.

F. VARIABLE OPERATING COSTS -----

1. CONSUMABLES (\$/year)

Description	Unit Cost	Material Requirement	Annual Operating Cost
Sulfuric Acid:	110.00 \$/ton	1442.1 ton/yr	158629.
NaOH:	220.00 \$/ton	322.4 ton/yr	70920.
Na2 HPO4:	0.70 \$/lb	1412.1 lb/yr	989.
Hydrazine:	3.20 \$/lb	6790.0 lb/yr	21728.
Morpholine:	1.30 \$/lb	6311.7 lb/yr	8205.
Lime:	80.00 \$/ton	520.5 ton/yr	41637.
Soda Ash:	160.00 \$/ton	578.7 ton/yr	92592.
Corrosion Inh.:	1.90 \$/lb	103492.0 lb/yr	196635.
Surfactant:	1.25 \$/lb	103492.0 lb/yr	129365.
Chlorine:	250.00 \$/ton	16.7 ton/yr	4180.
Biocide:	3.60 \$/lb	18602.0 lb/yr	66967.
Zinc Fer Sorb:	3.30 \$/lb	120856.5 lb/yr	398826.
Limestone:	18.00 \$/ton	290798.3 ton/yr	5234369.
Fuel Oil:	42.00 \$/bbl	36633.2 bbl/yr	1538593.
Plant Air Ads.:	2.80 \$/lb	2747.5 lb/yr	7693.
Raw Water:	0.73 \$/Kgal	389021.8 Kgal/yr	283986.
Waste Water:	840.00 \$/gpm ww	0.0 lb/hr	0.
LPG - Flare:	11.70 \$/bbl	3205.4 bbl/yr	37503.
TOTAL CONSUMABLES (\$/year) ----->			8292817.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT (\$/year)		
Coal:	1.61 \$/MMBtu	471973.5 lb/hr
		55269795.

Ash Disposal: 10.00 \$/ton 1732.1 ton/day 4109377.

TOTAL VARIABLE OPERATING COST (\$/year) -----> 67671989.

G. COST OF ELECTRICITY -----

Power Summary (MWe)		Auxiliary Loads (MWe)			
Gas Turbine Output	437.67	Coal Handling	3.40	Sulfation	1.47
Steam Turbine Output	258.03	Limestone Hdlg	0.46	Steam Cycle	2.98
Total Auxiliary Loads	49.92	Oxidant Feed	32.97	General Fac	5.53
		Gasification	3.11		
Net Electricity	645.78	Zinc Ferrite	0.00		

	Capital Cost:	1500.68 \$/kW
	Fixed Operating Cost:	52.86 \$/(kW-yr)
Incremental Variable Costs:	3.37 mills/kWh	
Byproduct Credit:	0.00 mills/kWh	
Fuel Cost:	15.03 mills/kWh	
	Variable Operating Cost:	18.40 mills/kWh
COST OF ELECTRICITY		54.94 mills/kWh

(Heat Rate is: 8221. Efficiency is: 0.4154)

H. ENVIRONMENTAL SUMMARY -----

INPUTS:	Coal	0.731 lb/kWh
	Water	0.883 lb/kWh
	Limestone	0.158 lb/kWh
OUTPUTS	Ash	0.224 lb/kWh
	CO2	1.676 lb/kWh
	CO	0.000 lb/kWh
	SO2	0.0131 lb/MMBtu
	NOx	0.0944 lb/MMBtu

APPENDIX C: COST MODEL FOR LURGI-BASED IGCC SYSTEM WITH HOT GAS CLEANUP

The cost model of an air-blown Lurgi-based IGCC system with hot gas cleanup is implemented in Fortran as a subroutine. This subroutine estimates direct capital costs based on the equations developed in Section 4 of the report, and estimates the total capital cost and annual costs based on the approach documented in Section 5 and 6, respectively. The values of the performance variables required for the cost model are taken from the ASPEN flowsheet of a Lurgi-based IGCC system with hot gas cleanup developed by Klara, Rastogi, and Craig (1988). 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 a Lurgi-based IGCC system with hot gas cleanup. The code includes the ASPEN Fortran block ALHCOST for inclusion in the ASPEN simulation model of the IGCC system and a subroutine USRALHC 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.

C.1 ASPEN Fortran Block ALHCOST

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.

FORTTRAN ALHCOST

```
;
;   Author:  H.C. Frey, Carnegie-Mellon University
;   Date:    March 1, 1990
;   Description: This block accesses the key variables required to
;   estimate capital and annual costs, and then calls subroutine
;   USRALHC (User Air-blown Lurgi-based igcc with Hot gas cleanup
;   Cost model).
;
F    REAL*8
F    1  TWPC1I, TWPC1O,  QAPC1, QW1PC1, QW2PC1, QW3PC1, PSNZFI, TSNZFI,
F    2  DSNZFI, MSNZFI, MOGSAI, MSOSAI, TOGSAI, MHSZFI, MCSZFI, MDISP1,
F    3  MDISP2, MDISP3, MDISP4,  MRW,  MCOND,  MAGTI, PHSHRO, MSHSHRO,
F    4  GTNETS, STNETS, STAUXE, ACAUXE, PERH2O, PERASH,  TA, HVCOAL,
F    5  MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT, MCH4OUT, MH2SOUT, MNH3IN, MRECYC
F    COMMON/USRCST3/  MCFGI, SORB1, SORB2,  WCME, TAPC1I, TAPC1O,
F    1  TWPC1I, TWPC1O,  QAPC1, QW1PC1, QW2PC1, QW3PC1, PSNZFI, TSNZFI,
```

```

F 2 DSNZFI, MSNZFI, MOGSAI, MSOSAI, TOGSAI, MHSZFI, MCSZFI, MDISP1,
F 3 MDISP2, MDISP3, MDISP4, MRW, MCOND, MAGTI, PHSHRO, MHSHRO,
F 4 GTNETS, STNETS, STAUXE, ACAUXE, PERH2O, PERASH, TA, HVCOAL,
F 5 MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT, MCH4OUT, MH2SOUT, MNH3IN
F COMMON/USRHOC/ HOC
F COMMON/USRREC/ RECYC
C
C NOTE: get fines recycle value from stochastic simulation
C
;
DESC 'Sample Cost Model Inputs and Call Cost Subroutine'
IN-UNITS ENG
;
; VARIABLE DEFINITIONS -----
;
; MCFG1 = Mass flow of coal to gasifier, lb/hr
; SORB1 = Mass flow of limestone sorbent to gasifier (mixed), lb/hr
; SORB2 = Mass flow of limestone sorbent to gasifier (cisolid), lb/hr
; WCME = Work to boost air compressor, BTU/hr
; TAPC1I = Temperature of boost air at first precooler inlet, F
; TAPC1O = Temperature of boost air at first precooler outlet, F
; TWPC1I = Temperature of BFW at first precooler inlet, F
; TWPC1O = Temperature of BFW at first precooler outlet, F
; QAPC1 = Heat transferred from air in first precooler, BTU/hr
; QW1PC1 = Heat transferred to BFW from HRSG, BTU/hr
; QW2PC1 = Heat transferred to BFW from regenerator, BTU/hr
; QW3PC1 = Heat leaving from economizer, BTU/hr
; PSNZFI = Pressure of syngas at zinc ferrite inlet, psia
; TSNZFI = Temperature of syngas at zinc ferrite inlet, F
; DSNZFI = Molar density of syngas at ZF inlet, lbmole/ft3
; MSNZFI = Molar flow of syngas at zinc ferrite inlet, lbmole/hr
; MOGSAI = Molar flow of off-gas to sulfuric acid plant, lbmole/hr
; MSOSAI = Molar flow of SO2 to sulfuric acid plant, lbmole/hr
; TSOSAI = Temperature of off-gas to sulfuric acid plant, F
; MHSZFI = Molar flow of hydrogen sulfide (H2S) at zinc ferrite
; inlet, lbmole/hr
; MCSZFI = Molar flow of carbonyl sulfide (COS) at zinc ferrite
; inlet, lbmole/hr
; MDISP1 = Mass flow (NC) of ash leaving gasifier, lb/hr
; MDISP2 = Mass flow (CISOLID) of ash leaving gasifier, lb/hr
; MDISP3 = Mass flow (NC) of fines leaving gasifier, lb/hr
; MDISP4 = Mass flow (CISOLID) of fines leaving gasifier, lb/hr
; MRW = Mass flow of raw water to power plant, lb/hr
; MCOND = Mass flow of steam turbine condensation, lb/hr
; MAGTI = Mass flow of air to gas turbine inlet, lb/hr
; PHSHRO = Pressure of high pressure steam at HRSG outlet, psia
; MHSHRO = Mass flow of high pressure steam at HRSG outlet, lb/hr
; GTNETS = Gas turbine net shaft work, BTU/hr
; STNETS = Steam turbines net shaft work, BTU/hr
; STAUXE = Steam cycle auxiliaries power consumption, BTU/hr
; ACAUXE = Acid gas removal auxiliaries power consumption, BTU/hr
; PERH2O = Moisture content of coal feed, weight percent
; PERASH = Ash content of coal, weight percent
; TA = Temperature of ambient air, F
; HVCOAL = Heating value of coal, BTU/lb
; MCO2OUT = CO2 leaving gas turbine, lbmole/hr
; MCOOUT = CO leaving gas turbine, lbmole/hr
; MSO2OUT = SO2 leaving gas turbine, lbmole/hr

```

```

;      MCOSOUT= COS leaving gas turbine, lbmole/hr
;      MCh4OUT= CH4 leaving gas turbine, lbmole/hr
;      MH2SOUT= H2S leaving gas turbine, lbmole/hr
;      MNH3IN = NH3 entering gas turbine, lbmole/hr
;
DEFINE MCFG1    SUBSTREAM-VAR  STREAM=COALFEED  SUBS=NC        VAR=MASSFLOW
DEFINE SORB1    SUBSTREAM-VAR  STREAM=CALSORB   SUBS=MIXED     VAR=MASSFLOW
DEFINE SORB2    SUBSTREAM-VAR  STREAM=CALSORB   SUBS=CISOLID   VAR=MASSFLOW
DEFINE WCME     STRM-ATTR-VAR  ATTR=WORK        VAR=POWER      STREAM=WBOOST
DEFINE TAPC1I   SUBSTREAM-VAR  STREAM=TOBOOST   SUBS=MIXED     VAR=TEMP
DEFINE TAPC1O   SUBSTREAM-VAR  STREAM=BOOSTAIR   SUBS=MIXED     VAR=TEMP
DEFINE TWPC1I   SUBSTREAM-VAR  STREAM=VHPBFW     SUBS=MIXED     VAR=TEMP
DEFINE TWPC1O   SUBSTREAM-VAR  STREAM=HOTHFBFW   SUBS=MIXED     VAR=TEMP
DEFINE QAPC1    STRM-ATTR-VAR  ATTR=HEAT        VAR=Q          STREAM=QAIRHX1
DEFINE QW1PC1   STRM-ATTR-VAR  ATTR=HEAT        VAR=Q          STREAM=QHRSG2
DEFINE QW2PC1   STRM-ATTR-VAR  ATTR=HEAT        VAR=Q          STREAM=QREGEN1
DEFINE QW3PC1   STRM-ATTR-VAR  ATTR=HEAT        VAR=Q          STREAM=QLPBOIL
DEFINE PSNZFI   STREAM-VAR     STREAM=RAWGAS    VAR=PRES
DEFINE TSNZFI   STREAM-VAR     STREAM=RAWGAS    VAR=TEMP
DEFINE DSNZFI   STREAM-VAR     STREAM=RAWGAS    VAR=MOLE-DENSITY
DEFINE MSNZFI   SUBSTREAM-VAR  STREAM=RAWGAS    SUBS=MIXED     VAR=MOLEFLOW
DEFINE MOGSAI   SUBSTREAM-VAR  STREAM=ACIDFD     SUBS=MIXED     VAR=MOLEFLOW
DEFINE MSOSAI   MOLE-FLOW      STREAM=ACIDFD     SUBS=MIXED     COMP=SO2
DEFINE TOGSAI   SUBSTREAM-VAR  STREAM=ACIDFD     SUBS=MIXED     VAR=TEMP
DEFINE MHSZFI   MOLE-FLOW      STREAM=RAWGAS    SUBS=MIXED     COMP=H2S
DEFINE MCSZFI   MOLE-FLOW      STREAM=RAWGAS    SUBS=MIXED     COMP=COS
DEFINE MDISP1   SUBSTREAM-VAR  STREAM=SOLID      SUBS=NC        VAR=MASSFLOW
DEFINE MDISP2   SUBSTREAM-VAR  STREAM=SOLID      SUBS=CISOLID   VAR=MASSFLOW
DEFINE MDISP3   SUBSTREAM-VAR  STREAM=FINES      SUBS=NC        VAR=MASSFLOW
DEFINE MDISP4   SUBSTREAM-VAR  STREAM=FINES      SUBS=CISOLID   VAR=MASSFLOW
DEFINE MRW       SUBSTREAM-VAR  STREAM=MAKEUP    SUBS=MIXED     VAR=MASSFLOW
DEFINE MCOND     SUBSTREAM-VAR  STREAM=WATER25   SUBS=MIXED     VAR=MASSFLOW
DEFINE MAGTI     SUBSTREAM-VAR  STREAM=GTAIR     SUBS=MIXED     VAR=MASSFLOW
DEFINE PHSHRO    SUBSTREAM-VAR  STREAM=SHSTEAM   SUBS=MIXED     VAR=PRES
DEFINE MHSHRO    SUBSTREAM-VAR  STREAM=SHSTEAM   SUBS=MIXED     VAR=MASSFLOW
DEFINE GTNETS    STRM-ATTR-VAR  ATTR=WORK        VAR=POWER      STREAM=WGASTURB
DEFINE STNETS    STRM-ATTR-VAR  ATTR=WORK        VAR=POWER      STREAM=WSTMTURB
DEFINE STAUXE    STRM-ATTR-VAR  ATTR=WORK        VAR=POWER      STREAM=WSTMISC
DEFINE ACAUXE    STRM-ATTR-VAR  ATTR=WORK        VAR=POWER      STREAM=ZF-WKMSC
;
;  Miscellaneous other inputs
;
DEFINE PERH2O    COMP-ATTR-VAR  STREAM=COALFEED  SUBS=NC        COMP=COAL &
                  ATTR=PROXANAL ELEMENT=1
DEFINE PERASH    COMP-ATTR-VAR  STREAM=COALFEED  SUBS=NC        COMP=COAL &
                  ATTR=PROXANAL ELEMENT=4
DEFINE TA        SUBSTREAM-VAR  STREAM=GTAIR     SUBS=MIXED     VAR=TEMP
DEFINE HVCOAL    COMP-ATTR-VAR  STREAM=COALFEED  SUBS=NC        COMP=COAL &
                  ATTR=COALMISC ELEMENT=1
;
;  Environmental Discharges
;
DEFINE MCO2OUT   MOLE-FLOW      STREAM=OUTGAS    SUBS=MIXED     COMP=CO2
DEFINE MCOOUT    MOLE-FLOW      STREAM=OUTGAS    SUBS=MIXED     COMP=CO
DEFINE MSO2OUT   MOLE-FLOW      STREAM=OUTGAS    SUBS=MIXED     COMP=SO2
DEFINE MCOSOUT   MOLE-FLOW      STREAM=OUTGAS    SUBS=MIXED     COMP=COS
DEFINE MCH4OUT   MOLE-FLOW      STREAM=OUTGAS    SUBS=MIXED     COMP=CH4
DEFINE MH2SOUT   MOLE-FLOW      STREAM=OUTGAS    SUBS=MIXED     COMP=H2S

```

```
DEFINE MNH3IN MOLE-FLOW      STREAM=BURNFD      SUBS=MIXED      COMP=NH3
;
F      CALL USRALHC
;
;      EXECUTE AFTER ST-MISC; (ST-MISC IS THE LAST BLOCK IN THE SIMULATION)
;
```

C.2 Cost Model Subroutine USRALHC

The cost model subroutine calculates the capital and annual costs for the Lurgi-based IGCC system with hot gas cleanup discussed in Section 4. This subroutine calls subroutines USRWCK, USRRCK, USRFCF, and USRVCLF. USRWCK and USRRCK are used for range-checking and are listed in Appendix E. USRFCF and USRVCLF are used to calculate the fixed charge and variable cost levelization factors required to estimate the cost of electricity. These are listed in Appendix F.

```
SUBROUTINE USRALHC
IMPLICIT DOUBLE PRECISION (A-H,O-Z)

C
C   Revision 1.00: H.C. Frey                               Date: March 1, 1990
C   Short Title: Cost Model for Air-blown Lurgi-IGCC With Hot Gas Cleanup
C   Description: Capital and annual cost model of Lurgi-based integrated
C   gasification combined cycle system with hot gas cleanup. Based on
C   ASPEN input file developed by Klara, Rastogi, Regenhardt, Stone,
C   and Craig over time.
C   Cost model documented in Frey and Rubin (1990).
C
C   The cost basis for the model is January 1989 (Chemical Engineering
C   Plant Cost Index = 351.5).
C
C   This subroutine requires the use of ENGLISH units in ASPEN.
C
REAL*8      MCFG1, SORB1, SORB2, WCME, TAPC1I, TAPC1O,
1 TWPC1I, TWPC1O, QAPC1, QW1PC1, QW2PC1, QW3PC1, PSNZFI, TSNZFI,
2 DSNZFI, MSNZFI, MOGSAI, MSOSAI, TOGSAI, MHSZFI, MCSZFI, MDISP1,
3 MDISP2, MDISP3, MDISP4, MRW, MCOND, MAGTI, PHSHRO, MSHSHRO,
4 GTNETS, STNETS, STAUKE, ACAUXE, PERH2O, PERASH, TA, HVCOAL,
5 MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT, MCH4OUT, MH2SOUT, MNH3IN

C
REAL*8 MSABFI, MSHBFI, MSPBFI, MHYBFI, MMOBFI, MSABFPI, MSHBFPI,
1 MCWI, MLMCWI, MSOCWI, MSACWI, MCICWI, MSUCWI, MCLCWI, MBIOCWI,
2 MSSSI, MCATCI, MCATBSI, MFOGTI, MADSGFI, MWATER, MLPGI, MSAC,
3 IC, MNAOH, MLMANN, MSOZFI, MSACAT, MCOKE, MBENT, MSZFI, MACID,
4 MCOALIN, MH2OIN, MASHOUT, MACIDO
REAL*8 INTRST, MASH, MCMGI, MPW, NSHIFTS, NOPERS, MSULFR,
1 DCCH, DCL, DCCM, DCPC, DCOF, DCG, DCZF, DCS, DCBF, DCGT,
2 DCHR, DCST, DCGF
REAL*8 GENEFF

C
C   Common blocks for input performance variables:
C   USRCST3 passes simulation variables accessed in block ALHCOST
C   USER passes variables from the ASPEN simulation, including NRPT
C   USRAK1 passes cost model parameters
C   USRAK2 passes capital cost uncertainty factors
C   USRAK3 passes maintenance cost factors
C   USRAK4 passes regression model error terms
C   USRAK5 passes miscellaneous other uncertainties, including unit
costs.
C
COMMON/USRCST3/ MCFG1, SORB1, SORB2, WCME, TAPC1I, TAPC1O,
```



```

1 TWPC1I, TWPC1O, QAPC1, QW1PC1, QW2PC1, QW3PC1, PSNZFI, TSNZFI,
2 DSNZFI, MSNZFI, MOGSAI, MSOSAI, TOGSAI, MHSZFI, MCSZFI, MDISP1,
3 MDISP2, MDISP3, MDISP4, MRW, MCOND, MAGTI, PHSHRO, MSHSHRO,
4 GTNETS, STNETS, STAUXE, ACAUXE, PERH2O, PERASH, TA, HVCOAL,
5 MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT, MCH4OUT, MH2SOUT, MNH3IN
COMMON/USRHOC/ HOC
COMMON /USER/ RMISS, IMISS, NGBAL, IPASS, IRESTR, ICONVG, LMSG,
1 LPMMSG, KFLAG, NHSTRY, NRPT, NTRMNL
COMMON /USRRPT/ NREPT
COMMON /USRAL1/ CF, EREAL, FBM, FEHO, FGF, FICC, FPJ,
1 INTRST, EINF, NOYEARS, IBKLIF, RTAX, ALABOR,
2 NSHIFTS
COMMON /USRAL2/ FPCCH, FPCL, FPCOF, FPCG, FPCSS, FPCCY,
1 FPCZF, FPCS, FPCSA, FPCBF, FPCGT, FPCR,
2 FPCST, FPCGF
COMMON /USRAL3/ FMCOF, FMCCCH, FMCL, FMCG, FMCCS, FMCCY,
1 FMCZF, FMCS, FMCSA, FMCBF, FMCCC, FMCCGF
COMMON /USRAL4/ ERRCH, ERROF, ERRSA,
1 ERBBF, ERBBR, ERBST, ERBWH
COMMON /USRAL5/ UCLMST, UPACID, UPSULF, UCASHD,
1 UCZFSO, SLZF, VSZF, TAZF
COMMON/USRREC/RECYC
NREPT=NRPT

```

C
C
C

CONVERT UNITS FROM BTU/HR TO WATTS

```

WCME = 0.292875*WCME
GTNETS = 0.292875*GTNETS
STNETS = 0.292875*STNETS
STAUXE = 0.292875*STAUXE
ACAUXE = 0.292875*ACAUXE
HOC = 0.000430210*HOC

```

C
C

XO=0.0

C
C
C

--- Print out the input performance parameters -----

```

WRITE(NRPT,100) MCFGI, SORB1, SORB2, WCME,
& TAPC1I, TAPC1O, TWPC1I, TWPC1O
WRITE(NRPT,110) QAPC1, QW1PC1, QW2PC1, QW3PC1,
& PSNZFI, TSNZFI, DSNZFI, MSNZFI
WRITE(NRPT,120) MOGSAI, MSOSAI, TOGSAI, MHSZFI,
& MCSZFI, MDISP1, MDISP2, MDISP3, MDISP4
WRITE(NRPT,130) MRW, MCOND, MAGTI, PHSHRO, MSHSHRO,
& GTNETS, STNETS, STAUXE, ACAUXE, PERH2O,
& PERASH, TA, HVCOAL
WRITE(NRPT,135) MCO2OUT, MCOOUT, MSO2OUT, MCOSOUT,
& MCH4OUT, MH2SOUT, MNH3IN, HOC

```

C

```

100 FORMAT(' *** SUMMARY OF COST MODEL INPUT ',
& 'PERFORMANCE PARAMETERS ***'//
& 'PLANT SECTION ', 'INPUT PERFORMANCE PARAMETERS',
& 8X, 'VALUE', 'UNITS',/
& '-----', '-----',/
& 8X, '-----', '-----',/
& 'Coal Feed ',
& 'Mass flow of coal to gasifier ', F12.2, ' lb/hr ',/
& 'Limestone Feed ',

```

```

&      'Mass flow of limestone (mixed)      ',F12.2,' lb/hr ',/
& 16X,'Mass flow of limestone (cisolid)    ',F12.2,' lb/hr ',/
&      'Oxidant Feed      ',
&      'Work to boost air compressor        ',E12.6,' Watts',/
& 16X,'First precoolers inlet air temp.    ',F12.2,' F ',/
& 16X,'First precoolers outlet air temp.   ',F12.2,' F ',/
& 16X,'First precoolers inlet BFW temp.    ',F12.2,' F ',/
& 16X,'First precoolers outlet BFW temp.   ',F12.2,' F ')
110 FORMAT(
& 16X,'Heat transfer in first precoolers   ',E12.6,' BTU/hr',/
& 16X,'Heat trans. to BFW from HRSG        ',E12.6,' BTU/hr',/
& 16X,'Heat trans. to BFW from Regen.      ',E12.6,' BTU/hr',/
& 16X,'Heat leaving from economizer        ',E12.6,' BTU/hr',/
&      'Gasification      ',
&      'Gasifier output syngas pres.       ',F12.2,' psia ',/
& 16X,'Gasifier output syngas temp.        ',F12.2,' F ',/
& 16X,'Gasifier output syngas density      ',E12.6,' lbmole/ft3',/
&      'Zinc Ferrite      ',
&      'Zinc Ferrite inlet syngas flow     ',F12.2,' lbmole/hr')
120 FORMAT(
& 16X,'Off-gas to sulfuric acid plant      ',F12.2,' lbmole/hr',/
& 16X,'SO2 to sulfuric acid plant          ',F12.2,' lbmole/hr',/
& 16X,'Off-gas temp. to sul. acid plant    ',F12.2,' lbmole/hr',/
& 16X,'Zinc ferrite inlet H2S flow         ',F12.2,' lbmole/hr',/
& 16X,'Zinc ferrite inlet COS flow        ',F12.2,' lbmole/hr',/
&      'Sulfation         ',
&      'Gasifier ash removal                ',F12.2,' lb/hr',/
& 16X,'Gasifier ash removal                ',F12.2,' lb/hr',/
& 16X,'Gasifier fines removal              ',F12.2,' lb/hr',/
& 16X,'Gasifier fines removal              ',F12.2,' lb/hr')
130 FORMAT(
&      'BFW Treating      ',
&      'Raw water to power plant            ',F12.2,' lb/hr',/
& 16X,'Steam turbine condensate             ',F12.2,' lb/hr',/
&      'Gas Turbine       ',
&      'Gas turbine inlet air flow          ',F12.2,' lb/hr',/
&      'Steam Cycle       ',
&      'HRSG outlet HP steam pres.         ',F12.2,' psia ',/
& 16X,'HRSG outlet HP steam flow           ',F12.2,' lb/hr',/
& 16X,'Gas turbine net shaft work          ',E12.6,' Watts',/
& 16X,'Steam turbine net shaft work        ',E12.6,' Watts',/
& 16X,'Steam cycle auxiliary power         ',E12.6,' Watts',/
& 16X,'Acid gas auxiliary power            ',E12.6,' Watts',/
&      'Miscellaneous     ',
&      'Moisture in coal feed               ',F12.2,' wt %',/
& 16X,'Ash in coal                       ',F12.2,' wt %',/
& 16X,'Temp. of ambient air               ',F12.2,' F ',/
& 16X,'Heating value of coal              ',F12.2,' BTU/lb')
135 FORMAT(
&      'Environmental     ',
&      'CO2 from gas turbine                ',F12.2,' lbmole/hr',/
& 16X,'CO from gas turbine                 ',F12.2,' lbmole/hr',/
& 16X,'SO2 from gas turbine                ',F12.2,' lbmole/hr',/
& 16X,'COS from gas turbine                ',F12.2,' lbmole/hr',/
& 16X,'CH4 from gas turbine                ',F12.2,' lbmole/hr',/
& 16X,'H2S from gas turbine                ',F12.2,' lbmole/hr',/
& 16X,'NH3 to gas turbine                  ',F12.2,' lbmole/hr',/
& 16X,'Actual coal heating value           ',F12.2,' BTU/lb')

```

```

C
C ----- INPUT VARIABLE DEFINITIONS -----
C
C   The variable notation in this subroutine is the same as in Frey and
C   Rubin (1989).
C
C   MCFG1 = Mass flow of coal to gasifier, lb/hr
C   SORB1 = Mass flow of limestone sorbent to gasifier (mixed), lb/hr
C   SORB2 = Mass flow of limestone sorbent to gasifier (cisolid), lb/hr
C   WCME  = Work to boost air compressor, BTU/hr
C   TAPC1I = Temperature of boost air at first precooler inlet, F
C   TAPC1O = Temperature of boost air at first precooler outlet, F
C   TWPC1I = Temperature of BFW at first precooler inlet, F
C   TWPC1O = Temperature of BFW at first precooler outlet, F
C   QAPC1  = Heat transferred from air in first precooler, BTU/hr
C   QW1PC1 = Heat transferred to BFW from HRSG, BTU/hr
C   QW2PC1 = Heat transferred to BFW from regenerator, BTU/hr
C   QW3PC1 = Heat leaving from economizer, BTU/hr
C   PSNZFI = Pressure of syngas at zinc ferrite inlet, psia
C   TSNZFI = Temperature of syngas at zinc ferrite inlet, F
C   VSNZFI = Volume flow of syngas at zinc ferrite inlet, ft3/hr
C   MSNZFI = Molar flow of syngas at zinc ferrite inlet, lbmole/hr
C   MOGSAI = Molar flow of off-gas to sulfuric acid plant, lbmole/hr
C   MSOSAI = Molar flow of SO2 to sulfuric acid plant, lbmole/hr
C   TSOSAI = Temperature of off-gas to sulfuric acid plant, F
C   MHSZFI = Molar flow of hydrogen sulfide (H2S) at zinc ferrite
C           inlet, lbmole/hr
C   MCSZFI = Molar flow of carbonyl sulfide (COS) at zinc ferrite
C           inlet, lbmole/hr
C   MDISP1 = Mass flow (NC) of ash leaving gasifier, lb/hr
C   MDISP2 = Mass flow (CISOLID) of ash leaving gasifier, lb/hr
C   MDISP3 = Mass flow (NC) of fines leaving gasifier, lb/hr
C   MDISP4 = Mass flow (CISOLID) of fines leaving gasifier, lb/hr
C   MRW    = Mass flow of raw water to power plant, lb/hr
C   MCOND  = Mass flow of steam turbine condensation, lb/hr
C   MAGTI  = Mass flow of air to gas turbine inlet, lb/hr
C   PHSHRO = Pressure of high pressure steam at HRSG outlet, psia
C   MHSHRO = Mass flow of high pressure steam at HRSG outlet, lb/hr
C   GTNETS = Gas turbine net shaft work, BTU/hr
C   STNETS = Steam turbines net shaft work, BTU/hr
C   STAUXE = Steam cycle auxiliaries power consumption, BTU/hr
C   ACAUXE = Acid gas removal auxiliaries power consumption, BTU/hr
C   PERH2O = Moisture content of coal feed, weight percent
C   PERASH = Ash content of coal, weight percent
C   TA     = Temperature of ambient air, F
C   HVCOAL = Heating value of coal, BTU/lb
C   MCO2OUT= CO2 leaving gas turbine, lbmole/hr
C   MCOOUT = CO leaving gas turbine, lbmole/hr
C   MSO2OUT= SO2 leaving gas turbine, lbmole/hr
C   MCOSOUT= COS leaving gas turbine, lbmole/hr
C   MCh4OUT= CH4 leaving gas turbine, lbmole/hr
C   MH2SOUT= H2S leaving gas turbine, lbmole/hr
C   MNH3IN = NH3 entering gas turbine, lbmole/hr
C
C ----- COST MODEL PARAMETERS -----
C
C   The following variables are cost model parameters.
C
C   Set the Chemical Engineering Plant Cost Index and Industrial
C   Chemicals Producer Price indicator to January of the cost

```

```

C   year NCSTYR
C
C   NCSTYR=1989
C
C   IF (NCSTYR.EQ.1989) THEN
C       CI=351.5
C       CICPPI=411.25
C   ELSEIF (NCSTYR.EQ.1988) THEN
C       CI=336.3
C       CICPPI=349.9
C   ELSEIF (NCSTYR.EQ.1987) THEN
C       CI=318.3
C       CICPPI=323.9
C   ELSEIF (NCSTYR.EQ.1986) THEN
C       CI=323.5
C       CICPPI=341.7
C   ELSEIF (NCSTYR.EQ.1985) THEN
C       CI=324.7
C       CICPPI=337.7
C   ELSEIF (NCSTYR.EQ.1984) THEN
C       CI=351.5
C       CICPPI=347.4
C   ELSEIF (NCSTYR.EQ.1983) THEN
C       CI=315.5
C       CICPPI=339.9
C   ENDIF
C
C   ----- COST MODEL PARAMETER DEFINITIONS -----
C
C   CF      = Annual plant capacity factor (decimal)
C   EREAL   = Annual real escalation rate (decimal)
C   FBM     = Fraction of byproduct sales to marketing costs (decimal)
C   FEHO    = Engineering and home office fee factor (decimal)
C   FGF     = General facilities cost, fraction of other directs (decimal)
C   FICC    = Indirect Construction Cost factor (decimal)
C   FPJ     = Project Contingency factor (decimal)
C   INTRST  = Cost of capital during construction (decimal)
C   EINFL   = Inflation rate (decimal)
C   NOYEARS = Number of years of construction (years)
C   IBKLIF  = Book life of project (years)
C   RTAX    = Sales tax rate (decimal)
C   ALABOR  = Average labor rate, including burdens ($/hour)
C   NSHIFTS = Number of work shifts per day (including vacations)
C   FCF     = Fixed charge factor (decimal)
C   DISC    = Discount rate (decimal)
C   VCLF    = Variable cost levelization factor (decimal)
C   GENEFF  = Generator Efficiency
C
C   CF=0.65
C   EREAL=0.0
C   FBM=0.10
C   FEHO=0.15
C   FGF=0.15
C   FICC=0.25
C   FPJ=0.20
C   INTRST=0.05
C   EINFL=0.0
C   NOYEARS=4

```

```

IBKLIF=30
RTAX=0.06
ALABOR=19.70
NSHIFTS=4.75
CALL USRFCF (IBKLIF,FCF,DISC)
CALL USRVCLF (EINFL,EREAL,DISC,IBKLIF,VCLF)
INTRST=DISC
GENEFF=0.985

C
C ----- PROCESS CONTINGENCY FACTORS -----
C   The following are process contingency factors for each process area.
C
FPCCH = 0.050
FPCL  = 0.000
FPCOF = 0.050
FPCG  = 0.050
FPCSS = 0.050
FPCCY = 0.100
FPCZF = 0.400
FPCS  = 0.500
FPCSA = 0.060
FPCBF = 0.000
FPCGT = 0.030
FPCHR = 0.030
FPCST = 0.030
FPCGF = 0.100

C
C ----- REGRESSION MODEL ERROR TERMS -----
C   The error terms for the direct cost regression models have "ERR" as the
C   first three letters. The error terms for the auxiliary power consumption
C   models have "ERRW" as the first four letters. Models with negligible
error
C   terms are not included.
C
ERRCH = 0.0
ERROF = 0.0
ERRSA = 0.0
ERRBF = 0.0
ERRHR = 0.0
ERRST = 0.0
ERRWCH = 0.0

C
C ----- MAINTENANCE COST FACTORS -----
C   Maintenance cost factors, based on plant section capital cost
C   (Based on Table 16 in Frey and Rubin (1989))
C
FMCCH = 0.02
FMCL  = 0.03
FMCOF = 0.02
FMCG  = 0.045
FMCSS = 0.03
FMCCY = 0.015
FMCZF = 0.045
FMCS  = 0.04
FMCSA = 0.02
FMCBF = 0.015
FMCCC = 0.015
FMCGF = 0.015

```

```

C
C ----- UNIT COSTS FOR FUEL, CHEMICALS, AND CATALYSTS -----
C   The following are unit costs for fuels, chemicals, and catalysts
C   based on data in Table 17 from Frey and Rubin (1989). The base
C   year for these costs is January, 1989. The costs are adjusted to
C   other years using the Chemical Engineering magazine industrial
C   chemicals producers prices index.
C
C   UCCOAL=1.61*(CICPPI/411.25)
C   UCSAI=110*(CICPPI/411.25)
C   UCNAOH=220*(CICPPI/411.25)
C   UCNA2H=0.70*(CICPPI/411.25)
C   UCHYDR=3.20*(CICPPI/411.25)
C   UCMORP=1.30*(CICPPI/411.25)
C   UCLIME=80*(CICPPI/411.25)
C   UCSODA=160*(CICPPI/411.25)
C   UCCORI=1.90*(CICPPI/411.25)
C   UCSURF=1.25*(CICPPI/411.25)
C   UCCHLR=250*(CICPPI/411.25)
C   UCBIOC=3.60*(CICPPI/411.25)
C   UCSELS=1.80*(CICPPI/411.25)
C   UCCLSC=440*(CICPPI/411.25)
C   UCSACA=1.90*(CICPPI/411.25)
C   UCSCCA=230*(CICPPI/411.25)
C   UCSCCH=0.36*(CICPPI/411.25)
C   UCBSCT=170*(CICPPI/411.25)
C   UCZFSO=3.30*(CICPPI/411.25)
C   UCPLTA=2.80*(CICPPI/411.25)
C   UCFLRL=11.70*(CICPPI/411.25)
C   UCWWCH=840*(CICPPI/411.25)
C   UCFUEL=42*(CICPPI/411.25)
C   UCRAWW=0.73*(CICPPI/411.25)
C   UCLMST=18*(CICPPI/411.25)
C   UCBNTN=0.029*(CICPPI/411.25)
C   UCCOKE=150*(CICPPI/411.25)
C   UCASHD=10*(CICPPI/411.25)
C   UCSSLGD=11.30*(CICPPI/411.25)
C
C ----- UNIT PRICES FOR BYPRODUCTS -----
C   The following are unit PRICES for byproducts, and may differ from
C   unit costs of the same commodities purchased for consumption in
C   the plant
C
C   UPSULF=125*(CICPPI/411.25)
C   UPACID=40*(CICPPI/411.25)
C
C ----- UNIT COST OF LAND -----
C   The following is the cost of land, $/acre
C
C   UCLAND=6500
C
C ----- DIRECT CAPITAL COSTS -----
C   The direct capital costs for 14 process areas are estimated. These
C   include: Coal Handling (CH)           Sulfation (S)
C             Limestone Handling (L)       Sulfuric Acid Plant (SA)
C             Oxidant Feed (OF)           Boiler Feedwater System (BF)
C             Gasification (G)            Gas Turbine (GT)

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C           Subsystems (SS)           Heat Recovery Steam Gen. (HR)
C           Cyclones (CY)            Steam Turbine (ST)
C           Zinc Ferrite (ZF)         General Facilities (GF)
C
C
C Calculate the direct cost for coal handling (Area 20). Coal feed
C rate is lb/hr. Original regression based on tons/day. Coefficient
C adjusted accordingly. The coal handling process area is considered
C to be a single train. A warning message is printed if the coal
C feed rate exceeds the range used to develop the regression model,
C indicating that the model is being extrapolated.
C
C NTCH=1
C NOCH=1
C CALL USRWCK (4.16D+5,MCFGI,3.0D+6,'MCFGI ','DCCH ')
C DCCH = (11200.+0.0204*MCFGI+ERRCH)*(CI/351.5)
C
C Calculate the direct cost for limestone handling (Area 25). Limestone
C feed rate MLGI is in lb/hr.
C
C NOTE: For the base case design, there is no in-bed desulfurization
C and therefore the number of limestone handling AND sulfation trains
C is zero, with capital costs also equal to zero. To include in-bed
C desulfurization in the cost model, simply set NTL = 1. This will turn
C on all the costs associated with these systems, including auxiliary
C power consumption.
C
C NTL=0
C NOL=NTL
C MLGI = SORB1 + SORB2
C DCL = 0.0
C IF (NTL.GT.0) DCL = (1160. + 0.026 * MLGI)*(CI/351.5)
C
C Calculate the direct cost for the oxidant feed section (Area 10).
C The cost includes a boost compressor and a compressor precooler.
C
C First estimate the number of gas turbines based on the compressor
C inlet air flow rate. The number of boost air compressors is equal
C to the number of gas turbines. There are no spares. The number
C of gas turbines is rounded to the nearest integer.
C
C A warning message is printed if the power requirement per train
C for the boost air compressor exceeds the range used to develop
C the regression model. If such a message is printed, the model
C is being extrapolated and the user should interpret the results
C with care.
C
C NOGT = INT(MAGTI/3000000.0+0.5)
C IF (NOGT.EQ.0) NOGT=1
C NTGT=NOGT
C
C NOOF=NOGT
C NTOF=NTGT
C WCME=WCME/1000.
C CALL USRWCK (1.9D+3,(WCME/NOOF),8.2D+3,'WCME/N','DCCM ')
C DCCM= (-98.0 + 26. * NTOF * (WCME/NOOF)**0.485)
C
C UPC = Universal heat transfer coefficient for precooler, Btu/(ft2-hr-F)

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```

C      DTLMPC = Log mean temperature difference for precoolers, F
C      APC      = Heat transfer area of precoolers, ft2
C
C      The number of precoolers is the same as the number of boost air
C      compressors. A warning message is printed if the heat exchanger
C      area exceeds the range used to develop the model, indicating
C      that the model is being extrapolated.
C
      UPC=40.0
      TWPC10 = TWPC1I + (TWPC10-TWPC1I)*(QAPC1/(QAPC1+QW1PC1+
&      QW2PC1-QW3PC1))
      TLMPC1=200.0
      IF(((TAPC1I-TWPC10).LT.0).AND.((TAPC10-TWPC1I).LT.0)) THEN
          TLMPC1 = ((TAPC1I-TWPC10)-(TAPC10-TWPC1I))/
&      LOG((TAPC1I-TWPC10)/(TAPC10-TWPC1I))
      ELSEIF(((TAPC1I-TWPC10).GT.0).AND.((TAPC10-TWPC1I).GT.0)) THEN
          TLMPC1 = ((TAPC1I-TWPC10)-(TAPC10-TWPC1I))/
&      LOG((TAPC1I-TWPC10)/(TAPC10-TWPC1I))
      ENDIF
      APC1= QAPC1/(UPC*TLMPC1*NOOF)
      CALL USRWCK (0.0D+0,APC1,1.0D+4,'APC1/N','DCPC1 ')
      DCPC1 = (24.+0.067*APC1-5.5E-6*APC1**2+2.8E-10*APC1**3)*NTOF
C
      DCOF= (DCCM + DCPC1 + ERROF)*(CI/351.5)
C      WRITE(NRPT,700) DCCM, DCPC1, DCOF, WCME, NOOF, NTOF, TLMPC1, APC1
C 700 FORMAT(' DCCM - DCPC - DCOF - WCME - NOOF - NTOF - TLMPC1 - APC1',
C      &      /3F10.2,1X,E12.6,2I5,F10.2,F10.2)
C
C      Direct capital cost of the Gasification Section (Area 30). For the
C      Lurgi-based system, the cost of the gasifiers is estimated separately
C      from the costs for coke, fines, and ash handling subsystems and for
C      the high temperature cyclones.
C
C      The number of trains is estimated based on the coal throughput
C      per train. The value of coal throughput is estimated based on
C      the gasifier pressure and assuming an Illinois No. 6 coal. The
C      use of the model for pressures outside the 300 to 600 psia range
C      used to determine the coal throughput, or for other than an
C      Illinois No. 6 coal is not recommended. The number of operating
C      gasifiers is based on the total coal throughput divided by the
C      estimated coal throughput per gasifier, rounded to the nearest
C      integer. The number of total gasifiers is estimated by dividing
C      the number of operating gasifiers by the availability of each
C      gasifier, rounded to the nearest integer.
C
C      AVLG = Availability of the gasifiers
C
      AVLG=0.87
      PERH2O=PERH2O/100.
      PERASH=PERASH/100.
      MCMGI = MCFGI*(1-PERH2O-PERASH)
      CTP = (100. + 0.6666667*PSNZFI)*127.
      NOG = INT(MCFGI/CTP + 0.5)
      NTG = INT(REAL(NOG)/AVLG + 0.5)
C      Cost of coke, fines, and ash handling subsystem
C      A warning message is printed if the coal mass flow rate exceeds the
C      limits used to develop the regression model. This model is applicable
C      only to Illinois No. 6 coal.

```



```

      CALL USRWCK (83.0D+3,MCFGI,433.0D+3,'MCFGI ','DCSS ')
      DCSS = (-1200. + 0.030*MCFGI)*(CI/351.5)
C   Cost of cyclones subsystem. A warning message is printed if either
C   the pressure or volume flow rate per cyclone exceeds the range of
C   values used to develop the regression model.
      NOCY=2*NOG
      NTCY=2*NTG
      VSNZFI=MSNZFI/(DSNZFI*60.)
      CALL USRWCK (0.260D+3,PSNZFI,0.630D+3,'PSNZFI','DCCY ')
      CALL USRWCK (3.9D+3,(2*VSNZFI/NOCY),6.0D+3,'VSNZFI','DCCY ')
      DCCY = 0.98*NTCY*(PSNZFI**0.28)*(2*VSNZFI/NOCY)**0.43
      &      *(CI/351.5)
C
C   Cost of gasifier vessels. Includes a pressure correction term
C   on the first line of the equation. The reference pressure is
C   450 psia. The pressure correction term is taken from the
C   pressure correction equation developed for the zinc ferrite system.
C
      DCG = ((2.5+0.093*PSNZFI**0.64)/7.14)*
      &      (7700.*NTG)*(CI/351.5)
C
C   Direct capital cost of zinc ferrite desulfurization
C
C   This model estimates the cost of a fixed-bed system. The number of
C   trains is adjusted so that each reactor vessel is rail-shippable.
C   For each operating vessel, assumed to be in absorption mode, there
C   is a vessel in regeneration mode. As a base case, it is assumed that
C   there are two absorbing vessels per gasifier. The number of vessels
C   is adjusted to conform to rail-shippable sizes and to keep the
C   superficial velocity below VSZF.
C   A warning message is printed if the vessel diameter exceeds the
C   rail shippable upper limit (DMAX) or the minimum diameter DZFMIN
C   needed to assure that the superficial velocity is less than or
C   equal to VSZF.
C
C   NMZF   = Number of operating vessels per operating gasifier
C   DMAX   = Maximum rail-shippable diameter, ft
C   TAZF   = Time of absorption cycle in zinc ferrite section, hours
C   SLZF   = Sorbent loading, long term, in zinc ferrite, wt fraction
C   DBZF   = Bulk density of zinc ferrite sorbent, lb/ft2
C   VSZF   = Superficial velocity, ft/sec
C
C   NMZF   = 1
C   ILOOP  = 0
C   JLOOP  = 0
C   DMAX   = 12.5
C   TAZF   = 168.0
C   SLZF   = 0.10
C   DBZF   = 82.0
C   VSZF   = 2.0
C   MSZFI  = (MHSZFI+ MCSZFI)*TAZF
C   SCZF   = (32-10.666667*SLZF)*MSZFI/SLZF
800 CONTINUE
      IF(NMZF.LT.1) NMZF = 1
      NOZF  = NMZF*NOG
      NTZF  = 2*NMZF*NTG
      VCZF  = SCZF/(DBZF*NOZF)
      DZF   = (VCZF/3.141592654)**(1/3.0)

```

```

DZFMIN = 1.128*( 10.73*MSNZFI*(TSNZFI+460)
& / (3600.*VSZF*PSNZFI*NOZF) )**0.5
IF (DZF.LT.DZFMIN) THEN
  ILOOP = ILOOP+1
  NMZF = NMZF-1
  IF (NMZF.LE.1) THEN
    NMZF=1
    JLOOP=9
  ENDIF
  IF (ILOOP.LT.10) GOTO 800
ELSEIF (DZF.GT.DMAX) THEN
  JLOOP = JLOOP+1
  NMZF = NMZF+1
  IF (JLOOP.LT.10) GOTO 800
ENDIF
CALL USRWCK (DZFMIN,DZF,DMAX,'DZF ','DCZF ')
HZF = (4*VCZF)/(3.141592654*DZF**2)
AZF = 4.1002-4.3468*DZF+0.95848*DZF**2-0.039117*DZF**3
BZF = (0.76181+0.5505*DZF-0.059669*DZF**2+0.00258088*DZF**3)/
& (3.28084)
FPIZF = 2.5 + 0.093*(PSNZFI**0.64)
DCZF = 1477.*FPIZF*(AZF+BZF*HZF)*NTZF*(CI/351.5)/1000.

```

C Direct capital cost of the sulfation section
 C NOTE: See comment for limestone handling direct capital cost.
 C Only one train should be used if the sulfation unit is included.
 C

```

NOS=NTL
NTS=NOS
DCS = 0.0
IF (NTS.GT.0) DCS = 13.0*(MDISP1+MDISP2)**0.639*(CI/351.5)

```

C Direct capital cost of the sulfuric acid plant
 C Only one train is assumed, with no spare. Warning messages are
 C printed if the concentration of sulfur dioxide in the zinc ferrite
 C offgas exceeds the limits used to develop the model, or if the
 C offgas molar flow rate exceeds the limits used to develop the model.
 C If the offgas temperature exceeds the limits, it is reset to the
 C nearest limit, and a warning is printed.
 C

```

NOSA=1
NTSA=1
FSO2 = MSOSAI/MOGSAI
CALL USRWCK (0.01D+0,FSO2,0.13D+0,'FSO2 ','DCSA ')
CALL USRWCK (1.48D+3,MOGSAI,70.0D+3,'MOGSAI','DCSA ')
CALL USRRCK (0.500D+3,TOGSAI,1.500D+3,'TOGSAI','DCSA ')
DCSA = (52 + 18.2*(MOGSAI**0.67)*(FSO2**0.12)*(TOGSAI**0.12)
& + ERRSA)*(CI/351.5)

```

C Direct capital cost of the boiler feedwater system
 C

C Raw water is makeup water for the steam cycle
 C Polished water includes steam turbine condensate and makeup water.
 C This process area is considered a single train. Warnings are
 C printed if the flow rates of either raw or polished water exceed
 C the limits used to develop the model.
 C

```

NOBF=1

```

```

NTBF=1
MPW=MRW+MCOND
CALL USRWCK ( 24.000D+3,MRW,0.614000D+6,'MRW ','DCBF ')
CALL USRWCK (0.234000D+6,MPW,3.880000D+6,'MPW ','DCBF ')
DCBF=(0.145*MRW**0.307*MPW**0.435+ ERRBF )*(CI/351.5)

C
C Direct capital cost of the gas turbines
C
C The gross gas turbine electrical output is estimated in MW.
C The number of units has been previously estimated based on
C input air flow.
C
WGTE=-GENEFF*GTNETS/1000000
DCGT=32000*NTGT*(CI/351.5)

C
C Direct capital cost of heat recovery steam generator
C The number of HRSGs is the same as the number of gas turbines.
C A warning is printed if the steam flow rate per HRSG exceeds
C the range used to develop the model. If the pressure exceeds
C the range used to develop the model, it is reset to either the
C upper or lower limit and a warning is printed.
C
NOHR=NOGT
NTHR=NTGT
CALL USRRCK ( 0.650D+3,PHSHRO,1.545D+3,'PHSHRO','DCHR ')
CALL USRWCK ( 66.000D+3,(MHSIRO/NOHR),0.640000D+6,
& 'MHSI/N','DCHR ')
DCHR=(-5364+0.00724*NTHR*PHSHRO**1.526*(MHSIRO/NOHR)**0.242+ERRHR)
& *(CI/351.5)

C
C Direct capital cost of the steam turbine
C
C The gross electrical output of the steam turbine is estimated in MW.
C Only one steam turbine is required.
C
NOST=1
NTST=1
WSTE=-GENEFF*STNETS/1000000
DCST=(158.7*(WSTE)+ERRST)*(CI/351.5)

C
C Direct capital cost of general facilities
C
DCGF=FGF*(DCCH+DCL +DCOF+DCG +DCSS+DCCY+DCZF+DCSA+DCS +DCBF
& +DCGT+DCHR+DCST)

C
C ----- CAPITAL COSTS: TOTAL PLANT INVESTMENT -----
C
C The total plant investment is estimated in this section. The
C total capital requirement, which includes items dependent on
C operating costs, is not calculated until after the operating
C costs are estimated.
C
C Total Direct Cost (TDC)
C
TDC= DCCH+DCL +DCOF+DCG +DCSS+DCCY+DCZF+DCSA+DCS +DCBF
& +DCGT+DCHR+DCST+DCGF

C
C Total Indirect Cost (TIC)

```

```

C      CICC=FICC*TDC
      CTAX=RTAX*(0.8*TDC+0.1*CICC)
      CEHO=FEHO*(TDC+CICC+CTAX)
      CEP=1000
      TIC=CICC+CTAX+CEHO+CEP
C
C      Process Contingency Costs
C
      CPCCH = FPCCH*DCCH*(1.+TIC/TDC)
      CPCL  = FPCL *DCL *(1.+TIC/TDC)
      CPCOF = FPCOF*DCOF*(1.+TIC/TDC)
      CPCSS = FPCSS*DCSS*(1.+TIC/TDC)
      CPCCY = FPCCY*DCCY*(1.+TIC/TDC)
      CPCG  = FPCG *DCG *(1.+TIC/TDC)
      CPCZF = FPCZF*DCZF*(1.+TIC/TDC)
      CPCSA = FPCSA*DCSA*(1.+TIC/TDC)
      CPCS  = FPCS *DCS *(1.+TIC/TDC)
      CPCBF = FPCBF*DCBF*(1.+TIC/TDC)
      CPCGT = FPCGT*DCGT*(1.+TIC/TDC)
      CPCHR = FPCHR*DCHR*(1.+TIC/TDC)
      CPCST = FPCST*DCST*(1.+TIC/TDC)
      CPCGF = FPCGF*DCGF*(1.+TIC/TDC)
C
      CPC=CPCCH+CPCL+CPCOF+CPCSS+CPCCY+CPCG+CPCS+CPCZF+CPCSA
      & +CPCBF+CPCGT+CPCHR+CPCST+CPCGF
C
C      Project contingency cost
C
      CPJ=FPJ*(TDC+TIC+CPC)
C
C      Total Plant Cost (TPC)
C
      TPC=TDC+TIC+CPC+CPJ
C
C      Total plant investment (TPI)
C
      Z=(1+INTRST)/(1+EREAL)
      AF=(Z**NOYEARS-1)/(NOYEARS*(Z-1))
      TPI=AF*TPC
      AFDC=TPI-TPC
C
C      ----- NET PLANT ELECTRICAL OUTPUT -----
C
C      The net electrical output must be estimated 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      oxygen plant, the gasification section, limestone handling,
C      sulfation, general facilities, and power consumption in the
C      steam cycle. The pump and compressor power consumption
C      is estimated directly from the ASPEN simulation. The other areas
C      are estimated from regression relationships taken from Frey and Rubin
C      (1989).
C
C      Gas Turbine Output:  WGTE          Steam Turbine Output:  WSTE
C
C      The auxiliary loads have been converted to MW from Frey and Rubin (1989)
C

```

```

WCHE = 1.89D-06 * MCFGI + ERRWCH
WLE = 0.0
IF(NTL.GT.0) WLE = (-13.3 + 0.00461*MLGI)/1000.
WOFE = WCME/1000.
WGE = (0.005*MCFGI)/1000.
WSE = 0.0
IF(NTS.GT.0) WSE = (-3.7 + 0.0102*(MDISP1+MDISP2))/1000.
C   WSAE = (0.9/1000.0) * MOGSAI
WSAE = ACAUXE/1000000.0
WBFE = STAUXE/1000000.0
WGFE = 1.64 + 0.0877*(WCHE+WLE+WOFE+WGE+WSE+WBFE)
WAUXE = WCHE+WLE+WOFE+WGE+WSE+WSAE+WBFE+WGFE
WNETE = WGTE+WSTE-WAUXE

C
C ----- INITIAL CATALYSTS AND CHEMICALS -----
C
C Zinc Ferrite Sorbent: the total sorbent charge is:
C
C   TSCZF = (REAL(NTZF)/REAL(NOZF)) * SCZF
C
C Initial startup fuel
C
C   FUELI= 80000*(WNETE/550.)
C
C Cost of initial makeup cooling water treatment chemicals
C
C   CCHICW=1000*(CICPPI/337.7)
C
C Startup coke and bentonite
C NOTE: REPLACE MDISP3+MDISP4 WITH MRECYC
C
C   CICOKE = (200.0)*(WNETE/180.)
C   CIBENT = 5.88 * (RECYC)
C
C Total cost of initial catalyst and chemicals
C (These costs are in dollars, and must be divided by 1,000 for
C conversion to thousands of dollars)
C
C   TCICC=(UCFUEL*FUELI+CCHICW+UCZF*SO*TSCZF+
C   &   UCCOKE*CICOKE+UCBNTN*CIBENT)/1000.
C   WRITE(NRPT,760) TSCZF
C 760 FORMAT(' TOTAL SORBENT CHARGE ',E14.7)
C
C ----- FIXED OPERATING COSTS -----
C
C Fixed Operating Costs: Operating Labor
C
C   NOPERS=15+2*NOG+4*NOGT
C   OCL=ALABOR*2080*NSHIFTS*NOPERS
C
C Fixed Operating Costs: Maintenance
C
C   OCMOF = FMCOF * (DCOF * (1+(TIC+CPJ)/TDC) + CPCOF)
C   OCMCH = FMCCH * (DCCH * (1+(TIC+CPJ)/TDC) + CPCCH)
C   OCML = FMCL * (DCL * (1+(TIC+CPJ)/TDC) + CPCL)
C   OCMG = FMCG * (DCG * (1+(TIC+CPJ)/TDC) + CPCG)
C   OCMSS = FMCSS * (DCSS * (1+(TIC+CPJ)/TDC) + CPCSS)
C   OCMCY = FMCCY * (DCCY * (1+(TIC+CPJ)/TDC) + CPCCY)

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OCMSA = FMCSA * (DCSA * (1+(TIC+CPJ)/TDC) + CPCS)
OCMS = FMCS * (DCS * (1+(TIC+CPJ)/TDC) + CPCS)
OCMZ = FMCZF * (DCZF * (1+(TIC+CPJ)/TDC) + CPCZF)
OCMBF = FMCBF * (DCBF * (1+(TIC+CPJ)/TDC) + CPCBF)
OCMCC = FMCCC * ((DCGT+DCHR+DCST) *
& (1+(TIC+CPJ)/TDC)+CPCGT+CPCHR + CPCST)
OCMGF = FMCGF * (DCGF * (1+(TIC+CPJ)/TDC) + CPCGF)

OCM = (OCMOF+OCMCH+OCML+OCMG+OCMS+OCMZ+OCMBF+
& OCMSS+OCMCY+OCMSA+OCMCC+OCMGF)*1000
OCMM=0.60*OCM
OCML=0.40*OCM

Fixed Operating Costs: Administration and Supervision

OCAS=0.30*(OCL+OCML)

Total fixed operating costs

FOC=OCL+OCM+OCAS

----- VARIABLE OPERATING COSTS -----

Fuel Cost for Coal (UCCOAL in $/MMBtu)

QCOAL=MCFGI*HVCOAL
OCFUEL=CF*8760*QCOAL*UCCOAL/1E+06

Material requirements for consumables, and cost of consumables

Raw and polished water treatment chemical requirements

MSABFI=CF*(47.0+0.00209*MRW)
MSHBFI=CF*(9.5+0.00042*MRW)
MSPBFI=CF*(115+0.00361*MRW)
MHYBFI=CF*(529+0.0174*MRW)
MMOBFI=CF*(420+0.0163*MRW)

CALL USRWCK (1.2D+6,MPW ,2.2D+6,'MPW ', 'MSABFP')
MSABFPI=CF*(15+5.4E-05*MPW)
MSHBFPFI=CF*(30+0.000107*MPW)

The makeup cooling water requirement MCWI (gal/min) is estimated
based on the plant electrical output

MCWI=9.3763*WNETE - 273.8
CALL USRWCK (0.415D+3,MCWI ,7.700D+3,'MCWI ', 'M**CWI')

Cooling water treatment chemical requirements

MLMCWI=CF*(-26+0.143*MCWI)
MSOCWI=0.154*CF*MCWI
MSACWI=CF*(6+0.147*MCWI)
MCICWI=CF*(-7280+28.8*MCWI)
MSUCWI=MCICWI
MCLCWI=CF*(4.1+0.00374*MCWI)
MBIOCWI=CF*(4800+4.12*MCWI)

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```

C   Makeup Zinc Ferrite Sorbent Requirement
C
C       FATTZF = 0.20
C       MSOZFI = 8760.*CF*FATTZF*SCZF/(2*TAZF*80)
C
C   Makeup sulfuric acid plant catalyst, annual coke and bentonite
C
C       MSACAT = 1.1*MOGSAI
C       MCOKE = 610.*CF*(WNETE/180.)
C   NOTE:  REPLACE MDISP3+MDISP4 WITH MRECYC
C       MBENT = 0.043*8760*CF*(RECYC)
C
C   Annual Limestone Requirement (TON/YR)
C
C       MLMANN = 0.0
C       IF (NTL.GT.0) MLMANN = MLGI*8760*CF/2000.
C
C   Other consumables:  fuel oil, plant and instrument air adsorbent,
C   wastewater chemicals (COST, not mass basis), plant raw water
C   consumption, and liquified petroleum gas.
C
C   For wastewater treatment, the waste water flow rate is converted from
C   lb/hr to gpm, using 60 min/hr, 62.4 lb/ft3, and 7.48 gal/ft3.
C
C       MFOGTI=48000*CF*(WNETE/550)
C       MADSGFI=3600*CF*(WNETE/550)
C       WWATER=0.0
C       CWW = CF * UCWWCH * WWATER * 1/(60.0 * 62.4) * 7.48
C
C   Only raw water input to the plant (process and deaerator makeup)
C   is included in water costs.
C   Cooling water is assumed to be taken from nearby body of water
C
C       MWATER=CF*8760*(MRW*1/62.4*7.48)/1000
C       MLPGI=4200*CF*(WNETE/550)
C
C   Estimate of consumable operating costs
C
C       MSAC = MSABFI + MSABFPI + MSACWI
C       OCSAI = UCSAI * MSAC
C       MNAOH = MSHBFI + MSHBFPI
C       OCNAOH = UCNAOH * MNAOH
C       OCNA2H = UCNA2H * MSPBFI
C       OCHYDR = UCHYDR * MHYBFI
C       OCMORP = UCMORP * MMOBFI
C       OCLIME = UCLIME * MLMCWI
C       OCSODA = UCSODA * MSOCWI
C       OCCORI = UCCORI * MCICWI
C       OCSURF = UCSURF * MSUCWI
C       OCCHLR = UCCHLR * MCLCWI
C       OCBIOC = UCBIOC * MBIOCWI
C       OCZFSO = UCZFSO * MSOZFI
C       OCLMST = UCLMST * MLMANN
C       OCSACT = UCSACA * MSACAT
C       OCCOKE = UCCOKE * MCOKE
C       OCBENT = UCBNTN * MBENT
C       OCFOIL = UCFUEL * MFOGTI
C       OCPLTA = UCPLTA * MADSGFI

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OCRAWW = UCRAWW * MWATER
OCFLRL = UCFLRL * MLPGI

C
OCCONS = OCSAI + OCNAOH + OCNA2H + OCHYDR + OCMORP + OCLIME +
&          OCSODA + OCCORI + OCSURF + OCCHLR + OCBIOC + OCZFSSO +
&          OCLMST + OCFOIL + OCPLTA + OCRAWW +
&          CWW      + OCFLRL + OCSACT + OCCOKE + OCBENT

C
C Costs for ash disposal and byproduct credit
C
MACID = MSOSAI*98.0/(0.93*2000.)*CF*8760
OCBYP = (1.0-FBM)*MACID*UPACID
MASH = (MDISP1+MDISP2+MDISP3+MDISP4)*24./2000.
OCASH = UCASHD * (MASH) * CF * 365.0

C
C TOTAL VARIABLE OPERATING COST (No byproduct sales)
C
VOC = OCFUEL + OCCONS + OCASH - OCBYP

C
C ----- PREPRODUCTION CAPITAL COSTS -----
C
PPFC = FOC/(12000)
PPOC = (0.083/CF)*(OCCONS+OCASH)/1000
PPFUEL=(0.021/CF)*OCFUEL/1000
PPC=PPFC+PPOC+PPFUEL+0.02*TPI

C
C ----- Inventory Capital Costs -----
C
IC=(0.164/CF)*(OCFUEL+OCCONS)/1000

C
C ----- Total Capital Requirement (TCR) -----
C
ALAND=-93+0.065*QCOAL/1000000
TCLAND=ALAND*UCLAND/1000
TCR=1.01*TPI+PPC+IC+TCICC+TCLAND

C
C ----- Normalized Costs and Cost of Electricity -----
C
DPERKW = TCR/WNETE
FOCN = FOC/(WNETE*1000.)
VOCN = VOC/(WNETE*8760.*CF)
FUELN = OCFUEL/(WNETE*8760.*CF)
BYPN = OCBYP/(WNETE*8760.*CF)
VOCINC = VOCN-FUELN+BYPN
CELEC=(1000*FCF*TCR+VCLF*(FOC+VOC))/(WNETE*8760*CF)

C
C ----- Plant Efficiency -----
C
HEATRATE=MCFGI*HOC/(WNETE*1000.)
EFFICNCY=3414.98/HEATRATE

C
C ----- Environmental Summary -----
C
MCOALIN = MCFGI/(WNETE*1000.)
MH2OIN = MRW/(WNETE*1000.)
MASHOUT = MASH*(2000.0/24.0) / (WNETE*1000.)
MACIDO = MSOSAI*(98.0/0.93) / (WNETE*1000.)
ECO2 = MCO2OUT*44.0/(WNETE*1000.)

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ECO = MCOOUT*28.0/(WNETE*1000.)
C
C Sulfur dioxide emissions include 0.5 % of sulfuric acid plant
C inlet stream.
C
ESO2 = (MSO2OUT+MSOSAI*0.005)*64.0/(MCFGI*HOC/1.D+6)
ECOS = MCOSOUT*60.0/(MCFGI*HOC/1.D+6)
ECH4 = MCH4OUT*16.0/(WNETE*1000.)
EH2S = MH2SOUT*34.0/(MCFGI*HOC/1.D+6)
ENOX = 0.7*MNH3IN*30.8/(MCFGI*HOC/1D+6)
C
C ----- COST SUMMARY REPORT -----
C
WRITE(NRPT,200)
200 FORMAT(33X,'COST SUMMARY'
& /11X,
& 'Air Blown Lurgi-Based IGCC System with Hot Gas Cleanup')
C
C ----- Report Cost Model Parameters -----
C
WRITE(NRPT,210)
210 FORMAT(/
& 'A. COST MODEL PARAMETERS -----',
& '-----')
WRITE(NRPT,220) CF, NCSTYR, FGF, CI, FICC, CICPPI,
& RTAX, EREAL, FEHO, INTRST, FPJ, NOYEARS,
& NSHIFTS, FBM, FCF, ALABOR, VCLF, IBKLIF
220 FORMAT(' Plant Capacity Factor: ',F6.2,
& ' Cost Year: January',I5,
& /' General Facilities Factor: ',F6.2,
& ' Plant Cost Index: ',F6.1,
& /' Indirect Construction: ',F6.2,
& ' Chemicals Cost Index: ',F6.1,
& /' Sales Tax: ',F6.2,
& ' Escalation: ',F6.2,
& /' Engr & Home Office Fee: ',F6.2,
& ' Interest: ',F6.2,
& /' Project Contingency: ',F6.2,
& ' Years of construction: ',I6,
& /' Number of Shifts: ',F6.2,
& ' Byproduct marketing: ',F6.2,
& /' Fixed Charge Factor: ',F6.4,
& ' Average Labor Rate: ',F6.2,
& /'Variable Levelization Cost Factor: ',F6.4,
& ' Book Life (years): ',I6)
C
WRITE(NRPT,230)
230 FORMAT(/'B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS ',
& '-----',
& /26X' Process Maintenance',
& /26X'Plant Section Contingency Cost Factor',
& /26X'-----')
WRITE(NRPT,240) FPCCH, FMCCH, FPCL, FMCL, FPCOF, FMCOF,
& FPCG, FMCG, FPCSS, FMCSS, FPCCY, FMCCY,
& FPCZF, FMCZF, FPCSA, FMCSA, FPCS, FMCS,
& FPCBF, FMCBF, FPCGT, FMCCC, FPCHR, FMCCC,
& FPCST, FMCCC, FPCGF, FMCGF

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240 FORMAT(10X,'          Coal Handling ',10X,F6.3,10X,F6.3,
&      /10X,'          Limestone Handling ',10X,F6.3,10X,F6.3,
&      /10X,'          Oxidant Feed ',10X,F6.3,10X,F6.3,
&      /10X,'          Gasification ',10X,F6.3,10X,F6.3,
&      /10X,'Coke, Ash, & Bent. Subsystems ',10X,F6.3,10X,F6.3,
&      /10X,'          High Temp. Cyclones ',10X,F6.3,10X,F6.3,
&      /10X,'          Zinc Ferrite ',10X,F6.3,10X,F6.3,
&      /10X,'          Sulfuric Acid Plant ',10X,F6.3,10X,F6.3,
&      /10X,'          Sulfation ',10X,F6.3,10X,F6.3,
&      /10X,'    Boiler Feedwater Treatment ',10X,F6.3,10X,F6.3,
&      /10X,'          Gas Turbine ',10X,F6.3,10X,F6.3,
&      /10X,'Heat Recovery Steam Generator ',10X,F6.3,10X,F6.3,
&      /10X,'          Steam Turbine ',10X,F6.3,10X,F6.3,
&      /10X,'          General Facilities ',10X,F6.3,10X,F6.3)

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----- CAPITAL COSTS -----

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WRITE(NRPT,250)
250 FORMAT('/C.  DIRECT CAPITAL AND PROCESS ',
&      'CONTINGENCY COSTS ($1,000) ',
&      '-----',
&      /16X'          Number of Units          Direct          Process',
&      /16X'Plant Section Operating Total Capital Cost Contingency',
&      /16X'-----'-----'-----'-----')
WRITE(NRPT,260) NOCH, NTCH, DCCH, CPCCH, NOL, NTL, DCL, CPCL,
&      NOOF, NTOF, DCOF, CPCOF, NOG, NTG, DCG, CPCG,
&      NOG, NTG, DCSS, CPCSS, NOCY, NTCY, DCCY, CPCCY,
&      NOZF, NTZF, DCZF, CPCZF, NOSA, NTSA, DCSA, CPCSA,
&      NOS, NTS, DCS, CPCS,
&      NOBF, NTBFB, DCBFB, CPCBFB, NOGT, NTGT, DCGT, CPCGT,
&      NOHR, NTHR, DCHR, CPCHR, NOST, NTST, DCST, CPCST,
&      DCGF, CPCGF
260 FORMAT(
&      /'          Coal Handling ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Limestone Handling ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Oxidant Feed ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Gasification ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'Coke, Ash, & Bent. Subsystems ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          High Temp. Cyclones ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Zinc Ferrite ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Sulfuric Acid Plant ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Sulfation ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'    Boiler Feedwater Treatment ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Gas Turbine ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'Heat Recovery Steam Generator ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          Steam Turbine ',5X,I5,1X,I5,3X,F10.0,3X,F10.0,
&      /'          General Facilities ',5X,' N/A',1X,' N/A',
&      3X,F10.0,3X,F10.0)
WRITE(NRPT,270) TDC, CICC, CTAX, CEHO, CEP, TIC, CPC, CPJ, TPC,
&      AFDC, TPI, PPC, IC, TCICC, TCLAND, TCR
270 FORMAT('/D.  TOTAL CAPITAL REQUIREMENT ($1,000) ----',
&      '-----',
&      /10X,'          Description          '
&      ,8X,'Annual Cost',
&      /10X,'          -----          '
&      ,8X,'-----',
&      /10X,'          Total Direct Cost          ',8X,F10.0,
&      /10X,'          Indirect Construction Cost          ',8X,F10.0,

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& /10X,' Sales Tax ',8X,F10.0,
& /10X,' Engineering and Home Office Fees ',8X,F10.0,
& /10X,' Environmental Permitting ',8X,F10.0,
& /10X,' Total Indirect Costs ',8X,F10.0,
& /10X,' Total Process Contingencies ',8X,F10.0,
& /10X,' Project Contingency ',8X,F10.0,
& /10X,' Total Plant Cost ',8X,F10.0,
& /10X,' AFDC ',8X,F10.0,
& /10X,' Total Plant Investment ',8X,F10.0,
& /10X,' Preproduction (Startup) Costs ',8X,F10.0,
& /10X,' Inventory Capital ',8X,F10.0,
& /10X,' Initial Catalysts and Chemicals ',8X,F10.0,
& /10X,' Land ',8X,F10.0,
& /10X,'TOTAL CAPITAL REQUIREMENT ($1,000) ----->',8X,F10.0)

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----- FIXED OPERATING COSTS -----

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WRITE(NRPT,280) OCL, OCM, OCAS, FOC
280 FORMAT('/E. FIXED OPERATING COSTS ($/year) -----',
& '-----',
& /10X,' Description ',15X,'Annual Cost',
& /10X,' ----- ',15X,'-----',
& /10X,' Operating Labor ',11X,F11.0,
& /10X,' Maintenance Costs ',11X,F11.0,
& /10X,' Administration and Supervision',11X,F11.0,
& /10X,'TOTAL FIXED OPERATING COST ',
& '($/year) ----->',F11.0)

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----- Unit Costs for Fuel, Chemicals, and Catalysts -----

```

WRITE(NRPT,290) UCSAI, MSAC, OCSAI,
& UCNAOH, MNAOH, OCNAOH, UCNA2H, MSPBFI, OCNA2H,
& UCHYDR, MHYBFI, OCHYDR, UCMORP, MMOBFI, OCMORP,
& UCLIME, MLMCWI, OCLIME, UCSODA, MSOCWI, OCSODA,
& UCCORI, MCICWI, OCCORI
290 FORMAT (/
& 'F. VARIABLE OPERATING COSTS -----',
& '-----',
& /' 1. CONSUMABLES ($/year)',
& /6X,' ',
& 'Material Annual',
& /6X,'Description Unit Cost ',
& 'Requirement Operating Cost',
& /6X,'-----',
& '-----',
& /6X,' Sulfuric Acid: ',F7.2,' $/ton ',F10.1,' ton/yr ',F14.0,
& /6X,' NaOH: ',F7.2,' $/ton ',F10.1,' ton/yr ',F14.0,
& /6X,' Na2 HPO4: ',F7.2,' $/lb ',F10.1,' lb/yr ',F14.0,
& /6X,' Hydrazine: ',F7.2,' $/lb ',F10.1,' lb/yr ',F14.0,
& /6X,' Morpholine: ',F7.2,' $/lb ',F10.1,' lb/yr ',F14.0,
& /6X,' Lime: ',F7.2,' $/ton ',F10.1,' ton/yr ',F14.0,
& /6X,' Soda Ash: ',F7.2,' $/ton ',F10.1,' ton/yr ',F14.0,
& /6X,'Corrosion Inh.: ',F7.2,' $/lb ',F10.1,' lb/yr ',F14.0)
WRITE(NRPT,300) UCSURF, MSUCWI, OCSURF, UCCHLR, MCLCWI, OCCHLR,
& UCBIOC, MBIOCWI, OCBIOC, UCZFSO, MSOZFI, OCZFSO,
& UCLMST, MLMANN, OCLMST, UCSACA, MSACAT, OCSACT,
& UCCOKE, MCOKE, OCCOKE, UCBNTN, MBENT, OCBENT
300 FORMAT (

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& 6X,'      Surfactant: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Chlorine: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'      Biocide: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Zinc Fer Sorb: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Limestone: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'      Sul Acid. Cat.: ',F7.2,' $/l      ',F10.1,' l/yr      ',F14.0,
& /6X,'      Coke: ',F7.2,' $/ton      ',F10.1,' ton/yr      ',F14.0,
& /6X,'      Bentonite: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0)
WRITE(NRPT,310) UCFUEL, MFOGTI, OCFOIL, UCPLTA, MADSGFI, OCPLTA,
& UCRAWW, MWATER, OCRAWW, UCWWCH, WWATER, CWW,
& UCFLRL, MLPGI, OCFLRL, OCONNS
310 FORMAT(
& 6X,'      Fuel Oil: ',F7.2,' $/bbl      ',F10.1,' bbl/yr      ',F14.0,
& /6X,'      Plant Air Ads.: ',F7.2,' $/lb      ',F10.1,' lb/yr      ',F14.0,
& /6X,'      Raw Water: ',F7.2,' $/Kgal      ',F10.1,' Kgal/yr      ',F14.0,
& /6X,'      Waste Water: ',F7.2,' $/gpm ww',F10.1,' lb/hr      ',F14.0,
& /6X,'      LPG - Flare: ',F7.2,' $/bbl      ',F10.1,' bbl/yr      ',F14.0,
& /6X,'      TOTAL CONSUMABLES ($/year) ----->',F14.0)
WRITE(NRPT,320) UCCOAL, MCFGI, OCFUEL, UCASHD,
& MASH, OCASH, UPACID, MACID, OCBYP, VOC
320 FORMAT(
& /'      2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT ($/year)',
& /6X,'      Coal: ',F7.2,' $/MMBtu      ',F10.1,' lb/hr      ',F14.0,
& /6X,'      Ash Disposal: ',F7.2,' $/ton      ',F10.1,' ton/day      ',F14.0,
& /6X,'      Sulfuric Acid: ',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,330) WGTE, WCHE, WSE, WSTE, WLE, WSAE, WAUXE, WOFE,
& WBFEE,WGE, WGFE,WNETE, XO, DPERKW, FOCN, VOCINC,
& BYPN, FUELN, VOCN, CELEC
330 FORMAT (/
& 'G. COST OF ELECTRICITY -----',
& '-----',
& /'      Power Summary (MWe)      ',
& 'Auxiliary Loads (MWe)',
& /'-----',
& '-----',
& /'Gas Turbine Output      ',F7.2,
& '      Coal Handling ',F6.2,' Sulfation      ',F6.2,
& /'Steam Turbine Output ',F7.2,
& '      Limestone Hdlg ',F6.2,' Acid Rem.      ',F6.2,
& /'Total Auxiliary Loads ',F7.2,
& '      Oxidant Feed ',F6.2,' Steam Cycle ',F6.2,
& /'-----',
& '      Gasification ',F6.2,' General Fac ',F6.2,
& /'Net Electricity      ',F7.2,
& '      Zinc Ferrite ',F6.2,
& //27X,'      Capital Cost: ',F10.2,' $/kW',
& /27X,'      Fixed Operating Cost: ',F10.2,' $(kW-yr)',
& /8X,'Incremental Variable Costs: ',F6.2,' mills/kWh',
& /8X,'Byproduct Credit:      ',F6.2,' mills/kWh',
& /8X,'Fuel Cost:      ',F6.2,' mills/kWh',
& /27X,'Variable Operating Cost: ',F10.2,' mills/kWh',
& /'COST OF ELECTRICITY ----->',
& F7.2,' mills/kWh'/)

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WRITE(NRPT,340) HEATRATE, EFFICNCY
340 FORMAT(
&  /'The plant heatrate (HHV) is: ',F7.0,
&  ' BTU/kWh. Efficiency: ',F7.4/)
WRITE(NRPT,350) MCOALIN, MH2OIN, MASHOUT, MACIDO,
&  ECO2, ECO, ESO2, ECOS, ECH4, EH2S, ENOX
350 FORMAT(/
&  'H. ENVIRONMENTAL SUMMARY -----',
&  '-----',
&  /15X,'INPUTS:   Coal           ',F10.4,' lb/kWh ',
&  /15X,'           Water          ',F10.4,' lb/kWh ',
&  /15X,'OUTPUTS:  Ash            ',F10.4,' lb/kWh ',
&  /15X,'           Sulfuric Acid ',F10.4,' lb/kWh ',
&  /15X,'           CO2 emissions ',F10.4,' lb/kWh ',
&  /15X,'           CO emissions  ',F10.4,' lb/kWh ',
&  /15X,'           SO2 emissions ',F10.4,' lb/MMBtu ',
&  /15X,'           COS emissions ',F10.4,' lb/MMBtu ',
&  /15X,'           CH4 emissions ',F10.4,' lb/kWh ',
&  /15X,'           H2S emissions ',F10.4,' lb/MMBtu ',
&  /15X,'           NOx emissions ',F10.4,' lb/MMBtu ',
&  /'NOTE: NOx emissions are based on fuel bound ammonia '/')
RETURN
END

```

C.3 Sample Output

The following is a sample output from an ASPEN simulation of the air-blown Lurgi-based IGCC system with hot gas cleanup 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.

*** SUMMARY OF COST MODEL INPUT PERFORMANCE PARAMETERS ***

PLANT SECTION	INPUT PERFORMANCE PARAMETERS	VALUE	UNITS
Coal Feed	Mass flow of coal to gasifier	478701.30	lb/hr
Limestone Feed	Mass flow of limestone (mixed)	0.00	lb/hr
	Mass flow of limestone (cisolid)	65.36	lb/hr
Oxidant Feed	Work to boost air compressor	0.130020E+08	Watts
	First precooler inlet air temp.	707.93	F
	First precooler outlet air temp.	651.91	F
	First precooler inlet BFW temp.	505.58	F
	First precooler outlet BFW temp.	605.29	F
	Heat transfer in first precooler	0.166839E+08	BTU/hr
	Heat trans. to BFW from HRSG	0.514713E+09	BTU/hr
	Heat trans. to BFW from Regen.	0.727726E+08	BTU/hr
	Heat leaving from economizer	0.462867E+09	BTU/hr
Gasification	Gasifier output syngas pres.	300.00	psia
	Gasifier output syngas temp.	1100.00	F
	Gasifier output syngas density	0.178608E-01	lbmole/ft3
Zinc Ferrite	Zinc Ferrite inlet syngas flow	95213.42	lbmole/hr
	Off-gas to sulfuric acid plant	3623.83	lbmole/hr
	SO2 to sulfuric acid plant	491.49	lbmole/hr
	Off-gas temp. to sul. acid plant	200.00	lbmole/hr
	Zinc ferrite inlet H2S flow	492.62	lbmole/hr
	Zinc ferrite inlet COS flow	0.00	lbmole/hr
Sulfation	Gasifier ash removal	42076.83	lb/hr
	Gasifier ash removal	8951.34	lb/hr
	Gasifier fines removal	38.30	lb/hr
	Gasifier fines removal	344.66	lb/hr
BFW Treating	Raw water to power plant	613556.81	lb/hr
	Steam turbine condensate	1093213.22	lb/hr
Gas Turbine	Gas turbine inlet air flow	9000000.04	lb/hr
Steam Cycle	HRSG outlet HP steam pres.	1465.00	psia
	HRSG outlet HP steam flow	1059611.67	lb/hr
	Gas turbine net shaft work	-.491118E+09	Watts
	Steam turbine net shaft work	-.205499E+09	Watts
	Steam cycle auxiliary power	0.291185E+07	Watts
	Acid gas auxiliary power	0.855729E+07	Watts

Miscellaneous	Moisture in coal feed	12.00 wt %
	Ash in coal	10.00 wt %
	Temp. of ambient air	59.00 F
	Heating value of coal	12774.00 BTU/lb
Environmental	CO2 from gas turbine	23762.10 lbmole/hr
	CO from gas turbine	0.63 lbmole/hr
	SO2 from gas turbine	0.95 lbmole/hr
	COS from gas turbine	0.00 lbmole/hr
	CH4 from gas turbine	0.00 lbmole/hr
	H2S from gas turbine	0.00 lbmole/hr
	NH3 to gas turbine	69.57 lbmole/hr
	Actual coal heating value	11248.73 BTU/lb

COST VAR WARNING ----	Variable MCFGI value of	478701.305
	in DCSS above the upper limit of	433000.000

COST VAR WARNING ----	Variable VSNZFI value of	6834.423
	in DCCY above the upper limit of	6000.000

COST VAR WARNING ----	Variable FSO2 value of	0.136
	in DCSA above the upper limit of	0.130

COST VARIABLE RESET -	Variable TOGSAI value of	200.000
	in DCSA reset to the lower limit of	500.000

COST SUMMARY

Air Blown Lurgi-Based I&CC System with Hot Gas Cleanup

A. COST MODEL PARAMETERS -----			
Plant Capacity Factor:	0.65	Cost Year:	January 1989
General Facilities Factor:	0.15	Plant Cost Index:	351.5
Indirect Construction:	0.25	Chemicals Cost Index:	411.3
Sales Tax:	0.06	Escalation:	0.00
Engr & Home Office Fee:	0.15	Interest:	0.06
Project Contingency:	0.20	Years of construction:	4
Number of Shifts:	4.75	Byproduct marketing:	0.10
Fixed Charge Factor:	0.1034	Average Labor Rate:	19.70
Variable Levelization Cost Factor:	1.0000	Book Life (years):	30

B. PROCESS CONTINGENCY AND MAINTENANCE COST FACTORS -----		
Plant Section	Process Contingency	Maintenance Cost Factor
-----	-----	-----
Coal Handling	0.050	0.020
Limestone Handling	0.000	0.030
Oxidant Feed	0.050	0.020
Gasification	0.050	0.045
Coke, Ash, & Bent. Subsystems	0.050	0.030
High Temp. Cyclones	0.100	0.015
Zinc Ferrite	0.400	0.045
Sulfuric Acid Plant	0.060	0.020
Sulfation	0.500	0.040
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.030	0.015
Heat Recovery Steam Generator	0.030	0.015
Steam Turbine	0.030	0.015
General Facilities	0.100	0.015

C. DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS (\$1,000) -----				
Plant Section	Number of Units		Direct	Process
-----	Operating	Total	Capital Cost	Contingency
-----	-----	-----	-----	-----
Coal Handling	1	1	20966.	1569.
Limestone Handling	0	0	0.	0.
Oxidant Feed	3	3	4659.	349.
Gasification	13	15	98347.	7360.
Coke, Ash, & Bent. Subsystems	13	15	13161.	985.
High Temp. Cyclones	26	30	6469.	968.
Zinc Ferrite	52	120	79190.	47408.
Sulfuric Acid Plant	1	1	7371.	662.
Sulfation	0	0	0.	0.
Boiler Feedwater Treatment	1	1	4459.	0.
Gas Turbine	3	3	96000.	4310.
Heat Recovery Steam Generator	3	3	27036.	1214.
Steam Turbine	1	1	32124.	1442.
General Facilities	N/A	N/A	58467.	8751.

D. TOTAL CAPITAL REQUIREMENT (\$1,000) -----	
Description	Annual Cost
-----	-----
Total Direct Cost	448250.
Indirect Construction Cost	112062.
Sales Tax	22188.
Engineering and Home Office Fees	87375.
Environmental Permitting	1000.
Total Indirect Costs	222626.
Total Process Contingencies	75017.
Project Contingency	149179.
Total Plant Cost	895071.
AFDC	85644.
Total Plant Investment	980715.
Preproduction (Startup) Costs	25960.
Inventory Capital	15823.
Initial Catalysts and Chemicals	199074.
Land	1979.
TOTAL CAPITAL REQUIREMENT (\$1,000) ----->	1233358.

E. FIXED OPERATING COSTS (\$/year) -----	
Description	Annual Cost
-----	-----
Operating Labor	10315708.
Maintenance Costs	25503661.
Administration and Supervision	6155152.
TOTAL FIXED OPERATING COST (\$/year) ----->	41974521.

F. VARIABLE OPERATING COSTS -----

1. CONSUMABLES (\$/year)

Description	Unit Cost	Material Requirement	Annual Operating Cost
-----	-----	-----	-----
Sulfuric Acid:	110.00 \$/ton	1498.3 ton/yr	164818.
NaOH:	220.00 \$/ton	311.9 ton/yr	68614.
Na2 HPO4:	0.70 \$/lb	1514.5 lb/yr	1060.
Hydrazine:	3.20 \$/lb	7283.2 lb/yr	23306.
Morpholine:	1.30 \$/lb	6773.6 lb/yr	8806.

Lime:	80.00 \$/ton	528.6 ton/yr	42285.
Soda Ash:	160.00 \$/ton	587.4 ton/yr	93988.
Corrosion Inh.:	1.90 \$/lb	105123.9 lb/yr	199735.
Surfactant:	1.25 \$/lb	105123.9 lb/yr	131405.
Chlorine:	250.00 \$/ton	16.9 ton/yr	4233.
Biocide:	3.60 \$/lb	18835.5 lb/yr	67808.
Zinc Fer Sorb:	3.30 \$/lb	1084598.1 lb/yr	3579174.
Limestone:	18.00 \$/ton	0.0 ton/yr	0.
Sul Acid. Cat.:	1.90 \$/l	3986.2 l/yr	7574.
Coke:	150.00 \$/ton	1443.0 ton/yr	216447.
Bentonite:	0.03 \$/lb	4594482.7 lb/yr	133240.
Fuel Oil:	42.00 \$/bbl	37160.6 bbl/yr	1560744.
Plant Air Ads.:	2.80 \$/lb	2787.0 lb/yr	7804.
Raw Water:	0.73 \$/Kgal	418783.2 Kgal/yr	305712.
Waste Water:	840.00 \$/gpm ww	0.0 lb/hr	0.
LPG - Flare:	11.70 \$/bbl	3251.5 bbl/yr	38043.
TOTAL CONSUMABLES (\$/year) ----->			6654795.

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

Coal:	1.61 \$/MMBtu	478701.3 lb/hr	56057647.
Ash Disposal:	10.00 \$/ton	616.9 ton/day	1463675.
Sulfuric Acid:	40.00 \$/ton	147448.9 ton/yr (	5308161.)

TOTAL VARIABLE OPERATING COST (\$/year) -----> 58867956.

G. COST OF ELECTRICITY -----

Power Summary (MWe)		Auxiliary Loads (MWe)	
Gas Turbine Output	483.75	Coal Handling	0.90
Steam Turbine Output	202.42	Limestone Hdlg	0.00
Total Auxiliary Loads	31.09	Oxidant Feed	13.00
		Gasification	2.39
		Zinc Ferrite	0.00
Net Electricity	655.07		

Capital Cost:	1882.78 \$/kW
Fixed Operating Cost:	64.08 \$/(kW-yr)
Incremental Variable Costs:	2.18 mills/kWh
Byproduct Credit:	1.42 mills/kWh
Fuel Cost:	15.03 mills/kWh
Variable Operating Cost:	15.78 mills/kWh

COST OF ELECTRICITY -----> 61.23 mills/kWh

The plant heatrate (HHV) is: 8220. BTU/kWh. Efficiency: 0.4154

H. ENVIRONMENTAL SUMMARY -----

INPUTS:	Coal	0.7308 lb/kWh
	Water	0.9366 lb/kWh
OUTPUTS:	Ash	0.0785 lb/kWh
	Sulfuric Acid	0.0791 lb/kWh
	CO2 emissions	1.5961 lb/kWh
	CO emissions	0.0000 lb/kWh
	SO2 emissions	0.0405 lb/MMBtu
	COS emissions	0.0000 lb/MMBtu
	CH4 emissions	0.0000 lb/kWh
	H2S emissions	0.0000 lb/MMBtu
	NOx emissions	0.2786 lb/MMBtu

NOTE: NOx emissions are based on fuel bound ammonia

APPENDIX D. NUMBER OF OPERATING TRAINS SUBROUTINE

This subroutine is used to estimate the number of operating trains required for many of the direct cost models for the oxygen-blown KRW-based IGCC system with cold gas cleanup.

```
SUBROUTINE NTRAINS (VLL,VAR,VUL,NO)
IMPLICIT DOUBLE PRECISION (A-H,O-Z)

C
C Author: H.C. Frey                      Date: March 24, 1990
C Short Title: Calculation of Number of Operating Trains
C Description: This subroutine determines the number of operating
C trains for a process area based on the key mass flow rate for the
C the process area and the upper and lower limits for the mass flow
C rate per train. The subroutine searches for the smallest number
C of operating trains that would be required, because there is
C generally an economy of scale associated with larger train sizes.
C
C NO=0
C
C Search for the minimum number of trains required to bring
C the mass flow rate per train within the upper limit.
C
C DO 100 I=1,100
C NO=NO+1
C TEST=VAR/REAL(NO)
C
C If the mass flow per train is within the limits for the train,
C then exit the loop and return the number of trains
C
C IF((TEST.LE.VUL).AND.(TEST.GE.VLL)) GOTO 200
C
C If the mass flow per train is less than the lower limit,
C but the number of trains tested is only one, then it is not
C possible to increase the mass flow per train by changing
C the number of trains. Therefore, exit the loop, and report
C the number of operating trains as one.
C
C IF((TEST.LT.VLL).AND.(NO.EQ.1)) GOTO 200
C
C If the mass flow per train is less than the lower limit and
C the number of trains is great than one, then
C do not continue to increase the number of trains. Instead,
C compare the error of the current mass flow per train versus
C the lower limit with the error of the previous step (NO=NO-1).
C First, determine if the mass flow rate per train for NO=NO-1
C is greater than the upper limit or less than the lower limit.
C Then estimate the appropriate error. Choose the number of
C trains that minimizes the error.
C
C IF((TEST.LT.VLL).AND.(NO.GT.1)) THEN
C   ERR1 = ABS(TEST-VLL)
C   TEST2=VAR/REAL(NO-1)
C   IF (TEST2.GT.VUL) THEN
C     ERR2=ABS(TEST2-VUL)
C   ELSEIF (TEST2.LT.VLL) THEN
C     ERR2=ABS(TEST2-VLL)
```

```
        ELSE
            ERR2=0
        ENDIF
        IF (ERR2.LT.ERR1) THEN
            NO=NO-1
            GOTO 200
        ELSE
            GOTO 200
        ENDIF
    ENDIF
100 CONTINUE
200 CONTINUE
    RETURN
END
```

APPENDIX E. RANGE-CHECKING SUBROUTINES

These subroutines compare the value of key performance variables used in the direct cost regression models with the upper and lower limits of the ranges used to develop the models. Subroutine USRWCK will print a warning if the variable exceeds the limits, indicating that the regression model is being extrapolated and that the results should be interpreted with care. Subroutine USRRCK will print a warning that a variable exceeds the limits of the range used to develop the regression model, and will set the value of the variable to either the upper or lower limit of the range. This subroutine is used when extrapolation is not acceptable. The user should interpret the model results carefully when the value of a performance variable has been changed; in such cases, the performance assumed in the ASPEN simulation is different from the performance used to estimate costs.

```

      SUBROUTINE USRWCK (LL, VAR, UL, VNAME, CNAME)
      IMPLICIT DOUBLE PRECISION (A-H,O-Z)
      REAL*8 VAR, UL, LL
      CHARACTER*6 VNAME, CNAME
      COMMON /USRRPT/ NREPT
C
C   This subroutine compares the value of a selected variable with
C   acceptable upper and lower limits and prints a warning if the
C   variable exceeds the limits.
C
C   Variable Definitions
C       VAR   = Variable to be checked
C       UL    = Upper limit of variable
C       LL    = Lower limit of variable
C       VNAME = Name of the variable
C       CNAME = Name of the variable being calculated as a function of VNAME
C
      IF (VAR.LT.LL) WRITE(NREPT,100) VNAME, VAR, CNAME, LL
      IF (VAR.GT.UL) WRITE(NREPT,110) VNAME, VAR, CNAME, UL
100  FORMAT('COST VAR WARNING ---- Variable ',A6,' value of ',F14.3,
      &      /'                in ',A6,' below the lower limit of ',F14.3,/)
110  FORMAT('COST VAR WARNING ---- Variable ',A6,' value of ',F14.3,
      &      /'                in ',A6,' above the upper limit of ',F14.3,/)
      RETURN
      END
C
C   -----
C
      SUBROUTINE USRRCK (LL, VAR, UL, VNAME, CNAME)
      IMPLICIT DOUBLE PRECISION (A-H,O-Z)
      REAL*8 VAR, UL, LL
      CHARACTER*6 VNAME, CNAME
      COMMON /USRRPT/ NREPT
C
C   This subroutine compares the value of a selected variable with
C   acceptable upper and lower limits and prints a warning if the
C   variable exceeds the limits. It also resets the value of the
C   variable to the appropriate limit.
```

```

C
C   When using this subroutine, care must be taken to reset the variable
C   only when necessary, and to not unduly affect downstream calculations.
C
C   Variable Definitions
C       VAR  = Variable to be checked
C       UL   = Upper limit of variable
C       LL   = Lower limit of variable
C       VNAME = Name of the variable
C       CNAME = Name of the variable being calculated
C               as a function of VNAME
C
C       IF (VAR.LT.LL) THEN
C           WRITE(NREPT,100) VNAME, VAR, CNAME, LL
C           VAR=LL
C       ELSEIF (VAR.GT.UL) THEN
C           WRITE(NREPT,110) VNAME, VAR, CNAME, UL
C           VAR=UL
C       ENDIF
100 FORMAT('COST VARIABLE RESET - Variable ',A6,' value of ',F14.3,
&         /'           in ',A6,' reset to the lower limit of ',F14.3,/)
110 FORMAT('COST VARIABLE RESET - Variable ',A6,' value of ',F14.3,
&         /'           in ',A6,' reset to the upper limit of ',F14.3,/)
C       RETURN
C       END

```

APPENDIX F. FINANCIAL SUBROUTINES

Two subroutines were developed to estimate the cost of electricity. The first subroutine, USRFCF, calculates the fixed charge factor. The second subroutine, USRVCLF, calculates the variable cost levelization factor. Both are based on BASIC programs listing in the EPRI Technical Assessment Guide (1986).

```
SUBROUTINE USRFCF (IBL, FCF, EFI)
IMPLICIT DOUBLE PRECISION (A-H,O-Z)
DIMENSION DEPR(60)
```

```
C
C      Financial analysis program to calculate carrying charges
C      From EPRI TAG, Volume 3, May 1987.  EPRI P-4463-SR
C      ECONCC.BAS  Version 4, 12/15/86
C
C      Adapted to Fortran by H.C. Frey, 2/1/90
C
C      Variables:
C      CD      = Cost of debt (decimal)
C      CP      = Cost of preferred shares (decimal)
C      CE      = Cost of common shares (decimal)
C      DR      = Debt ratio (decimal)
C      PR      = Preferred ratio (decimal)
C      TX      = Income Tax Rate (decimal)
C      XITC     = Investment tax credit (decimal)
C      ITDM     = Switch for depreciation method
C                  1 = Accelerated depreciation based on ACRS class
C                  (THIS OPTION IS NOT OPERATIONAL HERE)
C                  2 = Straight-line depreciation over book life
C                  3 = Straight-line depreciation based on ACRS class
C      ICL      = ACRS class (ONLY OPTION 4 IS AVAILABLE HERE)
C                  1 = Autos, light trucks, and computers - 5 yrs
C                  2 = Mining equipment - 7 yrs
C                  3 = Nuclear plant and combustion turbine - 15 yrs
C                  4 = Fossil plant and T&D equipment - 20 yrs
C      IBL      = Useful booklife (1 to 60 years if SL depr, 20 to 60
C                  if SL based on ACRS class is used)
C      SLTXDP   = Straight-Line Tax Depreciation
C      DEPR(i)  = Depreciation for each year i
C      EFI      = Discount rate (effective interest rate)
C      ER       = Equity ratio
C      EC       = Equity cost
C      PTI      = Property taxes and insurance
C      BD       = Book depreciation rate net of ITC
C      SB       = Remaining book value
C      V        = Present value factor
C      A        = Annuity factor
C      TD       = Tax depreciation rate
C      RE       = Return on Equity
C      RD       = Return on Debt
C      DT       = Deferred Income Taxes
C      TP       = Taxes paid
C      CC       = Carrying charges (year by year)
C      SP       = Cumulative present value of carrying charges
C      SB       = Depreciation book value, net def. tax
```

```

C      FCF      = Fixed charge factor
C
C
DATA DEPR/0.075, 0.069, 0.064, 0.059, 0.055, 0.051, 0.047,
&      8*0.045, 5*0.044, 40*0./
CD=0.046
CP=0.052
CE=0.087
DR=0.50
PR=0.15
TX=0.38
XITC=0.00
ITDM=3
ICL=4
SLTXDP=1.0/REAL (IBL)
IF (ITDM.EQ.3) GOTO 930
DO 900 I=1, IBL
      DEPR(I)=SLTXDP
900 CONTINUE
930 CONTINUE
      EFI=(CD*DR+CP*PR+CE*(1-DR-PR))
      ER=1-DR
      IF (ER.NE.0) THEN
          EC=(CP*PR+CE*(1-DR-PR))/ER
      ELSE
          EC=0
      ENDIF
      PTI=0.02
      BD=(1-XITC)/IBL
      SB=1-XITC
      SP=0.0
      PG=0.0
      DO 910 I=1, IBL
          V=(1+EFI)**(-I)
          IF (EFI.NE.0) THEN
              A=(1-V)/(EFI)
          ELSE
              A=1
          ENDIF
          TD=DEPR(I)
          RE=SB*ER*EC
          RD=SB*DR*CD
          DT=(TD-SLTXDP)*TX
          TP = TX / ( 1 - TX ) * ( BD - TD + DT + RE )
          CC=BD+DT+RD+RE+TP+PTI
          SP=SP+CC*V
          SB=SB-BD-DT
910 CONTINUE
      FCF=SP/A
      RETURN
      END
C
C -----
C
SUBROUTINE USRVCLF (EI, ER, DISC, ITIME, VCLF)
IMPLICIT DOUBLE PRECISION (A-H,O-Z)
C
C      Financial analysis program to calculate levelizing factors

```

```

C      From:  EPRI TAG, Volume 3, May 1987.  EPRI P-4463-SR
C      ECONLN.BAS  Version 3.0  12/04/86
C
C      Levelized value of a series of uniform payments with constant
C      apparent escalation.  Converts to beginning of first year dollars.
C
C      Adapted to Fortran by H.C. Frey  1/31/90
C
C      SUBROUTINE ARGUMENTS
C          EI      = annual inflation rate (decimal)
C          ER      = annual real escalation rate (decimal)
C          DISC    = annual discount rate
C          ITIME   = time period, years
C          VCLF    = variable cost levelization factor
C
C      EA=(1+ER)*(1+EI)-1
C      XK=(1+EA)/(1+DISC)
C      AN=((1+DISC)**ITIME-1)/(DISC*(1+DISC)**ITIME)
C      VCLF=XK*(1-XK**ITIME)/(AN*(1-XK))
C      RETURN
C      END

```


DOE/MC/24248-2901

Stochastic Modeling of Coal Gasification Combined Cycle
Systems: Cost Models for Selected Integrated Gasification
Combined Cycle (IGCC) Systems

DOE