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

**Topical Report
July 1995**

September 1995

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

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

By
Carnegie Mellon University
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Performance and Cost Models For The Direct Sulfur Recovery Process - Task 1, Volume 3

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For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
P.O. Box 880
Morgantown, West Virginia 26507-0880

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September 1995

**Performance and Cost Models for
the Direct Sulfur Recovery Process**

Task 1 Topical Report, Volume 3

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

For

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September 1995

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PREFACE

Insert one page introduction to the METC project, to be provided by Dr. E.S. Rubin.

Nomenclature

English Letter Symbols

$[A]_i$	= concentration of species A entering Stage I reactor (volume %)
$[A]_o$	= concentration of species A leaving Stage I reactor (volume %)
$[A]_G$	= concentration of species A generated in Stage I reactor by a certain reaction (volume %)
$[A]_{II,i}$	= concentration of species A entering Stage II reactor (volume %)
$[A]_{II,o}$	= concentration of species A leaving Stage II reactor (volume %)
$[A]_{II,G}$	= concentration of species A generated in Stage II reactor by a certain reaction (volume %)
AFDC	= Allowance for funds during construction, \$1,000
C_{ProcC}	= Process contingencies costs, \$1,000
C_{ProjC}	= Project contingencies costs, \$1,000
C_{EHO}	= Engineering and home office fees costs, \$1,000
C_{GF}	= General facilities costs, \$1,000
C_{IC}	= Initial catalyst and chemicals costs, \$1,000
C_{InvC}	= Inventory capital costs, \$1,000
C_{PP}	= Preproduction (startup) costs, \$1,000
C_{Roy}	= Royalties costs, \$1,000
DC_{DSRP}	= Direct capital cost of the DSRP, \$1,000
$DC_{GH,1}$	= Direct capital cost of the gas handling system for Stage I, \$1,000
$DC_{GH,2}$	= Direct capital cost of the gas handling system for Stage II, \$1,000
DC_i	= Direct capital cost of process area i, \$1,000
$DC_{R,1}$	= Direct capital cost of the reactor for Stage I, \$1,000
$DC_{R,2}$	= Direct capital cost of the reactor for Stage II, \$1,000
$DC_{S,1}$	= Direct capital cost of the solids handling system for Stage I, \$1,000
$DC_{S,2}$	= Direct capital cost of the solids handling system for Stage II, \$1,000
f_{InvC}	= Inventory capital cost factor
f_p	= Pressure correction factor (ratio)
f_{pp}	= Pre-production cost factor
$f_{ProjC,i}$	= Project contingency factors for each process area (fraction of direct cost)
f_{Roy}	= Royalties cost factor
FOC	= Fixed operating cost, \$1,000/yr
L_{cat}	= Catalyst life, years

m_{cat}	= Mass amount of initial catalyst, lb
$m_{s,R1,o}$	= Mass flow rate of sulfur from Stage I, lb/hr
$m_{s,R2,o}$	= Mass flow rate of sulfur from Stage II, lb/hr
$N_{m,FOC}$	= Number of months of FOC to be included in preproduction costs (months)
$N_{m,VOC}$	= Number of months of VOC to be included in preproduction costs (months)
$N_{O,DSRP}$	= Number of operating (on-line) equipment trains for the DSRP
$N_{T,DSRP}$	= Total number of equipment trains for the DSRP
OC_{cat}	= Operating cost for catalyst, \$/yr
$P_{g,R1,i}$	= Pressure of the gas entering Reactor No. 1, psia
PCI	= <i>Chemical Engineering Plant Cost Index</i>
R	= mole ratio of CO ₂ to COS, varying with space velocity
$SV_{IA,H}$	= space velocity for Stage I using Catalyst A at high space velocity (scc/cc-hr)
$SV_{IA,L}$	= space velocity for Stage I using Catalyst A at low space velocity (scc/cc-hr)
SV_{IIB}	= space velocity for Stage II using Catalyst B (scc/cc-hr)
SV_{IIC}	= space velocity for Stage II using Catalyst C (scc/cc-hr)
$SV_{R1,actual}$	= Actual space velocity for Reactor No. 1, scc/cc-hr.
$SV_{R1,ref}$	= Reference space velocity for Reactor No. 1, scc/cc-hr
$SV_{R2,actual}$	= Actual space velocity for Reactor No. 2, scc/cc-hr.
$SV_{R2,ref}$	= Reference space velocity for Reactor No. 2, scc/cc-hr
TCR	= Total capital requirement, \$1,000
TDC	= Total direct cost, \$1,000
TPC	= Total plant cost, \$1,000
TPI	= Total plant investment, \$1,000
UC_{cat}	= Unit cost of catalyst, \$/lb
$V_{g,R1,i}$	= Volumetric flowrate of gas entering Reactor No. 1, ft ³ /hr.
$V_{g,R2,i}$	= Volumetric flowrate of gas entering Reactor No. 2, ft ³ /hr.
VOC	= Variable operating costs, \$1,000/yr

Greek Letter Symbols

$\eta_{SO_2,I}$	= fraction of SO ₂ entering Stage I which is converted to elemental sulfur in that reactor
$\eta_{SO_2,II}$	= fraction of SO ₂ entering Stage II which is converted to elemental sulfur in that reactor
$\eta_{\#}$	= portion of $\eta_{SO_2,I}$ accounted for by reaction (#)

1.0 INTRODUCTION

The purpose of this project is to develop performance and cost models of the Direct Sulfur Recovery Process (DSRP). The DSRP is an emerging technology for sulfur recovery from advanced power generation technologies such as Integrated Gasification Combined Cycle (IGCC) systems. In IGCC systems, sulfur present in the coal is captured by gas cleanup technologies to avoid creating emissions of sulfur dioxide to the atmosphere. The sulfur that is separated from the coal gas stream must be collected. Leading options for dealing with the sulfur include byproduct recovery as either sulfur or sulfuric acid. Sulfur is a preferred byproduct, because it is easier to handle and therefore does not depend as strongly upon the location of potential customers as is the case for sulfuric acid.

Many concepts for coal gasification power generation systems have been modeled using the ASPEN chemical process simulator. The U.S. Department of Energy (DOE) is among many organizations that have used process simulation tools to evaluate and optimize power generation concepts. The work described here is in support of DOE efforts to: (1) develop improved performance models for key components of the power generation systems; (2) develop new engineering-economic models; and (3) apply the models to gain insights into research priorities for future technology development and the comparison of technology alternatives.

In this report, the need for new sulfur recovery technologies for IGCC systems is reviewed. The DSRP is identified as one of the leading alternatives sulfur recovery. It offers the potential of low cost and high efficiency sulfur recovery. Technical background information is used to provide a basis for the development of a performance model of the DSRP. A new engineering-economic model of the DSRP is developed and documented. The performance and cost models are illustrated through several case studies. Needs for future work regarding model refinement and model applications are discussed. Because the DSRP has not yet been proven at full-scale, one key area for future work is the quantification of uncertainties inherent in making estimates of DSRP performance and cost for advanced IGCC systems. Thus, future case studies will feature a probabilistic approach to technology assessment.

2.0 ENVIRONMENTAL CONTROL IN IGCC SYSTEMS

IGCC systems offer the potential of lower environmental discharges than conventional coal-fired power plants. First, general features of IGCC technology are reviewed, followed by more detailed discussion of cold gas cleanup, hot gas cleanup, and sulfur recovery options.

2.1 IGCC Systems

The major components of coal are carbon and hydrogen. In conventional combustion, coal reacts with oxygen to produce carbon dioxide and water. The heat released from coal combustion is used to power a steam cycle with a steam turbine generator. During the coal gasification process, coal reacts with steam and a substoichiometric amount of oxygen, typically one-fifth to one-third the amount needed for complete combustion, at high temperature (e.g., 2000°C) and pressure (e.g., 400 psi) to produce a gas containing carbon monoxide and hydrogen. The partial combustion of the coal provides thermal energy required for the endothermic gasification reactions. The coal gas is highly combustible and can serve as fuel to gas turbines. However, the raw fuel gas contains contaminants such as particulate matter and hydrogen sulfide that can damage the turbines and result in unacceptable environmental discharges. Therefore, fuel gas cleanup steps are required to remove these contaminants. The cleaned coal gas is employed as a fuel for a gas turbine combined cycle (GTCC) system. A GTCC is composed of a simple cycle gas turbine and a bottoming steam cycle used to extract additional energy from the hot exhaust gases of the gas turbine. When a GTCC is integrated with a coal gasifier, the resulting process is known as an Integrated Gasification Combined Cycle (IGCC) (Simbeck, 1983). This system is shown schematically in Figure 1. This report focuses on the development of an analytic performance model for a component of an advanced IGCC system.

Coal contains components such as carbon, hydrogen, oxygen, nitrogen, sulfur, and ash. Sulfur and ash in particular are of key environmental and operational concern in coal gasification systems. When coal is gasified, sulfur-containing species such as hydrogen sulfide (H_2S) and carbonyl sulfide (COS) are formed. Sulfur species combusted in the gas turbine are converted to sulfur dioxide (SO_2) and are released into the atmosphere. Particulate matter or "fly ash" from the gasifier is also a source of emissions. Both sulfur and particulate matter emissions must be controlled to comply with federal, state, and local emission regulations and environmental permit agreements. For example, the Environmental Protection Agency (EPA) has regulations regarding industrial utility

Figure 1. Schematic Diagram of an Integrated Gasification Combined Cycle (IGCC) System

emissions, embodied in New Source Performance Standards (NSPS) for coal-fired boilers and gas turbines. While it is not clear which of these regulations would apply to IGCC systems, it is likely that state or local restrictions would be more stringent than others. In addition to environmental concerns, there are operational reasons to reduce emissions. Sulfur and ash contribute to corrosion, erosion, and fouling of turbine blades. For these reasons, it is essential to remove a high percentage (typically 95-99 percent) of the sulfur and more than 99 percent of the fly ash from the syngas before it is combusted to produce electricity.

2.2 Cold Gas Cleanup

Sulfur-containing species are typically removed from the coal gas by either hot gas desulfurization or cold gas desulfurization. The methods of cold gas desulfurization in use commercially can be divided into the two main categories of chemical and physical absorption. Chemical absorption involves reaction of acid gases (H_2S and CO_2) in the syngas with various solutions of alkanolamines or alkaline salts in water. In the absorber, the lean solution is contacted with the acid gas at high pressure. The temperature is typically between 80°F (27°C) and 120°F (49°C). The sulfur-rich solution is then reduced to low pressure and heated in the stripper to reverse the reaction and regenerate the absorbent solution. The temperature in the regenerator is the boiling temperature of the solution, usually 220-260°F (105-127°C). This solution is then cooled, pressurized, and recycled back to the absorber. This process removes acid gas from the syngas at low to moderate pressures, but the selectivity of H_2S absorption over CO_2 absorption is low (SFA, 1987).

The physical absorption processes rely on the solubility of the acid gases in certain organic solvents and are generally favored at high pressures (Simbeck, 1983). The hot coal gas from the gasifier is first cooled to approximately 100°F (38°C) using a series of shell and tube heat exchangers. The cooled gas is then cleaned through counter-current contact with the organic solvent in a packed bed. The removed sulfur is stripped from the solvent by lowering the pressure and raising the temperature. The cleaned gas is then sent on to the gas turbines and the sulfur-laden gas is sent to byproduct recovery (Fluor, 1985). Two popular methods of physical cold gas desulfurization are Selexol, which uses dimethyl ether of polyethylene glycol as a solvent, and Rectisol, which uses methanol as a solvent. These processes can effectively remove up to 95 percent or more of the H_2S in the coal gas (SFA, 1987). The offgas from cold gas desulfurization processes usually has a high sulfur concentration (in the form of H_2S).

2.3 Hot Gas Desulfurization

Since gas usually leaves the coal gasifier at an elevated temperature, hot gas desulfurization would be more efficient and economical than cold gas desulfurization. It would eliminate the need for heat exchangers to cool the hot gases, thereby potentially reducing the capital cost of the plant. Furthermore, hot gas cleanup enables more of the sensible heat of the fuel gas to be delivered to the gas turbine, where it can be efficiently converted to electricity. Thus, hot gas cleanup also offers the potential benefit of higher plant efficiency.

Mixed-metal oxide sorbents, such as zinc ferrite (ZnFe_2O_4) and zinc titanate (ZnTiO_3), are proposed to remove H_2S from the raw coal gas (e.g., Gangwal, 1991). As an example of hot gas desulfurization systems, we consider a fixed bed design as illustrated by Figure 2. Two parallel reactor vessels each contain a fixed bed of mixed-metal oxide sorbent pellets. As the diagram shows, each vessel cycles between absorption and regeneration phases. In Figure 2, the reactor on the left is in the absorption mode while the reactor on the right is in the regeneration mode. The reactors usually operate at pressures between 300 psia (20 atm) and 400 psia (27 atm) and inlet gas temperature of about 1100°F (600°C). The relevant reaction mechanisms occurring during absorption, using zinc ferrite sorbent, are:

This yields an overall absorption phase reaction of:

As this reaction illustrates the sorbent, sulfur species, and reducing gases (CO and H_2) react to form carbon dioxide, water, and two sulfated sorbent species (ZnS and FeS).

Once the absorption of hydrogen sulfide is complete, the cleaned syngas is sent onward to be used as fuel for gas turbines. The sorbent must then be regenerated so it can be reused. The sulfated sorbent is reacted with a stoichiometric amount of oxygen. This produces regenerated sorbent and an offgas stream typically containing one to three percent SO_2 for fixed-bed systems, although concentrations as high as 12 percent have been

Figure 2. Absorption/Regeneration System for Hot Gas Desulfurization.

observed for moving bed systems. Because the regeneration reactions are highly exothermic, a thermal diluent such as steam must be added to control temperature. Temperatures above 650 °C may cause sintering of the sorbent, so steam is necessary to keep the temperature below this level (Frey, 1991). Note that the sulfur concentration of the offgas is much lower than that occurring in conventional cold gas desulfurization systems. The relevant reaction mechanisms occurring during regeneration of zinc ferrite sorbent are (METC, 1987):

This yields an overall regeneration phase reaction of:

The reactor vessels for absorption and regeneration are generally vertical cylinders with a maximum diameter of 12.5 feet to enable rail shipment and a maximum length-to-diameter ratio of four (Kasper, 1988). Research has indicated that maximum absorption occurs at temperatures between 1000°F (540°C) and 1200°F (650°C) (Underkoffler, 1986). For best results, a maximum space velocity of 2,000/hr should also be observed. Space velocity is defined as the volume of gas fed to the reactor at standard temperature and pressure (0°C, 1 atm) per volume of sorbent per unit time. The sorbent should be replaced periodically under normal operating conditions due to attrition (Kasper, 1988).

2.4 Sulfur Recovery

The sulfur dioxide-containing offgas from hot gas desulfurization systems must be handled in an environmentally acceptable manner. The sulfur dioxide is typically converted to elemental sulfur or to sulfuric acid. Historically, between 85 and 90 percent of all elemental sulfur consumed in the United States has been converted first to sulfuric acid. However, the market for sulfuric acid appears to be diminishing. Sulfuric acid is very expensive to store and can be costly and dangerous to transport. Therefore, unless the energy plant needs sulfuric acid on-site or the demand is high at a nearby site, it is more advantageous and profitable to convert the SO₂ to elemental sulfur (BRSC, 1987).

Sulfur is the second most consumed mineral in the world (after coal). Conversion of SO₂ to elemental sulfur is both environmentally and economically desirable. Elemental sulfur is easily stored and transported, and it is a valuable industrial commodity. The most

common method of converting hydrogen sulfide and sulfur dioxide to elemental sulfur is use of a Claus plant. Elemental sulfur is typically produced by reacting H_2S and SO_2 in the presence of a catalyst via the Claus reaction:

Since the regeneration offgas from conventional cold gas desulfurization processes contains H_2S at fairly high concentrations and contains no SO_2 , part of the feed stream is usually combusted to form SO_2 by the following reaction:

Thus, the overall reaction in a typical Claus plant is:

A typical Claus plant consists of two reactors, each followed by a sulfur condenser to remove elemental sulfur. The first reaction is rapid and occurs at a high temperature of 2300-2500°F (1260-1370°C). The second reaction is much slower and requires several stages of catalytic reactors at 400-500°F (205-260°C). Sulfur recovery in a Claus plant is usually between 91 and 98 percent (SFA, 1987).

Claus plants are adept at handling feed streams with high sulfur concentrations, but they are ineffective for feed streams with less than about 15 percent H_2S . There are some modified Claus processes, such as Selectox, which can handle low sulfur concentrations, but the conversion to elemental sulfur is typically too low (70 to 90 percent). Other processes, such as wet oxidation, produce elemental sulfur in water-based solutions. These processes yield very high conversion rates for low-sulfur feed streams, but they require special reagents such as vanadium salts and anthraquinone disulfonic acid (ADA). They are also quite expensive, especially if the reagent consumption is high. Therefore, wet oxidation processes are used mostly to further clean tail-gases from Claus plants if the sulfur recovery is not high enough to meet environmental and operational standards. The most widely used wet oxidation processes are Stretford, Takahax, Unisulf, and LO-CAT (SFA, 1987).

3.0 DIRECT SULFUR RECOVERY PROCESS

Since 1988, researchers at the Research Triangle Institute (RTI) have been developing a process to convert feed streams with low sulfur concentration to elemental sulfur. The Direct Sulfur Recovery Process (DSRP) appears to be capable of effectively converting the SO_2 in the regenerator offgas from a zinc ferrite-based hot gas desulfurization unit to elemental sulfur. Figure 3 is a schematic diagram of the DSRP unit. The DSRP promises to be an efficient way to lower sulfur emissions, but to date it has only been tested on a bench-scale. The major advantage of the DSRP over conventional Claus processes is the match in operating conditions between the DSRP and the hot gas cleanup portions of an IGCC plant. The DSRP operates at the same temperature and pressure as the zinc-ferrite desulfurization process. Using a proprietary catalyst, the DSRP can accommodate feed streams with low sulfur concentrations (as low as 0.85 percent by volume) with no pre-concentration necessary. This makes it especially attractive since the regenerator offgas from many hot gas desulfurization systems typically contains only one to three percent SO_2 . The capital and operating costs for a two-stage DSRP unit also appear to be low compared with conventional elemental sulfur recovery processes (Notestein, 1990).

3.1 DSRP Bench Scale Experimental Apparatus

RTI has been conducting bench-scale testing on the DSRP since 1990. The DSRP bench-scale unit consists of four subunits: gas delivery, Reactor No. 1, Reactor No. 2, and gas conditioning/pressure control. Each reactor is followed by a sulfur condenser to remove elemental sulfur, and a reheater follows the first condenser to heat the gas back to an appropriate temperature for Reactor No. 2. The gas delivery system is capable of simulating regeneration offgas mixtures of up to 12 percent SO_2 , and it is also capable of simulating typical gasifier effluent streams. Up to one liter of catalyst may be contained within each reactor. The reactors may be configured with space velocities up to 10,000 scc/(cc·hr).

The fixed-bed reactors for the bench-scale DSRP unit are built from four inch schedule 160 316H stainless pipe. The reactors are approximately 24 inches in height. The catalyst is held in a removable catalyst cage made of stainless steel. The catalyst bed is supported by a porous ceramic or perforated plate gas distribution. One liter of catalyst occupies nine inches of packing height. Figure 4 shows the structure of the DSRP reactors (Gangwal, 1990). Typical sulfur recovery values for the fixed-bed reactors are approximately 95 percent in Stage I and 80 to 90 percent in Stage II. Recent tests have also

Figure 3. Simple Schematic Diagram of the Direct Sulfur Recovery Process (DSRP) Bench-Scale Unit

Figure 4. Cross-Sectional View of a DSRP Bench-Scale Reactor

been conducted using a fluidized-bed reactor for Stage I and a fixed bed reactor in Stage II. Fluidization of the catalyst bed provides better temperature control over the exothermic SO₂ reduction reactions. Overall sulfur recoveries as high as 99.1 percent have been obtained with this reactor configuration and interstage water removal (Gangwal, 1993).

Once the elemental sulfur-laden gas leaves Reactor No. 1, it enters a sulfur condenser (see Figure 5) where it is cooled by surrounding steam at about 250°F (120°C) in a coil. The cooled gas is then flashed into the sulfur collection pot where gas and sulfur are separated. The elemental sulfur may then be drained. The separated gas goes to the reheater where its temperature is raised to a level required for the reactions in Reactor No. 2. Sulfur recovery as high as 96 percent has been observed after the first reactor. However, the overall percent recovery is often decreased in Reactor No. 2. This is due to insufficient separation of gas and sulfur in the condensers. If all the elemental sulfur exiting Reactor No. 1 is not removed before the gas enters Reactor No. 2, the reverse Claus reaction may occur and the process can become counter-productive. Improvements need to be made to the condensers to achieve sufficient sulfur recovery of more than 99 percent (Gangwal, 1992). The researchers at RTI have recently devised a technique that can be used to completely condense the sulfur interstage, but information regarding this technique has not yet been disclosed (Gangwal, 1993).

Once the gas has been through both Claus reactors and the elemental sulfur has been recovered, the gas enters the gas conditioning subsystem. This system is unique to the bench scale test apparatus, and is required to remove pollutants from the gas stream prior to discharge to the atmosphere. First a high pressure Dryrite trap removes essentially all water vapor from the stream. Particulate matter is removed by a Balston filter, then the gas enters a back pressure regulator to reduce pressure. Finally, the nearly atmospheric pressure gas will pass through a two-stage NaOH scrubber to remove any remaining H₂S and SO₂ prior to venting. If the mass output of sulfur dioxide in the vented stream is still too high, this tail-gas may need further treatment (Gangwal, 1990). The goal for the DSRP is to achieve high enough conversions such that tail-gas treatment will be unnecessary. As part of integration with an IGCC system, the DSRP tail gas may be combined with the coal gas stream prior to gas turbine combustion (e.g., Robin et al., 1993).

More recent experimental work has focused on the development of a skid-mounted bench scale zinc titanate fluidized bed desulfurization and DSRP reactor system. The skid-mounted system will be employed in testing with coal gas slip streams from larger scale coal gasification facilities. Future plans include the design, fabrication, and commissioning of a larger, pilot-plant scale DSRP reactor system. The pilot-plant system will be designed

- A, G: 1/2" Female Weld Fitting
- B: SS 3/8" Tubing, 0.035" Wall
- C: Blind Flange
- D, F: 3/8" Weld Fitting
- E, H, I: 3/8" Bored-through Union

Figure 5. DSRP Bench-Scale Sulfur Condenser

for gas volume flow rates six times larger than for the bench-scale design (Gangwal et al., 1994).

3.2 DSRP Process Chemistry

In bench-scale tests, the feed to the DSRP typically consisted of a simulated SO₂ offgas from the regenerator of a zinc ferrite desulfurization process and a stream of reducing gas (H₂S, H₂, CO). The flowrates were adjusted so that there was a 2 to 1 ratio of reducing gas to SO₂. Since the raw coal gas effluent from a gasifier contains H₂S, a slip stream of this gas could be used as the source of the reducing gas. This mixture of offgas and reducing gas was reacted in the presence of a catalyst at a pressure of 20 atm and temperature of 450-700°C in Reactor No. 1, which can operate as either a fixed-bed or fluidized-bed catalytic reactor. In the tests reported to date, only the fixed-bed mode has been used extensively. Under these conditions, the following reactions network is postulated:

Reaction (13) is the water gas shift reaction. It is presumed to be at equilibrium as the gas streams enter the reactor. Reaction (14) is a desired conversion of SO₂ to elemental sulfur via reaction with hydrogen. Reaction (15) is desired because it produces the H₂S needed to react with SO₂ in Stage II. Reactions (16) and (17) are governed by catalyst space velocity. At high space velocities (greater than about 6x10⁵ cc/hr·g), only reaction (16) occurs (See Section 4.1.1 for complete explanation). At lower space velocities, both reactions (16) and (17) occur. Reaction (17) becomes more prominent as space velocity is lowered, thus increasing sulfur conversion and lowering the amount of COS that is sent on to Stage II. Reaction (18) is clearly desirable because it converts two gaseous sulfur species to elemental sulfur.

After interstage sulfur condensation, the gas streams were sent to Reactor No. 2. This reactor was always operated in the fixed-bed mode, typically at temperatures between

250 and 300°C and a pressure of about 20 atm. Under these conditions, the following reactions are believed to occur:

Recall that reaction (19) is the Claus reaction. The "x" is typically between 2 and 8 depending on temperature and residence time. Reaction (20) converts any remaining COS to hydrogen sulfide, which is then converted to elemental sulfur via the Claus reaction. Bench-scale tests indicate that reaction (20) proceeds to completion. Due to the reaction mechanism for Stage II, the inlet stream should contain a 2 to 1 ratio of H₂S and COS to SO₂ (Gangwal, 1993).

3.3 DSRP Design Studies

There have been two major published design studies that feature a performance and cost evaluation of the DSRP as applied to several IGCC flowsheets. The first of these is a 1993 study by Texaco which was published in May 1993 (Robin et al., 1993). In this study, the DSRP was evaluated as part of two different hot gas cleanup systems as applied to a Texaco gasifier-based IGCC system. In one case, a moving bed zinc titanate desulfurization system was evaluated, while in the other case a fluidized bed zinc titanate desulfurization system was studied. In both cases, detailed performance and cost data were reported. The performance data are summarized in Tables 1 and 2 for the two Texaco design cases. The tables reference stream numbers which are depicted in Figure 6.

The second design study was developed by Gilbert/Commonwealth, Inc (Buchanan et al., 1994). This study focused on evaluation of gas stream cleanup systems as applied to KRW gasifier-based IGCC systems. The DSRP was evaluated in conjunction with two IGCC designs that featured fluidized bed zinc titanate desulfurization. The difference in the two cases was that one was for an oxygen-blown gasifier, while the other was for an air-blown system. The report by Buchanan et al. does not provide as much detail regarding the DSRP performance and cost as does the report by Robin et al. Therefore, much of the model development effort benefited primarily from the latter report.

Figure 6. Simple Schematic Diagram of the Direct Sulfur Recovery Process (DSRP).

Table 1. DSRP Mass Balance for a Texaco IGCC with a Moving Bed Zinc Titanate Desulfurization System (Robin et al., 1993)^a

Stream Number	Description	Temperature (°F)	Pressure (psia)	Flow Rate (lb/hr)
1	Regenerator Off-Gas	1,500	423	88,250
2	Coal Gas Slip Stream	1,496	476	10,420
3	Reactor No. 1 Inlet	962	411	98,650
4	Reactor No. 1 Outlet	1,292	386	95,900
5	Condenser No. 1 Outlet	320	376	95,900
6	Reheater Outlet	544	376	96,100
7	Reactor No. 2 Outlet	550	351	95,900
8	Condenser No. 2 Outlet	320	341	95,900
9	DSRP Off-Gas to Gas Turbine	320	341	61,750
10	DSRP Off-Gas to Other Areas	320	341	34,100
11	Elemental Sulfur from Cond. No. 1	320	376	2,750
12	Elemental Sulfur from Cond. No. 2	320	341	220
13	Total Elemental Sulfur			2,970

^a The regenerator off-gas and coal gas slip stream are cooled in a heat recovery steam generator prior to going to the DSRP system.

Figure 6 (repeated). Simple Schematic Diagram of the Direct Sulfur Recovery Process.

Table 2. DSRP Mass Balance for a Texaco IGCC with a Fluidized Bed Zinc Titanate Desulfurization System (Robin et al., 1993)^a

Stream Number	Description	Temperature (°F)	Pressure (psia)	Flow Rate (lb/hr)
1	Regenerator Off-Gas	1400	473	45,140
2	Coal Gas Slip Stream	1250	454	7,810
3	Reactor No. 1 Inlet	791	444	52,950
4	Reactor No. 1 Outlet	1292	419	50,190
5	Condenser No. 1 Outlet	350	409	50,190
6	Reheater Outlet	547	409	50,190
7	Reactor No. 2 Outlet	550	384	50,040
8	Condenser No. 2 Outlet	350	374	50,040
9	DSRP Off-Gas to Gas Turbine			
10	DSRP Off-Gas to Other Areas	349	523	46,750
11	Elemental Sulfur from Cond. No. 1	350	409	2,750
12	Elemental Sulfur from Cond. No. 2	350	374	150
13	Total Elemental Sulfur			2,900

^a The regenerator off-gas and coal gas slip stream are cooled a heat recovery steam generator prior to going to the DSRP system.

4.0 PERFORMANCE MODEL

There are several key factors which must be fully understood in order to create a model to simulate the operation of the DSRP. The most obvious of these are the specific reaction mechanisms which occur in the Stage I and Stage II reactors. Understanding the reaction mechanisms makes it possible to predict outlet compositions given a certain feed stream. Another important design consideration is the catalyst requirement of the reactors. It is necessary to understand the dependence of catalyst volume on key design and performance characteristics. Since space velocity is a key measure of catalyst size, an analysis of variables affecting space velocity is required to develop simple models for estimating catalyst requirements. This performance model will focus on these areas.

4.1 Modeling Reaction Mechanisms

One of the most important steps in modeling the performance of any chemical process is understanding reaction mechanisms and conversion rates. A typical DSRP inlet stream contains a mixture of hot gas desulfurization off-gas products and a slip stream of coal gas to supply the reducing agents required for the DSRP reactions. Thus, the components included in the mixed gas stream are SO_2 , H_2 , CO , CO_2 , H_2O , COS , H_2S , O_2 , and N_2 in varying amounts. The goal of the reaction model is to estimate the effluent composition given a certain inlet composition and reaction conditions.

The water gas shift reaction, Equation (13), is assumed to be at equilibrium at the entrance to Reactor No. 1. This condition can be modeled using standard features of the ASPEN chemical process simulator.

The reaction model described here concentrates on the other five Stage I reactions. RTI developed an equilibrium model for the DSRP reactors. This model predicted experimental reaction results extremely poorly. With this in mind, RTI also developed an equilibrium-modified kinetic reaction model. This model assumed the water gas shift reaction to be at equilibrium and the other Stage I reactions to be governed by kinetics. This model fell close to the experimental results, especially for the pressure range of interest (around 20 atm). An even more rigorous model including water gas shift kinetics would likely further improve model predictions (Gangwal, 1993). Unfortunately, only the results of the RTI model were reported and not the data acquired through kinetic testing. Due to lack of kinetic reaction data, a kinetic-based model was not a plausible approach for model development here. Therefore, a parametric stoichiometric reaction model was

created to enable the examination of the relative conversions of all relevant DSRP reactions through sensitivity analyses.

Several assumptions were made in order to create a simplified model. Key parameters of the model are the conversion efficiency of SO_2 to elemental sulfur in each reactor. Most of the testing done indicated a typical sulfur capture fraction of .90 to .95 for the Stage I reactor and .80 to .90 for the Stage II reactor. Research also indicates that there should be a 2 to 1 ratio of H_2S and COS to SO_2 at the entrance to Stage II, so this constraint was applied to the reaction model. Since no reactions are assumed to occur between the outlet of Reactor 1 and the inlet of Reactor 2, the Stage I effluent and Stage II inlet stream are used interchangeably. All elemental sulfur captured in Stages I and II is assumed to be completely condensed at the reactor outlets. The validity of this assumption is examined in a later section.

4.1.1 Reactor 1 Mass Balance

The concentration of SO_2 leaving Reactor 1 is estimated based on the inlet concentration and SO_2 removal efficiency (please see Nomenclature section):

$$[\text{SO}_2]_o = (1 - \eta_{\text{SO}_2, I}) [\text{SO}_2]_i \quad (21)$$

However, because of the reagent requirements of the Stage II reactor, the following constraint applies:

$$2 [\text{SO}_2]_o = [\text{H}_2\text{S}]_o + [\text{COS}]_o \quad (22)$$

As a result of this constraint, the Stage I outlet H_2S requirement is:

$$[\text{H}_2\text{S}]_o = 2 (1 - \eta_{\text{SO}_2, I}) [\text{SO}_2]_i - [\text{COS}]_o \quad (23)$$

The concentration of H_2S leaving the Stage I reactor is given by:

$$[\text{H}_2\text{S}]_o = [\text{H}_2\text{S}]_i + [\text{H}_2\text{S}]_G \quad (24)$$

where $[\text{H}_2\text{S}]_G$ is the concentration of H_2S generated by Reaction (15). It is assumed that all of the COS that enters the reactor is converted to elemental sulfur via reaction (18). Therefore, any COS leaving the reactor is produced within the reactor by Reaction (16):

$$[\text{COS}]_o = [\text{COS}]_G \quad (25)$$

The molar CO_2/COS ratio, R , at the outlet of Reactor No. 1 is an indication of the relative importance of Reactions (16) and (17). In kinetic and thermodynamic studies of the DSRP process chemistry, RTI estimated the molar CO_2/COS ratio as a function of catalyst space velocity for an $\text{SO}_2\text{-CO}$ reaction system (Gangwal, 1993). The relationship

of this ratio to space velocity is shown in Figure 7. For low space velocities, which are indicative of large catalyst volumes, the ratio is high. For space velocities above about 6×10^5 cc/hr·g, the ratio becomes nearly constant at a value near two.

By combining reactions (16) and (17), the following reaction is obtained:

where a, b, d, e, f are functions of the ratio R. The resulting stoichiometric equation, expressed as a function of R, is:

Given a value of the CO_2/COS ratio, Reaction (27) makes it possible to determine the contributions of reactions (16) and (17).

Although the data in Figure 7 may be useful for identifying qualitative trends, it is uncertain how applicable these results are to a large scale system due to the testing conditions used. The data were obtained from kinetic testing in a quartz-lined micro reactor containing approximately .5 g of catalyst. The reaction system contained only CO and SO_2 rather than a typical DSRP feed stream. The space velocities also may differ from those used in a large scale commercial system. For comparison, the space velocities on the curve can be converted to the same units as those used in bench-scale tests. This was accomplished by converting the reported space velocity from a catalyst mass basis (e.g., cc/hr·g) to a catalyst volume basis, assuming a catalyst density of about 0.5 g/cc (estimate obtained through personal communication with RTI). Using this conversion factor and adjusting for standard temperature and pressure, the space velocities tested were considerably higher (greater than about 200,000 scc/cc·hr) than those typical for DSRP bench-scale testing (less than about 5,500 scc/cc·hr). Thus, the results of Figure 7 are not directly applicable to commercial scale designs. Therefore, the CO_2/COS ratio is treated as a performance model parameter that can be adjusted by the user.

Reaction (27) can be used to determine the amount of COS generated in Stage I. Assuming that all inlet CO reacts in this system, since it is the only reaction system which consumes CO in the Stage I reactor, the amount of COS generated is:

$$[\text{COS}]_G = 1/(R+1) [\text{CO}]_i = [\text{COS}]_0 \quad (28)$$

Using the mass above, it is possible to determine the contributions of reactions (14) through (18) to the overall SO_2 conversion. The amount of SO_2 converted by reaction (15) should be equal to the amount of H_2S generated by this reaction. Since this reaction is

Figure 7. Plot of R vs. SV for CO-SO₂ System

the only reaction generating H_2S in the mechanism, this amount must be equal to the required H_2S outlet concentration minus the inlet concentration. Combining equations (23), (24), and (27), the efficiency of reaction (15) in removing SO_2 can be written as:

$$\eta_{15} = \frac{2(1-\eta_{\text{SO}_2})[\text{SO}_2]_i - [\text{H}_2\text{S}]_i - \frac{1}{R+1}[\text{CO}]_i}{[\text{SO}_2]_i} \quad (29)$$

In a similar manner, the efficiency for the system of reactions (16) and (17) can be written as:

$$\eta_{16-17} = \frac{\frac{R}{2(R+1)}[\text{CO}]_i}{[\text{SO}_2]_i} \quad (30)$$

Reaction (18) converts all inlet COS to elemental sulfur. The concentration of SO_2 consumed by this reaction will be equal to half of the inlet COS concentration as dictated by stoichiometry. The contribution of reaction (18) to the overall conversion of SO_2 to elemental sulfur in Stage I is:

$$\eta_{18} = \frac{\frac{1}{2}[\text{COS}]_i}{[\text{SO}_2]_i} \quad (31)$$

Thus, the contribution of reaction (14) can be determined by difference:

$$\eta_{14} = \eta_{\text{SO}_2, \text{I}} - (\eta_{15} + \eta_{16-17} + \eta_{18}) \quad (32)$$

4.1.2 Stage I Sulfur Condenser

The hot gas stream (typically 450 to 650°C) leaving the Stage I reactor enters a sulfur condenser where it is cooled in a heat exchanging coil surrounded by steam at about 120°C. There are no chemical reactions in this unit, but there is a change in the content of the stream. The elemental sulfur produced in the Stage I reactor is condensed and removed from the stream. Some of the water content is often also removed by this process. In the DSRP bench-scale tests, sulfur condensation has been less than 100 percent efficient. However, recent work at RTI has focused on developing a new technique which appears to be capable of condensing and removing all of the elemental sulfur before the stream is sent to Stage II. Information regarding this technique has not yet been disclosed. However, based on this development, it is assumed that 100 percent of the elemental sulfur exiting both the Stage I and Stage II reactors is condensed.

4.1.3 Interstage Reheater

The gas stream leaving the condenser is typically at a temperature of 150 to 200°C. It must be reheated to an appropriate temperature for the Stage II reactor, which generally operates most effectively at 250 to 300°C. This is accomplished by contacting the gas stream with superheated steam in a heat exchanger. There are no chemical or physical changes to the gas stream.

4.1.4 Reactor II Mass Balances

Modeling Stage II is considerably less complex because only two reactions are of concern. Testing has indicated that reaction (20) proceeds to completion, so all COS entering Stage II is converted to H₂S. This H₂S and entering H₂S react with entering SO₂ via the Claus reaction (19) to form elemental sulfur. In most of the bench-scale testing 80 to 90 percent of SO₂ entering Stage II was converted. As stated earlier, it is assumed that no interstage chemical reactions occur. Therefore, the SO₂ concentration at the entrance of the Stage II reactor is the same as the SO₂ concentration at the exit of the Stage I reactor:

$$[\text{SO}_2]_{\text{II},i} = [\text{SO}_2]_{\text{I},o} \quad (33)$$

Assuming a conversion efficiency ($\eta_{\text{SO}_2,\text{II}} = .8$ to $.9$), the outlet SO₂ concentration from Stage II is:

$$[\text{SO}_2]_{\text{II},o} = (1 - \eta_{\text{SO}_2,\text{II}})[\text{SO}_2]_{\text{II},i} \quad (34)$$

All inlet COS reacts to generate an equal concentration of H₂S:

$$[\text{H}_2\text{S}]_{\text{II},G} = [\text{COS}]_{\text{II},i} \quad (35)$$

$$[\text{COS}]_{\text{II},o} = 0 \quad (36)$$

Reaction (19) shows that H₂S reacts with SO₂ at a 2:1 mole ratio to form elemental sulfur. Enough H₂S will react to produce the required conversion efficiency. Using this knowledge, the outlet concentration of H₂S is:

$$[\text{H}_2\text{S}]_{\text{II},o} = 2(1 - \eta_{\text{SO}_2,\text{II}})\{[\text{H}_2\text{S}]_{\text{II},i} + [\text{H}_2\text{S}]_{\text{II},G}\} \quad (37)$$

Using these mass balances and the mass balances for the Stage I reactor, the stoichiometric model was implemented using the Excel spreadsheet program. This model can process a given inlet stream composition and reaction specifications and predict the outlet concentrations of key reaction species for both Stages I and II. Table 3 shows an analysis of a possible sample DSRP inlet stream (after the water gas shift has reached

Table 3. Reaction Model Using Hypothetical Inlet Composition

Performance Model Input Parameters (must be specified)

Stage I Inlet Concentrations (vol %)	
[SO ₂]	1.64%
[CO]	0.50%
[H ₂]	2.80%
[COS]	0.10%
[H ₂ S]	0.08%
Ratio: [CO ₂]/[COS]=R	2
Efficiency of SO ₂ Conversion (Stage I)	0.92
Efficiency of SO ₂ Conversion (Stage II)	0.85

Stage I Calculated Variables

Fraction of Inlet SO ₂ Consumed by:	
Reaction 14	0.778
Reaction 15	0.010
Reactions 16-17	0.102
Reaction 18	0.030

Stage II Inlet Variables (calculated)

Stage I Outlet = Stage II Inlet (vol %)	
[SO ₂]	0.1312%
[COS]	0.1667%
[H ₂ S]	0.0957%

Stage II Calculated Outlet (vol %)	
[SO ₂]	0.0197%
[H ₂ S]	0.0787%
[COS]	0.0000%

equilibrium). No DSRP outlet concentrations were reported from the bench-scale testing, so it is not possible to compare these results with actual experimental data at this time.

4.1.5 Stage II Sulfur Condenser

This condenser is the same as the condenser following the Stage I reactor. As noted above, it is assumed that the condensation and removal of elemental sulfur from the gas stream is complete.

4.1.6 Gas Conditioning/Pressure Control

Very little physical information or test data are available for these final DSRP units. All bench-scale testing has concentrated on the operation of the reactors and condensers. It is even possible that these units are unique to the bench-scale DSRP unit and would not be applicable to or necessary for a commercial scale unit.

4.2 Estimating Space Velocities

In the early stages of this project, an effort was made to develop an empirical statistical model that relates catalyst space velocity to key performance variables. The results of this work are reported in Section 4.2.1. The statistical analysis provides some insight into factors that may need to be considered in choosing a space velocity as part of the design of a new system. However, due to the evolving nature of this technology, and particularly due to improvements in catalyst formulation, assumptions regarding space velocity are changing. Therefore, the performance model that is implemented in ASPEN includes space velocity as a user-specified parameter. The basis for selecting a parameter value, based on currently available information, is described in Section 4.2.2.

4.2.1 Statistical Models for Space Velocity

In order to size a catalytic reactor for the Stage I and Stage II reactors, it is necessary to know the dependence of catalyst volume on key design or performance characteristics. Space velocity is a key measure of catalyst size. To gain insight into the key factors which influence the catalyst size, space velocity data for bench-scale tests were analyzed. Using data reported by McMichael and Gangwal (1991) for bench-scale tests with both Stage I and Stage II reactors in the fixed-bed mode, a number of alternative linear regression models were evaluated using the StatWorks program with catalyst space velocity as the dependent variable. The independent variables explored for both reactors included SO₂ removal efficiency (%), temperature (°C), pressure (atm), and inlet concentrations of relevant reaction species. For Stage I the species used were SO₂, CO, and H₂. For Stage

II the species used were SO₂ and H₂S. Although COS plays a key role in the Stage II reactor, none of the reported data sets included COS as a component of the inlet gas stream. RTI evaluated several catalysts for use in both the Stage I and Stage II reactors. Tests showed that Catalyst A provided the best results for Stage I and Catalysts B and C were preferred for Stage II. Therefore only test runs using these catalysts were analyzed.

Regression analyses were performed for space velocity as a function of each independent variable taken individually, as well as for groups of independent variables. The scatter plots of space velocity versus each independent variable for all three catalysts are given in Appendix A. The Stage I plots seemed to show two clusters of points, one for space velocities of about 2,000 scc/cc-hr (low SV) and the other for space velocities of about 5,000 scc/cc-hr (high SV). These groups of data points were re-analyzed separately (see Table 4). The low space velocity data indicated that space velocity depends most heavily on the inlet concentrations of SO₂, CO, and H₂. The statistical results for a linear regression model of space velocity versus these three independent variables gave a coefficient of determination (R² value) of 0.47 and a standard error of about 426 scc/cc-hr. The equation for this linear regression model is:

$$SV_{IA,L} = 5270 - 2390[SO_2]_i + 1350[H_2]_i - 856[CO]_i \quad (38)$$

This model is statistically significant at the .006 confidence level based on the F-test. However, the low value of R² and high standard error indicate that there is a large portion of variance in space velocity that cannot be explained by the linear model. Figure 8 contains a parity plot of results predicted by this model versus experimental values of space velocity. This plot illustrates the shortcomings of this particular model. If the model was perfect, all points would lie on the 45° line. Figure 8 shows that most of the data points are close to this line, but that there is considerable scatter.

At high space velocities, the regression analyses indicated that space velocity depends most heavily on SO₂ removal (%) and the inlet concentrations of SO₂ and CO. The statistical results for a linear regression model of space velocity versus these three independent variables gave a coefficient of determination (R² value) of 0.92 and a standard error of about 91 scc/cc-hr. The equation for this linear regression model is:

$$SV_{IA,H} = 11,000 - 10.1(\eta_{SO_2,I}) - 1490[SO_2]_i - 1100[CO]_i \quad (39)$$

This model is statistically significant at the .000 confidence level based on the F-test. The higher R² value and considerably lower standard error indicate that this is a much better

Table 4. Data for Stage I Regression Analyses

Data for Catalyst A at Low Space Velocity

Space Velocity (scc/cc-hr)	SO ₂ Removal (%)	Temp. (°C)	Pressure (atm)	[SO ₂] (vol. %)	[H ₂] (vol. %)	[CO] (vol. %)
950	96	594	20	1.78	1.52	1.99
950	96.1	554	20	1.78	1.52	1.99
950	96	501	20	1.78	1.52	1.99
1104	94.3	653	20	1.56	1.32	1.73
1104	94.9	599	20	1.56	1.32	1.73
1763	95.2	555	20	1.78	1.57	2.04
1763	94.4	597	20	1.78	1.57	2.04
1954	92.6	448	20	1.65	2.11	1.93
2004	93.5	644	20	2.01	2.15	1.97
2004	94.6	547	20	2.01	2.15	1.97
2004	84	455	20	2.01	2.15	1.97
2004	80.8	650	7	2.01	2.15	1.97
2159	93.4	647	20	1.82	1.89	1.73
2186	95.8	557	20	1.8	1.86	1.71
2212	92.3	640	20	1.82	1.89	1.73
2231	90.2	453	20	1.48	1.29	1.69
2235	83.7	447	20	1.72	1.83	1.68
2239	93.3	557	20	1.8	1.86	1.71
2279	95.7	551	20	1.72	1.83	1.68
2369	96.1	505	20	1.43	1.21	1.59
2430	96.4	599	20	1.38	1.03	1.35
2545	96.6	550	20	1.3	1.13	1.48
2787	96.9	650	20	1.38	1.03	1.35

Data for Catalyst A at High Space Velocity

Space Velocity (scc/cc-hr)	SO ₂ Removal (%)	Temp. (°C)	Pressure (atm)	[SO ₂] (vol. %)	[H ₂] (vol. %)	[CO] (vol. %)
4849	92.4	550	20	2.01	2.13	1.95
5000	93.1	649	20	2.02	2.07	1.89
5257	95.2	598	20	1.72	1.52	1.96
5333	94	646	20	1.89	1.93	1.77
5333	89.2	543	20	1.89	1.93	1.77
5440	94	648	20	1.66	1.46	1.89
5506	93.1	544	20	1.64	1.45	1.87
5532	93.4	644	20	1.7	1.32	1.73
5650	90.4	644	20	1.67	1.45	1.9
5650	84.9	601	20	1.67	1.45	1.9
5650	69.9	546	20	1.67	1.45	1.9

Figure 8. Comparison Between Experimental Values and Values of Space Velocity Predicted by the Model for Stage I at Low Space Velocities

Figure 9. Comparison Between Experimental Values and Values of Space Velocity Predicted by the Model for Stage I at High Space Velocities

model than for the low space velocity case above. Figure 9 contains a parity plot of results predicted by this model versus experimental values of space velocity. This plot reinforces the validity of this model. All data points lie quite close to the 45° line, indicating that the model is good at predicting experimental values.

Similar to the Stage I data, the data for Stage II test runs indicated clusters for low and high space velocities. However, there were not enough high velocity data points to create a statistical model. Therefore only low space velocities (around 2000-2300 scc/cc-hr) were used in the regression models (see Table 5). For the tests using Catalysts B and C at low space velocities, it appears that space velocity is strongly correlated with the inlet concentrations of SO₂ and H₂S. Since these values are collinear ([H₂S] = 2[SO₂]), it is not meaningful to create a model in which space velocity is a function of both. Therefore only inlet SO₂ concentration was used for the model. Space velocity is plotted as a function of SO₂ concentration in Figure 10 for Catalyst B and Figure 11 for Catalyst C. Also shown in the figures are the linear regression models. In Figure 10 there are essentially only two clusters of data. Thus, it is uncertain if a linear model is appropriate in this case. In Figure 11, there are three clusters of data which fall on the line, indicating greater confidence in the use of a linear model. The model using Catalyst B gave a coefficient of determination of 0.93 and standard error of about 43 scc/cc-hr. The equation for this linear regression model is:

$$SV_{IIB} = 3950 - 30,850[SO_2]_{II,i} \quad (40)$$

The model using Catalyst C gave a coefficient of determination of 0.997 and standard error of about 8.7 scc/cc-hr. The equation for this linear regression model is:

$$SV_{IIC} = 4090 - 32,600[SO_2]_{II,i} \quad (41)$$

Both of these models are statistically significant to the .000 confidence level based on the t-test. The high R² values and low values for standard error indicate that these models are good representations of space velocity dependence in Stage II.

4.2.2 Recent Space Velocity Estimates

As part of on-going work, Research Triangle Institute is conducting a series of slip-stream tests at METC's fluidized bed gasifier using actual coal gas. The objectives of the tests are to demonstrate the operation of fluidized bed desulfurization sorbents, the DSRP, and high-temperature catalytic ammonia decomposition. The test apparatus is a mobile laboratory which is comprised of a trailer containing a small-scale DSRP unit. The DSRP is reported to convert SO₂ in actual or synthetic tail-gas to elemental sulfur, achieving 95.1

Table 5. Data for Stage II Regression Analyses

Data for Catalyst B at Low Space Velocity

Space Velocity (scc/cc-hr)	SO ₂ Removal (%)	Temp. (°C)	Pressure (atm)	[SO ₂] (vol. %)	[H ₂ S] (vol. %)
1981	95.6	400	20	0.0619	0.1262
1981	99.5	248	20	0.0619	0.1262
1981	90.1	324	7	0.0639	0.1223
1981	96.7	254	7	0.0639	0.1223
2044	94.1	323	20	0.0619	0.1262
2088	78.9	396	7	0.0639	0.1223
2308	77.7	393	20	0.0531	0.1083
2308	79.1	321	20	0.0531	0.1083
2308	84	260	20	0.0531	0.1083
2308	47	395	7	0.0531	0.1083
2308	45.4	322	7	0.0531	0.1083
2308	62.9	247	7	0.0531	0.1083
2337	73.2	321	20	0.0537	0.107
2337	75.5	250	20	0.0537	0.107

Data for Catalyst C at Low Space Velocity

Space Velocity (scc/cc-hr)	SO ₂ Removal (%)	Temp. (°C)	Pressure (atm)	[SO ₂] (vol. %)	[H ₂ S] (vol. %)
2004	78.8	505	20	0.064	0.128
2004	90	450	20	0.064	0.128
2004	93.6	403	20	0.064	0.128
2133	54.8	600	20	0.06	0.12
2140	74.1	549	20	0.06	0.12
2157	51.7	645	20	0.06	0.12
2353	65.1	325	20	0.0533	0.1063
2353	66.5	255	20	0.0533	0.1063

Figure 10. Regression Plot for Stage II Using Catalyst B at Low SV

Figure 11. Regression Plot for Stage II Using Catalyst C at Low SV

to 99.5 percent conversion after the first stage of a two-stage system. As part of tests to be conducted later in 1995, a single-stage DSRP reactor system will be used in long-term testing. The purpose of the long-term testing is to assess new control logic and to evaluate the longer-term effect of trace contaminants in the coal gas. A larger pilot-scale DSRP unit is also planned. The pilot-scale unit will be six times larger than the mobile laboratory unit, and will be tested at Enviropower's 15 MW (thermal) gasifier, beginning in late 1995.

The mobile laboratory DSRP system has been operated with a Stage I space velocity of 4,560 scc/(cc-hr), in a temperature range of 550 to 610 °C, and at a pressure of 260 psi. The conversion of SO₂ to sulfur has been high, while the conversion of SO₂ to H₂S has been low (e.g., 1.2 to 4.9 percent of the inlet sulfur). In Stage II of the mobile laboratory system, it is believed that the reverse Claus reaction may be occurring, thereby reducing the overall sulfur production of the DSRP. For this reason, a one-stage design is preferred for future testing (Portzer, 1995).

4.2.3 Discussion

The regression analyses of space velocity provided insight regarding key parameters that affect space velocity. For example, the statistical analysis confirms that SO₂ removal efficiency decreases with increased space velocity. RTI tests indicated that temperature had little effect over a range of about 500 to 650°C for Stage I.

There are some questions raised by the regression analyses which cannot be answered at this time. It is unknown why the system appears to behave differently at low space velocities than it does at high space velocities. Also, space velocity is expected to be dependent on pressure. There were not enough data available to test this dependence, as almost all test runs were conducted at 20 atm of pressure.

Furthermore, space velocity also depends on catalyst formulations. The parametric test data which were used in the regression analyses represent one set of alternative catalyst formulations. The regression analysis results cannot be applied to other catalyst formulations. However, they do provide useful qualitative insights into the behavior of the DSRP system.

Because catalyst formulation is changing, and due to the limited availability of parametric data which can be used to related catalyst space velocity to other performance variables, catalyst space velocity for each stage of the DSRP is treated as a user-specified input variable.

4.3 Implementation of the Performance Model in ASPEN

The performance model of the DSRP has been implemented into ASPEN. A listing of the performance model code is given in Appendix B. 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, it is possible to include Fortran programs within the simulation models as "Fortran blocks," or to call external Fortran subroutines. ASPEN contains an extensive physical property database, which allows a modeler to include a wide range of chemical species in the model. ASPEN also includes convergence algorithms for calculating results in closed loop systems, which are modeled using "design specification." A design specification causes part or all of the simulation model to be executed iteratively, varying the value of a selected flowsheet input variable to achieve a specified value of a key design variable within a given tolerance. These combinations of features make ASPEN a powerful tool for process simulation.

Many advanced coal-based power generation technologies have been modeled by DOE using ASPEN. The DSRP performance model has been implemented as part of a previously developed and modified flowsheet for an air-blown Lurgi-gasifier-based IGCC system (Frey and Rubin, 1990, 1991, 1992).

The DSRP performance model consists of 15 unit operation blocks and three FORTRAN blocks, as illustrated in Figure 12. The key features of this model are described here.

The offgas from the high temperature, high pressure gas cleanup system enters the DSRP system via stream SO2PROD. This gas may typically have a temperature of approximately 1,200 °F to 1,500 °F. A portion of the clean fuel gas from the gas cleanup system, ZF-CLFL (which stands for "Zinc Ferrite-CLean Fuel"), is split in unit operation block REDUSPLT into a slip stream DSLIP and a fuel gas stream TOGTMIXR. The fuel gas stream later joins with the tail gas from the DSRP system before being fired in the gas turbine combustor. The slip stream mass flow is determined by FORTRAN block SET-RSPLT. This FORTRAN block ensures that the molar flow rate of the sum of CO and H₂ in the slip stream will be twice the molar flow rate of SO₂ in the offgas. The slipstream

Figure 12. Schematic Diagram of the ASPEN Performance Model for the DSRP

DSLIP is mixed with SO₂PROD in a mixer unit operation block DSRP-MIX. The outlet of this block is DCOOLIN.

The stream DCOOLIN enters a heater unit operation block DSRP-COL, which represents an inlet gas precooler. The heat released by cooling the gas may be used to heat high temperature streams such as in a superheater. As noted in Section 3.2, it is expected that the water-gas shift reaction will be at equilibrium as the gas stream enters the first DSRP reactor. Thus, the water-gas shift equilibrium reaction is modeled using the Gibbs free energy minimization optimization code contained in the RGIBBS unit operation block labelled DSRP-SHF. The outlet stream from this block, DR1IN, enters the first reactor, which is modeled as a stoichiometric chemical reaction in block DSRP-R1. The outlet temperature of the reactor is set to a specified temperature of 1,292 °F, which matches the maximum values assumed in a design study of the DSRP as applied to Texaco gasifier-based IGCC systems with hot gas cleanup (Robin et al., 1993). The specified temperature may be changed by the user. The specified reactor outlet temperature is achieved by varying the outlet temperature from the inlet gas precooler such that the heat transferred from the first reactor is set to zero. The variation of outlet temperature for the precooler is achieved using the design-specification DS-DSRPC. The design specification allows for the temperature change of the gas stream associated with the water-gas shift reaction in DSRP-SHF and with the exothermic chemical reactions in DSRP-R1 to be accounted for to ensure that sintering of the catalyst does not occur.

The FORTRAN block SET-DCNV is used to calculate the reaction conversion rates for the reactions that occur in both of the DSRP reactor vessels, which are modeled using the stoichiometric reaction blocks DSRP-R1 and DSRP-R2. The user may specify the SO₂ conversion efficiency for each of the reactors. Using the relationships developed in Equations (29) through (32), the conversion rates of each of five reactions occurring in DSRP-R1 and each of two reactions occurring in DSRP-R2 are calculated.

The gas leaving Reactor No. 1, DC1IN, enters a heater unit operation block, DSRP-C1, which represents the first condenser. The outlet temperature for this condenser may be specified by the user, and typically would have a value of approximately 350 °F. The heat stream from the condenser, QDSRP-C1, is used in an economizer. The separation of elemental sulfur is represented by a separate unit operation block, DSRP-SP1, in which the solid elemental sulfur is removed in stream DES1 and the remaining gas is contained in stream DRHIN. The stream DRHIN enters reheater, which is modeled by the heater unit operation block DSRP-RHX. Design studies by Robin et al. (1993) suggest that a temperature of approximately 550 °F would be typical of the reheater outlet

temperature. The user may set the reheater outlet temperature by changing its value in the unit operation block parameter sentence.

The reheated gas stream, DR2IN, enters the second reactor, DSRP-R2. As previously noted, the reaction conversion rates for this stoichiometric reactor model are determined by the FORTRAN block SET-DCNV. The reactions that occur in DSRP-R2 have only a minor effect on the gas temperature. As an approximation, it is assumed that the outlet gas temperature is the same as the inlet gas temperature. By comparison, studies by Robin et al. (1993) indicate a temperature change of approximately three to six degrees Fahrenheit. The gas stream that leaves the second reactor, DC2IN, enters the unit operation block DSRP-C2, which represents the second condenser. The heat released by the condenser is represented by the heat stream QDSRP-C2. The heat released by the second condenser is assumed to be available as input to the low pressure economizer.

The sulfur removed in the DSRP second condenser is modeled in unit operation block DSRP-SP2, in which the elemental sulfur stream DES2 is separated from the gas stream TAILGAS. The TAILGAS is compressed to a sufficient pressure in the compressor TAILCOMP for mixing with the clean fuel gas in stream TOGTMIXR in unit operation block GTMIXR, which produces the outlet fuel gas stream FUELIN.

The elemental sulfur streams from the two condensers are mixed together in a mixer block DSRP-SM and then cooled to an appropriate product exit temperature in the heater block DSRP-PRD.

The pressures of the gas streams are determined by the inlet pressure of the fuel gas slip stream and specified pressure drops in the gas precooler, Reactors Nos. 1 and 2, Condensers Nos. 1 and 2, and the reheater between Condenser 1 and Reactor 2. These pressure drops were estimated based upon design studies by Robin et al. (1993). They are summarized in Table 6.

The ASPEN performance model is used to estimate the energy requirements for the tailgas booster compressor. All of the work streams for the DSRP are grouped together for integration with the balance of the power systems model.

The heat streams from the DSRP performance model have been grouped into three categories, based upon the approximate temperature at which heat is transferred. There are four heat streams which are grouped as high temperature streams that are associated with an approximate outlet temperature of 1,200 °F. Three heat streams are grouped together as

Table 6. Typical Pressure Drop Assumptions for Key Components of the DSRP

Unit Operation	Typical Gas Stream Pressure Drop (psi)
Inlet Gas Precooler	2
Reactor No. 1	25
Condenser No. 1	10
Reheater	1
Reactor No. 2	25
Condenser No. 2	10

medium temperature streams (approximately 500 °F). One stream is included in the category of low temperature streams (approximately 300 °F).

5.0 DSRP COST MODEL

In this section, analytical models of direct, indirect, and total capital costs are developed for DSRP systems applied to IGCC systems.

5.1 Direct Capital Costs

Direct capital cost models for key components of the DSRP were developed based upon data presented by Robin et al. (1993). Total installed capital costs were reported for components that include Reactors No. 1 and 2, Condensers No. 1 and 2, gas precooling, gas reheat, and sulfur handling. These components can be categorized into two major groups: (1) gas handling; and (2) sulfur handling. Most of the components fall into the gas handling group. For the Case 11F cost estimate as described in Robin et al. (1993), the direct cost for the gas handling components is \$13.1 million, while the direct cost for the sulfur handling components is \$0.4 million. To a first order, the costs of the gas handling components are proportional to the volume flow rate of the gas streams, which in turn depend on pressure and temperature. In addition, the costs of the pressure vessels depend upon the thickness of the pressure vessel walls, which in turn are proportional to pressure. For the sulfur handling systems, the costs are proportional to the mass flow rate of molten sulfur.

A capacity-exponent cost model formulation is employed. Because few cost data are available, it was necessary to make a reasonable assumption regarding the appropriate functional form of the cost model. The capacity-exponent model is widely used for the purposes of developing preliminary, budgetary cost estimates of chemical processes. An exponential scaling factor of two-thirds is typically employed.

To enable evaluation of the costs of both one-stage and two-stage DSRP systems, the costs of the DSRP are broken into separate stages.

A listing of the DSRP cost model code is given in Appendix C.

5.1.1 Adjusting for Pressure

To adjust for differences in cost due to the thickness of pressure vessel walls, a correction factor developed by Frey and Rubin (1990) based upon data presented in Ulrich (1984) is employed:

$$f_p = 2.5 + 0.093 P^{0.64} \quad (42)$$

In the direct cost model for the gas handling section of the DSRP, the correction factor is normalized with respect to the value for the Reactor No. 1 pressure in the base case cost estimate obtained from Robin et al (1993), which is 444 psi. At this pressure, the correction factor has a value of 7.1.

5.1.2 Direct Costs for the First Stage of the DSRP

For each of the two DSRP reactor stages, the costs are divided into gas handling and sulfur handling. For the first stage, gas handling equipment costs include the gas cooler, reactor vessel, and sulfur condenser. The sulfur handling section includes equipment required to handle the sulfur once it has been condensed.

For the gas cooler and sulfur condenser, the costs are assumed to be proportional to the gas volumetric flow rate. The reported costs are \$3.38 million per train of equipment to handle a volume flow rate of 77,000 ft³/hr at 444 psi (Robin et al., 1993). Of this amount, \$1.75 million is attributable to gas cooling upstream of Reactor No. 1, and the balance is due to the sulfur condenser. To adjust for differences in pressure, the pressure correction factor is employed. Because costs are reported in 1993 dollars, the costs are normalized using the *Chemical Engineering* magazine 1993 plant cost index of 361.2.

$$DC_{GH,1} = N_{T,DSRP} \left(\frac{2.5 + 0.093 P_{g,R1,i}^{0.64}}{7.1} \right) \left\{ 3,380 \left(\frac{V_{g,R1,i}}{77,000 N_{O,DSRP}} \right)^{0.67} \right\} \left(\frac{PCI}{361.2} \right) \quad (43)$$

For the Stage I reactor vessel, the cost per vessel is reported to be \$1.02 million in 1993 dollars based upon a gas flow rate of 77,000 ft³/hr and a space velocity of approximately 2,000 scc/cc-hr. Unlike the gas cooler and sulfur condenser costs, the cost of the reactor vessel is proportional not only to the volumetric flow rate of the gas, but also to the volume required for the catalysts. Thus, all other things being equal, a pressure vessel which must hold a larger quantity of catalyst is presumed to cost more than one that must handle the same gas flow rate but with a smaller catalyst volume. Since catalyst volumes are typically expressed using "space velocity", the catalyst volume is typically normalized to the gas flow rate. Therefore, as a measure of reactor vessel size, both the gas volumetric flow rate and the catalyst space velocity are used. The purpose here is to develop a ratio of catalyst volumes, rather than gas flow rate, as the sizing variable for the reactor vessel. The resulting equation is:

$$DC_{R,1} = N_{T,DSRP} \left(\frac{2.5 + 0.093 P_{g,R1,i}^{0.64}}{7.1} \right) \left\{ 1,020 \left[\left(\frac{V_{g,R1,i}}{77,000 N_{O,DSRP}} \right) \left(\frac{SV_{R1,ref}}{SV_{R1,actual}} \right) \right]^{0.67} \right\} \left(\frac{PCI}{361.2} \right) \quad (44)$$

The sulfur handling system includes heaters to maintain the sulfur in a molten state. The costs for these are reported to be \$80,000 per unit for a unit capable of handling 2,900 lb/hr of sulfur. Thus, the cost of the solids handling system for the first stage of the DSRP are:

$$DC_{S,1} = N_{T,DSRP} 80 \left(\frac{m_{s,R1,o}}{2,900 N_{O,DSRP}} \right)^{0.67} \left(\frac{PCI}{361.2} \right) \quad (45)$$

5.1.3 Direct Costs for the Second Stage of the DSRP

For a two-stage system, there are additional direct costs for a reheater, reactor vessel, sulfur condenser, and sulfur handling systems. The reheater is reported to have a 1993 cost of \$480,000 based on a gas volumetric flow rate of 69,000 ft³/hr. The sulfur condenser, which costs \$726,000 per unit, is also sized for the same gas volumetric flow rate. Thus, the total cost for the reheater and sulfur condenser is:

$$DC_{GH,2} = N_{T,DSRP} \left(\frac{2.5 + 0.093 P_{g,R2,i}^{0.64}}{7.1} \right) \left\{ 1,206 \left(\frac{V_{g,R2,i}}{69,000 N_{O,DSRP}} \right)^{0.67} \right\} \left(\frac{PCI}{361.2} \right) \quad (46)$$

For the Stage II reactor vessel, the 1993 cost is \$960,000, based upon a gas flow rate of 69,000 ft³/hr and a space velocity of 2,000 scc/cc-hr. The direct cost, adjusted for other gas flow rates and space velocities, is:

$$DC_{R,2} = N_{T,DSRP} \left(\frac{2.5 + 0.093 P_{g,R2,i}^{0.64}}{7.1} \right) \left\{ 960 \left[\left(\frac{V_{g,R2,i}}{77,000 N_{O,DSRP}} \right) \left(\frac{SV_{R2,ref}}{SV_{R2,actual}} \right) \right]^{0.67} \right\} \left(\frac{PCI}{361.2} \right) \quad (47)$$

The sulfur handling system has approximately the same costs as for Stage II. Thus, the direct cost equation is:

$$DC_{S,2} = N_{T,DSRP} 80 \left(\frac{m_{s,R2,o}}{2,900 N_{O,DSRP}} \right)^{0.67} \left(\frac{PCI}{361.2} \right) \quad (48)$$

5.1.4 Total Direct Costs for the DSRP

The total direct cost of the DSRP is given by the sum of all of the direct cost equations for the gas handling, reactors, and solids handling components:

$$DC_{DSRP} = DC_{GH,1} + DC_{R,1} + DC_{S,1} + DC_{GH,2} + DC_{R,2} + DC_{S,2} \quad (49)$$

For a one-stage DSRP system, the direct costs for all second stage components are modeled as being zero.

5.2 Other Capital Costs for the DSRP

Other capital costs include general facilities, engineering and home office fees, contingencies, various indirect capital costs, preproduction costs associated with startup, and inventory costs associated with providing initial stocks of chemicals and fuels. For more details on other capital cost models, please see Frey and Rubin (1990).

General facilities include roads, office buildings, shops, laboratories, and other equipment or facilities that are not part of other process areas but that are associated with the process technology. General facilities costs are typically estimated as a percentage of the direct capital cost of all other process areas. These costs are handled by the power systems cost model. Like general facilities costs, engineering and home office fees are typically estimated as a percentage of the total direct cost. A typical value of 10 percent of the total direct cost is often assumed for engineering and home office fees.

Project contingency cost is estimated for the entire power plant as a multiplier of total direct cost (e.g., EPRI, 1986). Process contingency costs are typically evaluated separately for each process area. The total process contingency is given by:

$$C_{ProcC} = \sum_i f_{ProjC} DC_i \quad (50)$$

For the DSRP, a typical value for process contingency would be 25 percent, based upon the state of development of the technology. As shown in Table 7, contingency factors of 20 to 35 percent are often employed for technologies for which small pilot plant data are available.

The total plant cost, or overnight construction cost, is given by:

$$TPC = TDC + C_{GF} + C_{EHO} + C_{ProjC} + C_{ProcC} \quad (51)$$

An allowance for funds during construction (AFDC) is calculated based on the TPC as a function of the amount of time it would take to construct a power plant system. A

Table 7. 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)

three-to-five year construction period for a new plant is often assumed. Methods for computing the AFDC are documented elsewhere (e.g., EPRI, 1986; Frey and Rubin, 1990) and are not repeated here. The total plant investment (TPI) represents the sum of the total plant cost and the AFDC.

$$TPI = TPC + AFDC \quad (52)$$

The final measure of capital cost is the total capital requirement (TCR). The TCR includes the total plant investment plus costs for royalties, startup costs, and initial inventories of feedstocks.

The royalty cost is typically estimated as a multiplier of total direct cost:

$$C_{Roy} = f_{Roy} TDC \quad (53)$$

However, in this case, no costs are assumed for royalties ($f_{Roy} = 0$).

Preproduction costs are intended to cover operating, training, equipment checkout, changes in plant equipment, extra maintenance, and inefficient use of fuel, chemicals, catalysts, and other materials during plant startup. Pre-production costs typically include one month of both fixed and variable operating costs and two percent of total plant investment.

$$C_{PP} = \frac{N_{m,FOC} FOC + N_{m,VOC} VOC}{12} + f_{PP} TPI \quad (54)$$

where $N_{m,FOC}$ is the number of months of fixed operating cost to be included in the pre-production cost estimate, $N_{m,VOC}$ is the number of months of variable operating cost to be included, and f_{PP} is the pre-production cost factor, typically assigned a value of 0.02.

The inventory capital cost accounts for inventories of consumables required to be on-site and is typically estimated as 0.5 percent of total process capital excluding catalyst.

$$C_{\text{InvC}} = f_{\text{InvC}} \text{ TDC} \quad (55)$$

For the DSRP, the costs for initial catalysts and chemicals is based upon the initial catalyst requirements for each reactor vessel. The initial catalyst requirement is calculated based upon the molar flow rate of the gas in each reactor stage and the user-specified space velocity for each reactor stage. Because space velocity is expressed on a standard cubic centimeter (scc) basis, the molar flow rate of the gas in each reactor can be converted to a volumetric flow rate. The conversion factor is 10.762 million scc/lbmole. Because available cost data for the DSRP catalyst are on a \$/lb basis, it is necessary to convert from catalyst volume to catalyst weight. This is done by assuming a typical catalyst density of 0.72 g/cc, or 0.00159 lb/cc, based upon data reported by McMichael et al. (1991). Thus, the expression for the total weight of initial catalysts is:

$$m_{\text{cat}} = \left\{ \left(\frac{M_{\text{g,R1,in}}}{SV_{\text{R1,act}}} \right) + \left(\frac{M_{\text{g,R2,in}}}{SV_{\text{R2,act}}} \right) \right\} 10.762 \times 10^6 \cdot 0.00159 \quad (56)$$

The cost for the initial catalyst is:

$$C_{\text{IC}} = \frac{m_{\text{cat}} \text{ UC}_{\text{cat}}}{1,000} \quad (57)$$

The total capital requirement for a system that includes the DSRP is therefore:

$$\text{TCR} = \text{TPI} + C_{\text{PP}} + C_{\text{InvC}} + C_{\text{IC}} + C_{\text{Roy}} \quad (58)$$

5.3 Annual Costs

Fixed operating costs include operating labor, maintenance labor and materials, and overhead costs associated with administrative and support labor. Cost models for these items have previously been developed by Frey and Rubin (1990). The fixed operating costs are developed based upon the direct costs for each process area. A maintenance cost factor is used to estimate the annual maintenance costs as a fraction of direct capital costs for each process area. For the DSRP, a maintenance cost factor of 0.02 is assumed. The annual maintenance cost is divided into materials and labor components. It is assumed that the DSRP does not require any operators in addition to those already present at the power plant site or that would otherwise be needed to operate a conventional sulfuric acid byproduct recovery plant.

The major component of the variable operating cost is for catalyst replacement. Based upon data in Robie et al (1993), it is assumed that the DSRP catalyst is replaced every three years. Therefore, the annual cost for the DSRP catalyst replacement is:

$$OC_{cat} = \left(\frac{m_{cat}}{L_{cat}} \right) UC_{cat} \quad (59)$$

where $L_{cat} = 3$ years.

The performance model of the DSRP is fully integrated with the power plant model with respect to internal auxiliary loads, including electricity and steam. Any incremental water or fuel consumption due to the DSRP is modeled as part of the integrated power plant model simulation.

6.0 MODEL APPLICATIONS

The new performance and cost models of the DSRP have been implemented as part of previously developed performance and cost models of the air-blown Lurgi-gasifier-based IGCC system with hot gas cleanup. This system features a fixed bed zinc ferrite process operating at a pressure of approximately 300 psia.

6.1 Modeling IGCC Systems

The performance, emissions, and cost of the IGCC system are modeled using detailed engineering models. Performance models of a Lurgi-based system were developed by DOE/METC using the ASPEN chemical process simulator. These performance models have been significantly modified to more completely and accurately represent process performance and emissions (Frey, 1991; Frey and Rubin 1991). For example, gas cleanup and gas turbine performance and emissions are more completely characterized in the modified models. A new performance and cost model of selective catalytic reduction (SCR) NO_x control has been added to these IGCC models (Frey, 1994).

The cost models for the IGCC system are based on approximately 30 design studies of IGCC systems (Frey and Rubin, 1990). Direct capital costs are estimated for approximately one dozen major process areas. Typically, several performance and design variables are included in the direct cost models. Indirect and other capital costs are estimated based on approximately 60 cost model parameters. These include process area contingencies, project contingency, indirect construction costs, sales tax, allowance for funds used during construction, environmental permitting costs, spare parts inventory costs, costs for initial inventories of fuels and chemicals, land cost, and startup costs. Fixed and variable operating costs are estimated based on 40 to 50 parameters. Fixed operating costs include maintenance material and labor for each process area, plant operating labor, and administrative and support labor. Variable operating costs include consumables (e.g., water treatment chemicals, zinc ferrite sorbent), ash disposal, fuel, and byproduct credit. Total levelized costs are calculated using the methodology of EPRI (1986).

6.2 Modeling Assumptions

The case studies developed here are based on system performance and cost assumptions which have been documented elsewhere (Frey and Rubin, 1991; 1992; Frey, 1994). Here, we focus on assumptions that are unique to the DSRP process area. These assumptions are given in Table 8. These assumptions include performance and cost inputs

Table 8. Modeling Assumptions for DSRP Systems^a

Parameter	Units	Value
<i>DSRP Performance Assumptions</i>		
Stage I SO ₂ Removal Efficiency	%	95
Stage I SO ₂ Removal Efficiency	%	90
Stage I Space Velocity	scc/cc-hr	5,000
Stage II Space Velocity	scc/cc-hr	5,000
Stage I Reference Space Velocity	scc/cc-hr	2,000
Stage II Reference Space Velocity	scc/cc-hr	2,000
<i>DSRP Cost Assumptions</i>		
Number of Units		2
Process Contingency	% of direct cost	25
Maintenance Cost	% of direct cost	2
DSRP Catalyst Cost	\$/lb	12.28
Catalyst Layer Replacement Interval	Years	3
Sulfuric Acid Sale Price	\$/ton	40
Sulfur Sale Price	\$/ton	125

to the DSRP models. The sulfur removal efficiency assumptions are based upon typical values obtained in small scale testing of the DSRP. The space velocity assumptions are based in part on recently presented results for the DSRP (Portzer, 1995). The reference space velocity numbers are based upon data used in previous design studies (Robin et al., 1993). It is assumed that there are two parallel trains of equipment with no spares. The process contingency and maintenance cost factors have been previously presented. The DSRP catalyst cost is based upon data from Robin et al. (1993), as is the assumption regarding the catalyst life. The sulfuric acid and sulfur price assumptions are from Frey and Rubin (1990).

^a For pressure drop assumptions, please see Table 6.

6.3 Running the Model

The engineering models were exercised to characterize key measures of plant performance, emissions, and cost. The IGCC model was run on a DEC VAXStation 3200 mini-computer using the public version of ASPEN. Model results are reported in Tables 9, 10, and 11. The model assumptions include a current heavy duty gas turbine with a pressure ratio of 15 and a firing temperature of 2,350 °F.

6.4 Modeling Results

Results from application of the new performance and cost model of the DSRP are presented in this section. A total of three cases are considered. One is a base case in which a Lurgi-gasifier-based IGCC system is configured with a conventional sulfuric acid byproduct recovery plant. The second case features a one-stage DSRP system for sulfur

Table 9. Modeling Results for a One-Stage DSRP System: Mass Flow Rates, Temperatures, and Pressures of Key Gas Streams

Description	Mass Flow (lb/hr)	Temp. (deg F)	Pres. (psia)
HGD OFF-GAS	647,678	1234.	305.50
COAL GAS	3,031,798	1206.	286.72
SLIP STREAM	136,108	1206.	286.72
COOLER INLET	783,786	1228.	286.72
SHIFT INLET	783,786	1147.	285.72
REACTOR 1 INLET	783,786	1160.	284.72
CONDENSER 1 INLET	763,091	1292.	259.72
CONDENSER 1 OUTLET	763,091	350.	249.72
DSRP TAILGAS	763,091	350.	249.72
COMPRESSED TAILGAS	763,091	360.	286.72
FUEL TO GAS TURB.	3,662,061	801.	286.72

recovery, in place of the sulfuric acid plant. The third case features a two-stage DSRP. A complete listing of the model results for all three cases is given in Appendix D.

The performance model results for the one and two stage DSRP systems are presented in Tables 9 and 10, respectively. These tables identify the major gas streams in the DSRP, and include data on the calculated mass flows, temperatures, and pressure drops for each stream.

For both systems, a slip stream of coal gas is calculated based upon the requirements for the sum of the molar flow of carbon monoxide and hydrogen to be twice that of the sulfur dioxide in the off-gas from the hot gas desulfurization (HGD) system. The off-gas from the HGD system undergoes a pressure drop from 305.5 to 286.72 psia, which is the pressure level of the coal gas. The mixture of HGD off-gas and the coal gas slip stream enters the DSRP gas cooler at a temperature of approximately 1,228 °F. The gas is cooled to reactor inlet temperature of 1,160 °F, and the maximum gas temperature in the reactor is limited to 1,292 °F. During gas cooling, the water-gas shift reaction is modeled as going to equilibrium. The gas cooler has a pressure drop of approximately 2 psi, while Reactor No. 1 has a pressure drop of approximately 25 psi. For both the one and two stage systems, the gas exiting Reactor No. 1 is cooled to 350 °F, resulting in the condensation of elemental sulfur. In the one-stage system, the outlet gas from the first condenser is also the DSRP tailgas. This gas is recompressed to the pressure of the gas turbine fuel gas, and it is mixed with the fuel gas prior to combustion in the gas turbine.

For the two-stage system, the outlet gas from Condenser No. 1 enters a reheater, where it is increased in temperature to 550 °F. The gas then passes through Reactor No. 2

Table 10. Modeling Results for a Two-Stage DSRP System: Mass Flow Rates, Temperatures, and Pressures of Key Gas Streams

Description	Mass Flow (lb/hr)	Temp. (deg F)	Pres. (psia)
HGD OFF-GAS	637,315	1233.	305.50
COAL GAS	2,982,200	1206.	286.72
SLIP STREAM	133,882	1206.	286.72
COOLER INLET	771,198	1228.	286.72
SHIFT INLET	771,198	1148.	285.72
REACTOR 1 INLET	771,198	1161.	284.72
CONDENSER 1 INLET	752,777	1292.	259.72
CONDENSER 1 OUTLET	752,777	350.	249.72
REHEATER INLET	752,777	350.	249.72
REACTOR 2 INLET	752,777	550.	248.72
CONDENSER 2 INLET	749,874	550.	223.72
CONDENSER 2 OUTLET	749,874	350.	213.72
DSRP TAILGAS	749,874	350.	213.72
COMPRESSED TAILGAS	749,874	372.	286.72
FUEL TO GAS TURB.	3,601,418	852.	286.72

and Condenser No. 2. The tailgas is recompressed to the pressure of the gas turbine fuel gas.

The one-stage DSRP is assumed to have a Reactor No. 1 sulfur capture efficiency of 95 percent. The two-stage system is assumed to have an overall sulfur capture efficiency of approximately 99 percent. As noted earlier, there is uncertainty as to the effectiveness of a second stage. In particular, the reverse Claus reaction may occur in the second stage, which would substantially penalize, rather than enhance, the overall sulfur capture efficiency.

A comparison of results for the base case, one stage DSRP, and two stage DSRP is given in Table 11. The results include power plant performance, cost, byproduct, and emissions.

The DSRP appears to offer significant advantages over a conventional sulfuric acid plant with respect to plant performance. The plant thermal efficiency is approximately six-tenths of a percentage point higher for the two DSRP cases than for the base case system with sulfuric acid recovery. The DSRP system is highly integrated with the IGCC plant, particularly with respect to heat streams and the DSRP tailgas. The DSRP returns a pressurized gas to the gas turbine, enabling recovery of energy through the hot gas path of the gas turbine. The mass flow of fuel to the gas turbine is increased by approximately 20 percent for this reason. In contrast, the tail gas from the sulfuric acid plant is assumed to be vented to the atmosphere. Thus, for the DSRP cases, the ratio of fuel gas to inlet air is slightly greater than 0.4 lb/lb, compared to only 0.26 lb fuel gas per lb of inlet air for the

Table 11. Comparison of Modeling Results for One-Stage and Two-Stage DSRP Systems

Description	Base Case	1-Stage DSRP	2-Stage DSRP
<i>Performance</i>			
Thermal Efficiency, % HHV basis	38.31	38.93	38.96
Net Plant Output, MW	709	851	838
<i>Cost (January 1989 dollars)</i>			
Capital Cost, \$/kW	1,357	1,235	1,277
Fixed Operating Cost, \$/kW-yr	41.41	36.69	37.22
Variable Operating Cost, mills/kWh	19.23	19.15	19.18
Cost of Electricity, mills/kWh	51.14	48.01	48.91
<i>Byproduct</i>			
Sulfuric Acid, tons/yr	173,230		
Sulfur, tons/yr		58,918	60,710
<i>Emissions</i>			
SO ₂ , lb/10 ⁶ BTU	0.041	0.350	0.090

base case. The higher mass flow for the high pressure fuel stream enables greater energy recovery from the gas turbine.

The steam turbine energy recovery is higher for the DSRP cases due to a combination of factors. The gas turbine exhaust temperature is slightly higher for the DSRP cases, thereby enabling more energy recovery through the heat recovery steam generator. In addition, the integration of the DSRP with the steam cycle enables efficient use of steam, especially the steam generated due to gas cooling and condensation in the DSRP.

Thus, the net power produced by the IGCC systems with the DSRP is substantially higher than for the base case with sulfuric acid recovery.

The cost of the IGCC systems with the DSRP is lower than for the base case. The capital cost of the two-stage DSRP is comparable to that for the sulfuric acid plant (approximately \$25 million). However, the increased power available from the gas turbines, due to an increase in the flow of pressurized fuel gas, leads to a reduced cost on a per kilowatt basis. This is even though the IGCC system with the two-stage DSRP has a larger coal throughput and, hence, higher absolute costs for coal handling, oxidant feed, gasification and associated subsystems, hot gas desulfurization, heat recovery steam generator, and steam turbine.

The comparison of the capital costs of the one-stage DSRP to the base case is similar to the comparison of the two-stage DSRP to the base case, except that the direct

capital cost for the one-stage DSRP is approximately 35 percent less than for the two-stage design. Therefore, there is an additional cost savings for the one-stage design when compared to the two-stage design.

The cost comparisons are similar for fixed operating, variable operating, and levelized costs. The one-stage design offers the lowest costs for all of these categories.

The byproduct mass flow rates are significantly different for the base case versus the DSRP cases. The key shortcomings of sulfuric acid production are the large mass of acid that is produced, and the difficulty in storing and handling the acid. In contrast, elemental sulfur is easier to handle, store, and ship, and is produced in smaller mass quantities for the same amount of recovered sulfur.

A potential advantage of the sulfuric acid recovery option is the high sulfur capture efficiency and correspondingly low plant sulfur dioxide emission rate. The two-stage DSRP design has an estimated sulfur dioxide emission rate twice that of the sulfuric acid recovery option. This presumes that the second stage of the DSRP would be effective in removing sulfur, rather than in producing sulfur gases through the reverse Claus reaction. The one-stage DSRP is shown to have the highest SO_2 emission rate of the three options.

7.0 SUMMARY

Detailed performance and cost models of DSRP systems have been developed. These models have been implemented as computer code to augment previously developed performance, emissions, and cost models of selected IGCC systems.

The DSRP models allow the implications of alternative sulfur control strategies to be explored, and also permit sensitivity and optimization case studies to minimize the total cost of environmental control for advanced power generation systems. The illustrative case studies presented here indicate that there may be substantial performance and cost benefits to the use of the DSRP in lieu of more conventional approaches for sulfur byproduct recovery.

Because DSRP systems have not yet been applied to many of the advanced power generation systems, it is critically important to consider the role of uncertainties in making predictions of DSRP performance and cost. Because many of these technologies are of interest to research and development (R&D) planners and to electric utilities, it is necessary to fully understand their environmental characteristics and the costs of complying with current and future regulations. Thus, the use of probabilistic modeling techniques, as described by Frey and Rubin (1991), to evaluate the DSRP will be a key area for future work.

The DSRP performance and cost models developed here will need to be revised as new data and technologies become available. As experience is gained regarding long-term catalyst durability, model input assumptions may be revised. As better understanding is gained regarding the factors that affect catalyst sizing, perhaps more detailed mechanistic models can be developed to characterize catalyst requirements. Therefore, the models developed here should be viewed as a first step in an iterative model development and adaptation process, as new data become available and as the need emerges to adapt these models for applications to new types of systems not currently envisioned.

8.0 REFERENCES

- Buchanan, T.L., H.T. Chen, M.G. Klett, M.D. Rutkowski, and R. Zaharchuk (1994). "Optimization of Gas Stream Cleanup in Three IGCC Systems", Prepared by Gilbert/Commonwealth for U.S. Department of Energy, Contract DE-AC01-88FE61660, March 1994.
- BRSC (1987). *Flue Gas Cleanup Technologies By-Products*, Final Report. Contract No. DE-AC22-84PC72571. Prepared by Burns and Roe Services Corporation for U.S. Department of Energy, Pittsburgh, PA. September.
- EPRI (1986). TAG(tm) - Technical Assessment Guide, Volume 1: Electricity Supply - 1986. EPRI P-4463-SR. Electric Power Research Institute, Inc. December.
- Fluor (1985). Cost and Performance of Kellogg Rust Westinghouse-based Gasification-Combined-Cycle Plants. Prepared by Fluor Engineers, Inc. for Electric Power Research Institute, Palo Alto, CA. EPRI AP-4018. June.
- Frey, H.C. (1991), *Probabilistic Modeling of Innovative Clean Coal Technologies: Implications for Research Planning and Technology Evaluation*, Ph.D. Thesis, Department of Engineering and Public Policy, Carnegie Mellon University, Pittsburgh, Pennsylvania, May 1991.
- Frey, H.C. (1994). "Development and Application of Performance and Cost Models for Gas Turbine-Based Selective Catalytic Reduction NO_x Control", Contract DE-AC21-92MC29094. Prepared by Carnegie Mellon University for U.S. Department of Energy, Morgantown, WV, October.
- Frey, H.C. and E.S. Rubin (1990), *Stochastic Modeling of Coal Gasification Combined Cycle Systems: Cost Models of Selected Integrated Gasification Combined Cycle (IGCC) Systems*, Topical Report, prepared by Carnegie-Mellon University for the U.S. Department of Energy, Morgantown, West Virginia, DOE/MC/24248-2901, NTIS DE90015345, June 1990, 307p.
- Frey, H.C., and E.S. Rubin (1991), *Development and Application of a Probabilistic Evaluation Method for Advanced Process Technologies*, Final Report, prepared by Carnegie-Mellon University for the U.S. Department of Energy, Morgantown, West Virginia, DOE/MC/24248-3015, NTIS DE91002095, April 1991, 364p.
- Frey, H.C., and Rubin, E.S. (1992). "Evaluation of advanced coal gasification combined-cycle systems under uncertainty." *Ind. Eng. Chem. Res.*, 31:1299-1307.
- Gangwal, S.K., and W.J. McMichael. 1990. "Bench-Scale Testing of Sulfur Recovery Process", In *Proceedings of the Tenth Annual Gasification and Gas Stream Cleanup Systems Contractors Review Meeting*, DOE/METC-90/6115, Vol. 1. (DE90009689), August.
- Gangwal, S.K., and W.J. McMichael. 1991. "Bench-Scale Testing of Sulfur Recovery Process", In *Proceedings of the Eleventh Annual Gasification and Gas Stream Cleanup Systems Contractors Review Meeting*, DOE/METC-91/6123, Vol. 2. (DE92001102), August.
- Gangwal, S.K., and W.J. McMichael. 1992. "Bench-Scale Testing of Sulfur Recovery Process", In *Proceedings of the Twelfth Annual Gasification and Gas Stream Cleanup*

Systems Contractors Review Meeting, DOE/METC-92/6128, Vol. 1. (DE93000228), September.

Gangwal, S.K., et al. 1993. "DSRP, Direct Sulfur Production", In *Proceedings of Coal-Fired Power Systems 93--Advances in IGCC and PFBC Review Meeting*, Contracted by U.S. Department of Energy, Morgantown, WV, June 1993.

Gangwal, S.K., et al. (1994). "Slipstream Testing of the Direct Sulfur Recovery Process," In *Proceedings of Coal-Fired Power Systems 94--Advances in IGCC and PFBC Review Meeting*, U.S. Department of Energy, Morgantown, WV, June 1994.

Kasper, S. 1988. "Design Basis for Zinc Ferrite Hot Gas Desulfurization Systems", In *Proceedings of the Eighth Annual Gasification and Gas Stream Cleanup Systems Contractors Review Meeting*, Vol.1. V.P. Kothari and J.R. Longanbach, Eds. U.S. Dept. of Energy Morgantown Energy Technology Center, Morgantown, WV. May.

McMichael, W.J., and S.K. Gangwal. 1991. "Recovery of Sulfur from Hot Gas Desulfurization Processes". Final Report. DOE/METC Contract No. DE-AC21-86MC23260, April.

METC (1987). "Gas Stream Cleanup Technology Status Report". DOE/METC-87/0255. NTIS/DE87006493. U.S. Department of Energy, Morgantown Energy Technology Center, Morgantown, WV. October.

Notestein, John E. 1990. "Commercial Gasifier for IGCC Applications Study Report". DOE/METC-91/6118. (DE91002051), June.

Portzer, J.W. (1995). "Slip-Stream Testing of Hot-Gas Desulfurization with Sulfur Recovery". Presented at Advanced Coal-Fired Power Systems '95 Review Meeting, U.S. Department of Energy, Morgantown, WV, June 27-29.

Robin, A.M., L.A. Davis, T.F. Leininger, et al. (1993). "Integration and Testing of Hot Desulfurization and Entrained-Flow Gasification for Power Generation Systems", Volume 1, Prepared by Texaco, Inc. for the U.S. Department of Energy, Morgantown, WV, May 1993.

SFA Pacific (1987). Process Screening Study of Alternative Gas Treating and Sulfur Removal Systems for IGCC Power Plant Applications. Prepared by SFA Pacific, Inc. for Electric Power Research Institute. Palo Alto, CA. EPRI AP-5505. December.

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

Ulrich, G.D. (1984). *A Guide to Chemical Engineering Process Design and Economics*. John Wiley and Sons. New York. 1984.

Underkoffler, V.S. 1986. "Summary and Assessment of METC Zinc Ferrite Hot Coal Gas Desulfurization Test Program", Final Report. Volume I and Volume II, Appendices. Prepared by Gilbert/Commonwealth for the U.S. DOE-METC, Morgantown, WV. DOE/MC/21098-2247-Vol.1 and -Vol.2. December.

Appendix A

Regression Plots for Stage I Using Catalyst A for Low Space Velocities

Regression Plots for Stage I Using Catalyst A for Low Space Velocities

Regression Plots for Stage I Using Catalyst A for High Space Velocities

Regression Plots for Stage I Using Catalyst A for High Space Velocities

Regression Plots for Stage II Using Catalyst B for Low Space Velocities

Regression Plots for Stage II Using Catalyst B for Low Space Velocities

Regression Plots for Stage II Using Catalyst C for Low Space Velocities

Regression Plots for Stage II Using Catalyst C for Low Space Velocities

Appendix B

DSRP PERFORMANCE MODEL CODE

```

;* IN-UNITS are assumed to be ENG
;*
; //////////////////////////////////////
; //////////////////////////////////////
; ///          DIRECT SULFUR RECOVERY PROCESS          ///
; //////////////////////////////////////
; //////////////////////////////////////
;
; Date:      June 15, 1995
; Author:    H.C. Frey, NC State Univ.
; Developed originally for air-blown Lurgi flowsheet.
; Key Input Variables:
;   FUELGAZ is the clean fuel gas exiting the high temperature
;           desulfurization system.
;   SO2GAS  is the regenerator off-gas.  If the inlet gas variable is
;           different, be sure to search for it and change it
;           (e.g., in SET-RSPLT)
; Key Output Variables
;   FUELGAZ2 is what goes to the gas turbine
;
; Integration of heat streams:
;   High temperature streams ( approx 1200 deg F outlet temps)
;     QDSRP-CL QDSRP-SF QDSRP-R1 QDRSP-C1
;     These can be integrated with the superheater (SUPERHTR)
;     These are combined into one stream:  QDSRP-H
;   Medium temperature streams (approx 500 deg outlet temps)
;     QDSRP-RH QDSRP-R2 QDRSP-C2
;     These can be integrated with the low pressure boiler
; (LPBOILER)
;     These are combined into one stream:  QDSRP-M
;   Low temperature streams (approx 300 deg outlet temps)
;     QDSRP-SL
;     This can be integrated with LPECON (note, while the inlet
;     temp of LPECON is less than 300 deg, the outlet is not.
;     Therefore, this is an approximation.
;     These are combined into one stream:  QDSRP-L
;
; Integration of work streams
;   All work streams are mixed to form WDSRP.  The work from WDSRP
;   should be included in the cost model subroutine in determining
;   the plant efficiency.
;
; Integration of pressure specifications
;   Pressure drops are specified throughout the DSRP model.
;   The tail gas which returns to the fuel gas stream must be
;   compressed back up to the fuel gas pressure.  This is done
;   in TAILCOMP
;
;
; *****
;*                               DSRP STREAM DATA                               *
; *****

```

```

;
DEF-STREAMS &
    MIXCISLD DSRP      /
    CONVEN    FUELGAZ2  TAILCNV HPTCNV /
    HEAT      QDSRP-CL QDSRP-SF QDSRP-R1 QDRSP-C1 QDSRP-RH &
                                QDSRP-R2 QDRSP-C2 QDSRP-SL &
                                QDSRP-H   QDSRP-M   QDSRP-L   /
    WORK      TAILWK WDSRP
;
;*****
;*                      DSRP FLOWSHEET                      *
;*****
;
FLOWSHEET DSRP
    DSRP-CLC IN = FUELGAZ          OUT = TODSRP
    DSRP-CL2 IN = SO2GAS           OUT = SO2PROD
    REDUSPLT IN = TODSRP           OUT = DSLIP      TOGTMIXR
    DSRP-MIX IN = DSLIP            SO2PROD      OUT = DCOOLIN
    DSRP-COL IN = DCOOLIN          OUT = DSHFIN    QDSRP-CL
    DSRP-SHF IN = DSHFIN           OUT = DR1IN    D-DUMMY  QDSRP-SF
    DSRP-R1  IN = DR1IN            OUT = DC1IN    QDSRP-R1
    DSRP-C1  IN = DC1IN            OUT = DC1OUT   QDRSP-C1
    DSRP-SP1 IN = DC1OUT           OUT = DRHIN    DES1
    DSRP-RHX IN = DRHIN            OUT = DR2IN    QDSRP-RH
    DSRP-R2  IN = DR2IN            OUT = DC2IN    QDSRP-R2
    DSRP-C2  IN = DC2IN            OUT = DC2OUT   QDRSP-C2
    DSRP-SP2 IN = DC2OUT           OUT = TAILGAS   DES2
    DSRP-CL4 IN = TAILGAS          OUT = TAILCNV
    TAILCOMP IN = TAILCNV          OUT = HPTCNV    TAILWK
    DSRP-CL5 IN = HPTCNV           OUT = HPTAIL
    GTMIXR   IN = TOGTMIXR HPTAIL  OUT = FUELTEMP
    DSRP-CL3 IN = FUELTEMP          OUT = FUELGAZ2
    DSRP-SM   IN = DES1            DES2      OUT = DSPIN
    DSRP-PRD  IN = DSPIN           OUT = SULPROD  QDSRP-SL
    DSRP-WK   IN = TAILWK          OUT = WDSRP
    DSRP-HTM IN = QDSRP-CL QDSRP-SF &
                                QDSRP-R1 QDRSP-C1 OUT = QDSRP-H
    DSRP-MTM IN = QDSRP-RH QDSRP-R2 QDRSP-C2 OUT = QDSRP-M
    DSRP-LTM IN = QDSRP-SL          OUT = QDSRP-L
;
;*****
;*                      DSRP BLOCKS                          *
;*****
;
BLOCK DSRP-CLC CLCHNG
;  changes the class of the fuel gas from conventional to mixcisl
;
BLOCK DSRP-CL2 CLCHNG
;  changes the class of the SO2GAS from conven to mixcisl
;
BLOCK REDUSPLT FSPLIT ; ACTUAL MOLE FLOW SET BY FORTRAN BLOCK SET-RSPLT
    DESCRIPTION ' SPLIT SLIP STREAM OF REDUCING GAS '
    MOLE-FLOW DSLIP 50.5 1
    DEF-KEY ID=1 SUBS=MIXED COMP=CO H2
;
BLOCK DSRP-MIX MIXER
    DESCRIPTION ' CONTACT REDUCING GAS AND REGENERATOR OFF-GAS '

```

```

;
BLOCK DSRP-COL HEATER
  DESCRIPTION ' DSRP INLET GAS PRECOOLER '
  PARAM PRES=-1 TEMP=800.
;   Cooler outlet temperature is set by Design Spec DS-DSRPC
;
BLOCK DSRP-SHF REQUIL
  DESCRIPTION " DSRP WATER-GAS SHIFT EQUILIBRIUM "
  BLOCK-OPTIONS RESTART=0
  PARAM PRES=-1 NR=1 NPK=1 MXOL=90 MXIL=90 KODE=1 TEST=800.
  STOI   1   CO -1 0 / H2O -1 0 / H2 1 0 / CO2 1 0
;
BLOCK DSRP-R1 RSTOIC
  DESCRIPTION ' DSRP REACTOR NO. 1 '
;
; RATE OF CONVERSION CONTROLLED BY FORTRAN BLOCK SET-DCNV
;
  PARAM PRES = -25. TEMP = 1292.
  STOI 1 MIXED SO2 -1 / COS -2 / CO2 2 / CISOLID SULFUR 3 /
        2 MIXED SO2 -1 / H2 -2 / H2O 2 / CISOLID SULFUR 1 /
        3 MIXED SO2 -1 / H2 -3 / H2O 2 / H2S 1 /
        4 MIXED SO2 -1 / CO -3 / CO2 2 / COS 1 /
        5 MIXED SO2 -1 / CO -2 / CO2 2 / CISOLID SULFUR 1
  CONV 1 MIXED SO2 0. /
        2 MIXED SO2 0. /
        3 MIXED SO2 0. /
        4 MIXED SO2 0. /
        5 MIXED SO2 0.
;
BLOCK DSRP-C1 HEATER ; HEAT STREAM SENT TO HPECON
  DESCRIPTION ' ELEMENTAL SULFUR CONDENSER #1 '
  PARAM PRES=-10. TEMP=350.
;
BLOCK DSRP-SP1 SSPLIT
  DESCRIPTION ' SPLIT OUT THE CONDENSED ELEMENTAL SULFUR '
  FRAC CISOLID DES1 1.0 / MIXED DRHIN 1.0
;
BLOCK DSRP-RHX HEATER
  DESCRIPTION "DSRP REHEATER"
  PARAM PRES=-1. TEMP=550.
;
BLOCK DSRP-R2 RSTOIC
  DESCRIPTION ' DSRP REACTOR NO. 2 '
;
; RATE OF CONVERSION CONTROLLED BY FORTRAN BLOCK SET-DCNV
;
  PARAM PRES = -25. TEMP = 550.
  STOI 1 MIXED COS -1 / H2O -1 / CO2 1 / H2S 1 /
        2 MIXED SO2 -1 / H2S -2 / H2O 2 / CISOLID SULFUR 3
  CONV 1 MIXED COS 1.0 /
        2 MIXED SO2 0.
;
BLOCK DSRP-C2 HEATER ; HEAT STREAM SENT TO HPECON
  DESCRIPTION ' ELEMENTAL SULFUR CONDENSER #2 '
  PARAM PRES=-10. TEMP=350.
;
BLOCK DSRP-SP2 SSPLIT
  DESCRIPTION ' SPLIT OUT THE CONDENSED ELEMENTAL SULFUR '

```



```

      FRAC CISOLID DES2 1.0 / MIXED TAILGAS 1.0
;
BLOCK DSRP-CL4 CLCHNG
;
BLOCK TAILCOMP COMPR
  DESCRIPTION ' DSRP TAIL GAS COMPRESSOR '
  PARAM TYPE=1 PRES=286.72 EP=0.9 GFLAG=0
;
BLOCK DSRP-CL5 CLCHNG
;
BLOCK GTMIXR MIXER ; MODELS LINE LOSSES OF GAS STREAM TO GT
  DESCRIPTION ' MIX DSRP GAS EFFLUENT AND MAIN STREAM FOR GT FEED '
  PARAM PRES=286.72
;
BLOCK DSRP-CL3 CLCHNG
; Changes FUELTEMP (MIXCISLD) to FUELGAZ2 (CONVEN), which goes to GT
;
BLOCK DSRP-SM MIXER
  DESCRIPTION "MIX SULFUR FROM CONDENSERS 1 AND 2"
;
BLOCK DSRP-PRD HEATER
  PARAM PRES=14.7 TEMP=300.
;
BLOCK DSRP-WK MIXER
;
; Combine similar temperature heat streams for integration with
; the steam cycle
;
BLOCK DSRP-HTM MIXER
;
BLOCK DSRP-MTM MIXER
;
BLOCK DSRP-LTM MIXER
;
; *****
;                               DSRP SEQUENCING
; *****
;
CONVERGENCE DS-COOL ONE-VAR
  SPEC DS-DSRPC
;
SEQUENCE DSRP-SEQ &
  DSRP-CLC DSRP-CL2 SET-RSPLT REDUSPLT &
  DSRP-MIX DS-COOL DSRP-COL &
  DSRP-SHF SET-DCNV &
  DSRP-R1 (RETURN DS-COOL) DSRP-C1 DSRP-SP1 &
  DSRP-RHX DSRP-R2 DSRP-C2 DSRP-SP2 &
  DSRP-CL4 TAILCOMP DSRP-CL5 GTMIXR DSRP-CL3 &
  DSRP-SM DSRP-PRD DSRP-WK &
  DSRP-HTM DSRP-MTM DSRP-LTM &
  DSRP-SUM
;
; *****
;                               DSRP FORTRAN BLOCKS
; *****
;
FORTRAN SET-RSPLT ; FLOWRATE SET AT 0% ABOVE STOICHIOMETRIC
; DESCRIPTION ' SET REDUCING GAS FLOWRATE TO DSRP PLANT '

```

```

      DEFINE SFLOW MOLE-FLOW STREAM=SO2PROD SUBSTREAM=MIXED COMPONENT=SO2
      DEFINE RFRAC BLOCK-VAR BLOCK=REDUSPLT SENTENCE=MOLE-FLOW      &
                                         VARIABLE=FLOW ID1=DSLIP
F      RFRAC=2.00*SFLOW
;      EXECUTE BEFORE REDUSPLT
;
      FORTRAN SET-DCNV
F      COMMON /USRDSR/ DSRSV1, DSRSV2,
F      &                FDSPRE, VGR1IN, DCDSG1, DCDSS1,
F      &                FDSPR2, VGR2IN, DCDSG2, DCDSS2,
F      &                DCDS,   TDSCAT, ADSCAT, XMSULF
F      &                EFSO21, EFSO22, DSRPR
F      COMMON /USRDSI/ NSTGDS
      DESCRIPTION "SET CONVERSION RATES FOR REACTORS 1 AND 2"
      DEFINE XSO2I MOLE-FLOW STREAM=DR1IN SUBSTREAM=MIXED COMPONENT=SO2
      DEFINE XH2SI MOLE-FLOW STREAM=DR1IN SUBSTREAM=MIXED COMPONENT=H2S
      DEFINE XCOI  MOLE-FLOW STREAM=DR1IN SUBSTREAM=MIXED COMPONENT=CO
      DEFINE XCOSI MOLE-FLOW STREAM=DR1IN SUBSTREAM=MIXED COMPONENT=COS
      DEFINE XTOTI SUBSTREAM-VAR STREAM=DR1IN  SUBS=MIXED VAR=MOLEFLOW
;
      DEFINE CNV11 BLOCK-VAR BLOCK=DSRP-R1 SENTENCE=CONV VAR=CONV ID1=1
      DEFINE CNV12 BLOCK-VAR BLOCK=DSRP-R1 SENTENCE=CONV VAR=CONV ID1=2
      DEFINE CNV13 BLOCK-VAR BLOCK=DSRP-R1 SENTENCE=CONV VAR=CONV ID1=3
      DEFINE CNV14 BLOCK-VAR BLOCK=DSRP-R1 SENTENCE=CONV VAR=CONV ID1=4
      DEFINE CNV15 BLOCK-VAR BLOCK=DSRP-R1 SENTENCE=CONV VAR=CONV ID1=5
      DEFINE CNV21 BLOCK-VAR BLOCK=DSRP-R2 SENTENCE=CONV VAR=CONV ID1=1
      DEFINE CNV22 BLOCK-VAR BLOCK=DSRP-R2 SENTENCE=CONV VAR=CONV ID1=2
;
      DEFINE YSO2I MOLE-FLOW STREAM=TODSRP SUBSTREAM=MIXED COMPONENT=SO2
      DEFINE YH2SI MOLE-FLOW STREAM=TODSRP SUBSTREAM=MIXED COMPONENT=H2S
      DEFINE YCOI  MOLE-FLOW STREAM=TODSRP SUBSTREAM=MIXED COMPONENT=CO
      DEFINE YCOSI MOLE-FLOW STREAM=TODSRP SUBSTREAM=MIXED COMPONENT=COS
      DEFINE YTOTI SUBSTREAM-VAR STREAM=TODSRP  SUBS=MIXED VAR=MOLEFLOW
;
      DEFINE ZSO2I MOLE-FLOW STREAM=DSHFIN SUBSTREAM=MIXED COMPONENT=SO2
      DEFINE ZH2SI MOLE-FLOW STREAM=DSHFIN SUBSTREAM=MIXED COMPONENT=H2S
      DEFINE ZCOI  MOLE-FLOW STREAM=DSHFIN SUBSTREAM=MIXED COMPONENT=CO
      DEFINE ZCOSI MOLE-FLOW STREAM=DSHFIN SUBSTREAM=MIXED COMPONENT=COS
      DEFINE ZTOTI SUBSTREAM-VAR STREAM=DSHFIN  SUBS=MIXED VAR=MOLEFLOW
;
      DEFINE TDSRC1 BLOCK-VAR BLOCK=DSRP-C1 VAR=TEMP
      DEFINE PDSRC1 BLOCK-VAR BLOCK=DSRP-C1 VAR=PRES
      DEFINE TDSREH BLOCK-VAR BLOCK=DSRP-RHX VAR=TEMP
      DEFINE PDSREH BLOCK-VAR BLOCK=DSRP-RHX VAR=PRES
      DEFINE TDSRR2 BLOCK-VAR BLOCK=DSRP-R2 VAR=TEMP
      DEFINE PDSRR2 BLOCK-VAR BLOCK=DSRP-R2 VAR=PRES
      DEFINE TDSRC2 BLOCK-VAR BLOCK=DSRP-C2 VAR=TEMP
      DEFINE PDSRC2 BLOCK-VAR BLOCK=DSRP-C2 VAR=PRES
;
;      INITIALIZE KEY DSRP VARIABLES HERE FOR THE TIME BEING
;
;      EFSO21 = SO2 Removal Efficiency for Reactor 1
;      EFSO22 = SO2 Removal Efficiency for Reactor 2
;      DSRPR  = CO2/COS molar ratio
;
F      EFSO21 = 0.95
F      EFSO22 = 0.90

```

```

F      DSRPR      = 2.
F      NDSSTG = 2
F      IF(NDSSTG.EQ.1) THEN
F          EFSO22 = 0.0
F          TDSREH = TDSRC1
F          PDSREH = PDSRC1
F          TDSRR2 = TDSRC1
F          PDSRR2 = PDSRC1
F          TDSRC2 = TDSRC1
F          PDSRC2 = PDSRC1
F      ENDIF
;
;      echo input variables to the history file for debugging purposes
;
F      WRITE(NHSTRY,201) YSO2I,YH2SI,YCOI,YCOSI,YTOTI
F 201 FORMAT(/' Sampled Variables for stream TODSRP ',/,
F      &      '      SO2 Flow Rate ',F10.2,/,
F      &      '      H2S Flow Rate ',F10.2,/,
F      &      '      CO Flow Rate ',F10.2,/,
F      &      '      COS Flow Rate ',F10.2,/,
F      &      '      Total Flow Rate ',F10.2,/)
;
F      WRITE(NHSTRY,202) ZSO2I,ZH2SI,ZCOI,ZCOSI,ZTOTI
F 202 FORMAT(/' Sampled Variables for stream DSHFIN ',/,
F      &      '      SO2 Flow Rate ',F10.2,/,
F      &      '      H2S Flow Rate ',F10.2,/,
F      &      '      CO Flow Rate ',F10.2,/,
F      &      '      COS Flow Rate ',F10.2,/,
F      &      '      Total Flow Rate ',F10.2,/)
F      WRITE(NHSTRY,203) XSO2I,XH2SI,XCOI,XCOSI,XTOTI,
F      &      CNV11,CNV12,CNV13,CNV14,CNV15,CNV21,CNV22
F 203 FORMAT(/' Sampled Variables for stream DR1IN ',/,
F      &      '      SO2 Flow Rate ',F10.2,/,
F      &      '      H2S Flow Rate ',F10.2,/,
F      &      '      CO Flow Rate ',F10.2,/,
F      &      '      COS Flow Rate ',F10.2,/,
F      &      '      Total Flow Rate ',F10.2,/,
F      &      '      Reaction 11 cnv ',F10.4,/,
F      &      '      Reaction 12 cnv ',F10.4,/,
F      &      '      Reaction 13 cnv ',F10.4,/,
F      &      '      Reaction 14 cnv ',F10.4,/,
F      &      '      Reaction 15 cnv ',F10.4,/,
F      &      '      Reaction 21 cnv ',F10.4,/,
F      &      '      Reaction 22 cnv ',F10.4,/)
;
F      ETA1 = 0.
F      ETA2 = 0.
F      ETA3 = 0.
F      ETA4 = 0.
F      IF(XSO2I.LE.0) GOTO 101
F      ETA1 = 0.5*XCOSI/XSO2I
F      IF(ETA1.LT.0.) ETA1=0
F      IF(ETA1.GT.1.) ETA1=1.
F      ETA2 = (2*(1-EFSO21)*XSO2I-XH2SI-1/(DSRPR+1)*XCOI)/XSO2I
F      IF(ETA2.LT.0.) ETA2=0.
F      IF(ETA2.GT.1.) ETA2=1.
F      ETA3 = (DSRPR/(2*(DSRPR+1))*XCOI)/XSO2I
F      IF(ETA3.LT.0.) ETA3=0

```

```

F      IF(ETA3.GT.1.) ETA3=1.
F      ETA4 = EFSO21 - (ETA1+ETA2+ETA3)
F      IF(ETA4.LT.0.) ETA4=0
F      IF(ETA4.GT.1.) ETA4=1.
;
F      WRITE(NHSTRY,204) ETA1,ETA2,ETA3,ETA4
F 204 FORMAT(/' Reaction Conversion Efficiencies ',/,
F      &          '          ETA1 ',F10.4,/,
F      &          '          ETA2 ',F10.4,/,
F      &          '          ETA3 ',F10.4,/,
F      &          '          ETA4 ',F10.4,/)
;
F      CNV11 = ETA1
F      CNV12 = ETA4
F      CNV13 = ETA2
F      CNV14 = ETA3*.5
F      CNV15 = ETA3*.5
;
F      CNV21 = 1.
F      CNV22 = EFSO22
F      IF(NSTGDS.EQ.1) THEN
F          CNV21 = 0.
F          CNV22 = 0.
F      ENDIF
F 101 CONTINUE
;
FORTRAN DSRP-SUM
F      CHARACTER TIM*8
F      CHARACTER*3 MONTH(12)
F      DATA MONTH(1)/'JAN'/,MONTH(2)/'FEB'/,MONTH(3)/'MAR'/,
F      + MONTH(4)/'APR'/,MONTH(5)/'MAY'/,MONTH(6)/'JUN'/,
F      + MONTH(7)/'JUL'/,MONTH(8)/'AUG'/,MONTH(9)/'SEP'/,
F      + MONTH(10)/'OCT'/,MONTH(11)/'NOV'/,MONTH(12)/'DEC'/
F      COMMON/USRDSRP/ EFSO21, EFSO22, DSRPR
F      DESCRIPTION ' DSRP SUMMARY BLOCK '
F      IN-UNITS ENG
F      DEFINE XSO2GA SUBSTREAM-VAR STREAM=SO2PROD SUBS=MIXED VAR=MASSFLOW
F      DEFINE XZFCLF SUBSTREAM-VAR STREAM=ZF-CLFL SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XGTM SUBSTREAM-VAR STREAM=TOGTMIXR SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XSLIP SUBSTREAM-VAR STREAM=DSLIP SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XCOOLI SUBSTREAM-VAR STREAM=DCOOLIN SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XSHFIN SUBSTREAM-VAR STREAM=DSHFIN SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XR1IN SUBSTREAM-VAR STREAM=DR1IN SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XC1IN SUBSTREAM-VAR STREAM=DC1IN SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XC1OUT SUBSTREAM-VAR STREAM=DC1OUT SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XRHIN SUBSTREAM-VAR STREAM=DRHIN SUBS=MIXED
F      VAR=MASSFLOW
F      DEFINE XR2IN SUBSTREAM-VAR STREAM=DR2IN SUBS=MIXED
F      VAR=MASSFLOW

```

```

      DEFINE XC2IN      SUBSTREAM-VAR  STREAM=DC2IN      SUBS=MIXED
      VAR=MASSFLOW
      DEFINE XC2OUT     SUBSTREAM-VAR  STREAM=DC2OUT     SUBS=MIXED
      VAR=MASSFLOW
      DEFINE XTLGAS     SUBSTREAM-VAR  STREAM=TAILGAS    SUBS=MIXED
      VAR=MASSFLOW
      DEFINE XHPTL      SUBSTREAM-VAR  STREAM=HPTAIL     SUBS=MIXED
      VAR=MASSFLOW
      DEFINE XFUELT     SUBSTREAM-VAR  STREAM=FUELTEMP   SUBS=MIXED
      VAR=MASSFLOW
;
      DEFINE TSO2GA     SUBSTREAM-VAR  STREAM=SO2PROD    SUBS=MIXED  VAR=TEMP
      DEFINE TZFCLF     SUBSTREAM-VAR  STREAM=ZF-CLFL    SUBS=MIXED  VAR=TEMP
      DEFINE TGTM       SUBSTREAM-VAR  STREAM=TOGTMIXR   SUBS=MIXED  VAR=TEMP
      DEFINE TSLIP      SUBSTREAM-VAR  STREAM=DSLIP      SUBS=MIXED  VAR=TEMP
      DEFINE TCOOLI     SUBSTREAM-VAR  STREAM=DCOOLIN     SUBS=MIXED  VAR=TEMP
      DEFINE TSHFIN     SUBSTREAM-VAR  STREAM=DSHFIN     SUBS=MIXED  VAR=TEMP
      DEFINE TR1IN      SUBSTREAM-VAR  STREAM=DR1IN      SUBS=MIXED  VAR=TEMP
      DEFINE TC1IN      SUBSTREAM-VAR  STREAM=DC1IN      SUBS=MIXED  VAR=TEMP
      DEFINE TC1OUT     SUBSTREAM-VAR  STREAM=DC1OUT     SUBS=MIXED  VAR=TEMP
      DEFINE TRHIN      SUBSTREAM-VAR  STREAM=DRHIN      SUBS=MIXED  VAR=TEMP
      DEFINE TR2IN      SUBSTREAM-VAR  STREAM=DR2IN      SUBS=MIXED  VAR=TEMP
      DEFINE TC2IN      SUBSTREAM-VAR  STREAM=DC2IN      SUBS=MIXED  VAR=TEMP
      DEFINE TC2OUT     SUBSTREAM-VAR  STREAM=DC2OUT     SUBS=MIXED  VAR=TEMP
      DEFINE TTLGAS     SUBSTREAM-VAR  STREAM=TAILGAS    SUBS=MIXED  VAR=TEMP
      DEFINE THPTL      SUBSTREAM-VAR  STREAM=HPTAIL     SUBS=MIXED  VAR=TEMP
      DEFINE TFUELT     SUBSTREAM-VAR  STREAM=FUELTEMP   SUBS=MIXED  VAR=TEMP
;
      DEFINE PSO2GA     SUBSTREAM-VAR  STREAM=SO2PROD    SUBS=MIXED  VAR=PRES
      DEFINE PZFCLF     SUBSTREAM-VAR  STREAM=ZF-CLFL    SUBS=MIXED  VAR=PRES
      DEFINE PGTM       SUBSTREAM-VAR  STREAM=TOGTMIXR   SUBS=MIXED  VAR=PRES
      DEFINE PSLIP      SUBSTREAM-VAR  STREAM=DSLIP      SUBS=MIXED  VAR=PRES
      DEFINE PCOOLI     SUBSTREAM-VAR  STREAM=DCOOLIN     SUBS=MIXED  VAR=PRES
      DEFINE PSHFIN     SUBSTREAM-VAR  STREAM=DSHFIN     SUBS=MIXED  VAR=PRES
      DEFINE PR1IN      SUBSTREAM-VAR  STREAM=DR1IN      SUBS=MIXED  VAR=PRES
      DEFINE PC1IN      SUBSTREAM-VAR  STREAM=DC1IN      SUBS=MIXED  VAR=PRES
      DEFINE PC1OUT     SUBSTREAM-VAR  STREAM=DC1OUT     SUBS=MIXED  VAR=PRES
      DEFINE PRHIN      SUBSTREAM-VAR  STREAM=DRHIN      SUBS=MIXED  VAR=PRES
      DEFINE PR2IN      SUBSTREAM-VAR  STREAM=DR2IN      SUBS=MIXED  VAR=PRES
      DEFINE PC2IN      SUBSTREAM-VAR  STREAM=DC2IN      SUBS=MIXED  VAR=PRES
      DEFINE PC2OUT     SUBSTREAM-VAR  STREAM=DC2OUT     SUBS=MIXED  VAR=PRES
      DEFINE PTLGAS     SUBSTREAM-VAR  STREAM=TAILGAS    SUBS=MIXED  VAR=PRES
      DEFINE PHPTL      SUBSTREAM-VAR  STREAM=HPTAIL     SUBS=MIXED  VAR=PRES
      DEFINE PFUELT     SUBSTREAM-VAR  STREAM=FUELTEMP   SUBS=MIXED  VAR=PRES
;
;   INLET SULFUR
;
      DEFINE XOGSO2     MOLE-FLOW      STREAM=SO2PROD    SUBS=MIXED  COMP=SO2
      DEFINE XSLSO2     MOLE-FLOW      STREAM=DSLIP      SUBS=MIXED  COMP=SO2
      DEFINE XSLH2S     MOLE-FLOW      STREAM=DSLIP      SUBS=MIXED  COMP=H2S
      DEFINE XSLCOS     MOLE-FLOW      STREAM=DSHFIN     SUBS=MIXED  COMP=COS
;
;   OUTLET SULFUR
;
      DEFINE XES1       MOLE-FLOW      STREAM=DES1       SUBS=CISOLID  COMP=SULFUR
      DEFINE XES2       MOLE-FLOW      STREAM=DES2       SUBS=CISOLID  COMP=SULFUR
      DEFINE XSPROD     MOLE-FLOW      STREAM=SULPROD    SUBS=CISOLID  COMP=SULFUR
      DEFINE XTLSO2     MOLE-FLOW      STREAM=HPTAIL     SUBS=MIXED   COMP=SO2

```

```

DEFINE XTLH2S MOLE-FLOW STREAM=HPTAIL SUBS=MIXED COMP=H2S
;
; Print only after flowsheet simulation is completed.
;
F IF (IPASS.EQ.1) RETURN
;
F CALL IDATE(IM, ID, IY)
F CALL TIME(TIM)
F WRITE(NRPT,101) MONTH(IM), ID, IY, TIM(1:5)
F 101 FORMAT(/,22X,'DIRECT SULFUR RECOVERY PROCESS',
F + /19X,'RESULTS GENERATED ON ',A,I3', 19'I2' AT ',A/)
;
F WRITE(NRPT,102)
F 102 FORMAT(26X,'** DSRP MASS BALANCE (LB/HR) **',/)
;
F WRITE(NRPT,103) XSO2GA, TSO2GA, PSO2GA,
F 1 XZFCLF, TZFCLF, PZFCLF,
F 2 XSLIP, TSLIP, PSLIP,
F 3 XCOOLI, TCOOLI, PCOOI,
F 4 XSHFIN, TSHFIN, PSHFIN,
F 5 XR1IN, TR1IN, PR1IN,
F 6 XC1IN, TC1IN, PC1IN
;
F 103 FORMAT(26X,' DSRP GAS STREAMS',/
F & 10X,' Mass Flow Temp. Pres. ',/
F 1 10X,' Description (lb/hr) (deg F) (psia) ',/
F 2 10X,'-----',/
F & 10X,'HGD OFF-GAS ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'COAL GAS ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'SLIP STREAM ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'COOLER INLET ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'SHIFT INLET ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'REACTOR 1 INLET ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'CONDENSER 1 INLET ',3X,F11.2,4X,F9.0,1X,F9.2)
;
F WRITE(NRPT,104) XC1OUT, TC1OUT, PC1OUT,
F 1 XRHIN, TRHIN, PRHIN,
F 2 XR2IN, TR2IN, PR2IN,
F 3 XC2IN, TC2IN, PC2IN,
F 4 XC2OUT, TC2OUT, PC2OUT,
F 5 XTLGAS, TTLGAS, PTLGAS,
F 6 XHPTL, THPTL, PHPTL,
F 6 XFUELT, TFUELT, PFUELT
;
F 104 FORMAT(
F & 10X,'CONDENSER 1 OUTLET',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'REHEATER INLET ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'REACTOR 2 INLET ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'CONDENSER 2 INLET ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'CONDENSER 2 OUTLET',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'DSRP TAILGAS ',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'COMPRESSED TAILGAS',3X,F11.2,4X,F9.0,1X,F9.2,/
F & 10X,'FUEL TO GAS TURB. ',3X,F11.2,4X,F9.0,1X,F9.2,/)
;
F WRITE(NRPT,105)
F 105 FORMAT(/,20X,'** DSRP SULFUR MASS BALANCE (LBMOLE/HR) **',/)
;
F XSITOT = XOGSO2 + XSLSO2 + XSLH2S + XSLCOS

```

```

F      WRITE(NRPT,106) XOGSO2, XLSO2, XSLH2S, XSLCOS, XSITOT
F 106 FORMAT(
F      & 20X,'INLET SULFUR:          SO2 IN OFF-GAS ',F10.2,/
F      & 20X,'                      SO2 IN SLIP STREAM ',F10.2,/
F      & 20X,'                      H2S IN SLIP STREAM ',F10.2,/
F      & 20X,'                      COS IN SLIP STREAM ',F10.2,/
F      & 20X,'                      TOTAL INLET SULFUR---->',F10.2,/
F      ;
F      FSCAP = (XSPROD/XSITOT)*100.
F      XSOTOT = XES1+XES2+XTLSO2+XTLH2S
F      WRITE(NRPT,107) XES1, XES2, XTLSO2, XTLH2S, XSOTOT, FSCAP
F 107 FORMAT(
F      & 20X,'OUTLET SULFUR:      CONDENSER NO. 1 ',F10.2,/
F      & 20X,'                      CONDENSER NO. 2 ',F10.2,/
F      & 20X,'                      SO2 IN TAILGAS ',F10.2,/
F      & 20X,'                      H2S IN TAILGAS ',F10.2,/
F      & 20X,'                      TOTAL OUTLET SULFUR---->',F10.2,//
F      & 20X,'DSRP SULFUR CAPTURE EFFICIENCY: ',F6.2,' PERCENT')
F      ;
F      WRITE(NRPT,100) XSO2GA, XZFCLF, XSLIP, XCOOLI, XSHFIN,
F      1              XR1IN, XC1IN
C
F 100 FORMAT(26X,'DSRP GAS STREAMS',/
F      &          'HGD Off-gas: ',F14.2,/
F      &          'Coal Gas: ',F14.2,/
F      &          'Coal Gas Slip Stream: ',F14.2,/
F      &          'Gas Into Cooler: ',F14.2,/
F      &          'Gas Into Shift: ',F14.2,/
F      &          'Reactor No. 1 Inlet: ',F14.2,/
F      &          'Condenser No. 1 Inlet: ',F14.2,/
F      ;
F      ;*****
F      ;          DSRP DESIGN SPECIFICATIONS
F      ;*****
F      ;
F      DES-SPEC DS-DSRPC
F      DEFINE QR1 STRM-ATTR-VAR ATTR=HEAT VAR=Q STREAM=QDSRP-R1
F      SPEC QR1 TO 0.
F      TOL-SPEC 1000.
F      VARY BVAR BLOCK=DSRP-COL SENTENCE=PARAM VAR=TEMP
F      LIMITS 500. 1292.
F      ;

```


Appendix C
DSRP COST MODEL CODE

This appendix contains a listing of the FORTRAN code from the cost model subroutine for a Lurgi-based IGCC system which includes the DSRP. Only the code specific to the DSRP is included here.

Direct Capital Cost of the DSRP

```

C
C   Direct capital cost of the DIRECT SULFUR RECOVERY PROCESS
C
C   NTDS   = Number of total trains of DSRP units
C   NODS   = Number of on-line (operating) trains of DSRP units
C   NDSSTG = Number of DSRP stages (reactors): 1 or 2
C   RSV1DR = Reference space velocity for Reactor No. 1 (scc/(cc-
hr))
C   RSV2DR = Reference space velocity for Reactor No. 2 (scc/(cc-
hr))
C   FDSPRE = DSRP Pressure correction factor for pressure vessel
cost
C   XPR1IN = Reactor No. 1 inlet gas pressure, psia
C   XTR1IN = Reactor No. 1 inlet gas temperature, F
C   VGR1IN = Reactor No. 1 inlet gas volume flow rate, ft^3/hr
C   FDSPR2 = DSRP Pressure correction factor for pressure vessel
cost
C   XPR2IN = Reactor No. 2 inlet gas pressure, psia
C   XTR2IN = Reactor No. 2 inlet gas temperature, F
C   VGR2IN = Reactor No. 2 inlet gas volume flow rate, ft^3/hr
C   DCDSG1 = Direct cost for stage 1 gas handling equipment, $1,000
C   DCDSG2 = Direct cost for stage 2 gas handling equipment, $1,000
C   DCDSS1 = Direct cost for stage 1 sulfur handling equipment,
$1,000
C   DCDSS2 = Direct cost for stage 2 sulfur handling equipment,
$1,000
C
C   NTDS = 2
C   NODS = 2
C   NDSSTG = 2
C   RSV1DR = 2000.
C   RSV2DR = 2000.
C
C   also define space velocities. units are scc/(cc-h).
C
C   DSRSV1 = 5000.
C   DSRSV2 = 5000.
C
C   pressure correction factor for pressure vessel cost
C
C   FDSPRE = (2.5 + 0.093 * XPR1IN ** 0.64)/7.1
C
C   calculate gas volume flow through DSRP, units are ft^3/hr
C
C   VGR1IN = XR1IN * 0.73 * (XTR1IN+460)/ (XPR1IN/14.7)
C
C   calculate cost for the 1st stage of the DSRP
C   The first term represents the gas cooler and sulfur condenser
C   The second term represents the reactor vessel. The latter is
C   sensitive to the gas volume flow rate and the space velocity
C
C   DCDSG1 = NTDS*FDSPRE*(3380.*(VGR1IN/(77000.*NODS))**0.67

```

```

&          + 1020.*(VGR1IN/(77000.*NODS)*(RSV1DR/DSRSV1))**0.67)
&          *(CI/361.2)
DCDSS1 = 80.*NTDS*(XSPROD/(2900.*NODS))**0.67*(CI/361.2)
C
C calculate cost for the 2nd stage of the DSRP
C
DCDSG2 = 0.
DCDSS2 = 0.
C
IF (NDSSTG.EQ.2) THEN
VGR2IN = XR2IN * 0.73 * (XTR2IN+460)/ (XPR2IN/14.7)
FDSR2 = (2.5 + 0.093 * XPR2IN ** 0.64)/7.1
DCDSG2 = NTDS*FDSR2*(1205.*(VGR2IN/(69000.*NODS))**0.67
&          + 960.*(VGR2IN/(69000.*NODS)*(RSV2DR/DSRSV2))**0.67)
&          *(CI/361.2)
DCDSS2 = 80.*NTDS*(XSPROD/(2900.*NODS))**0.67*(CI/361.2)
ENDIF
C
DCDS = DCDSG1 + DCDSG2 + DCDSS1 + DCDSS2
C

```

Other Capital Costs

```

C
C Direct capital cost of general facilities
C
DCGF=FGF*(DCCH+DCL +DCOF+DCG +DCSS+DCCY+DCZF+DCSA+DCS +DCBF
&          +DCGT+DCHR+DCDS+DCCR+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+DCDS+DCCR+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)
TIC=CICC+CTAX+CEHO+CEP
C
C Process Contingency Costs
C
CPCDS = FPCDS*DCDS*(1.+TIC/TDC)
C
CPC=CPCCH+CPCL+CPCOF+CPCSS+CPCCY+CPCG+CPCS+CPCZF+CPCSA
&          +CPCBF+CPCGT+CPCHR+CPCDS+CPCCR+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       IF(Z.LE.1) THEN
C           AF = 1
C       ELSE
C           AF=(Z**NOYEARS-1)/(NOYEARS*(Z-1))
C       ENDIF
C       TPI=AF*TPC
C       AFDC=TPI-TPC
C

```

Net Plant Electrical Output (including DSRP auxiliary loads)

```

C
C   ----- NET PLANT ELECTRICAL OUTPUT -----
C
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
C   Rubin (1990).
C
C   Gas Turbine Output:  WGTE           Steam Turbine Output:  WSTE
C
C   The auxiliary loads have been converted to MW from Frey and Rubin
C   (1990)
C
C   WCHE = 1.89D-06 * MCFGI + ERRWCH
C   WLE = 0.0
C   IF (NTL.GT.0) WLE = (-13.3 + 0.00461*MLGI)/1000.
C   WOFE = WCME/1000.
C   WGE = (0.005*MCFGI)/1000.
C   WSE = 0.0
C   IF (NTS.GT.0) WSE = (-3.7 + 0.0102*(MDISP1+MDISP2))/1000.
C   WSCR = 3.47D-5 * XNH3IN * 2000./(8760. * CF)
C   WDSRP = WKDSRP/1000000.
C
C   WBFE = STAUXE/1000000.0
C   WGFE = 1.64 + 0.0877*(WCHE+WLE+WOFE+WGE+WSE+WBFE)
C   WAUXE = WCHE+WLE+WOFE+WGE+WSE+WSAE+WBFE+WGFE+WDSRP+WSCR
C   WNETE = WGTE+WSTE-WAUXE

```

```

C ----- INITIAL CATALYSTS AND CHEMICALS -----
C
C DSRP: the total catalyst charge is:
C NOTE: need to convert from lbmole/hr to scc/hr
C lbmole/hr * 0.73 (atm-ft^3)/(lbmole-R)
C * 520 R / 1 atm = ft^3/hr
C mult by 28,350 cc/ft^3 to get scc/hr
C Thus, lbmole/hr * 379.6 ft^3/lbmole * 28,350 cc/ft^3
C or lbmole/hr * 10,762,000 cc/lbmole = scc/hr
C Space velocity is scc/(cc-hr), thus cat volume is cc.
C Convert to weight basis to be consistent with cost data.
C cc * (0.72 g/cc (typical density) * lb/454 g) = lb
C or cc * 0.00159 = lb
C TDSCAT has units of lb.
C
C TDSCAT = (XR1IN/DSRSV1 + XR2IN/DSRSV2) * 10762000. * 0.00159
C

```

```
C
C      Fixed Operating Costs:  Maintenance
C
C      OCMDS = FMCDS * (DCDS * (1+(TIC+CPJ)/TDC) + CPCDS)
```

```
C
C Annual DSRP Catalyst Requirement
C
C ADSCAT = TDSCAT/3.
C
C
C OCDSCA = UCDSCA * ADSCAT
```


Appendix D

COST MODEL OUTPUTS

- a) Base Case**
(Lurgi IGCC with sulfuric acid recovery)
- b) One-Stage DSRP**
- c) Two-Stage DSRP**

Computer Model Output for Base Case Analysis: Lurgi-based IGCC with conventional sulfuric acid recovery

1 Air-Blown Fixed-Bed Gasifier IGCC w/SCR: Deterministic Analysis
SYSTEM SUMMARY

*** GASIFIER CONDITIONS ***

COAL FLOW RATE: 0.562244E+06 LB/HR
SORBENT FLOW RATE: 0.378160E+00 LB/HR
AIR FLOW RATE: 0.138042E+07 LB/HR
REQUIRED STEAM: 0.690210E+06 LB/HR
SUPPLIED STEAM: 0.998983E-03 LB/HR
SO2 FLOW RATE: 0.999983E-06 LB/HR
ASH FLOW RATE: 49220.832652 LB/HR
GASIFIER PRESSURE: 307.5 PSIA
GASIFIER TEMPERATURE: 1100.0 F

1 Air-Blown Fixed-Bed Gasifier IGCC w/SCR: Deterministic Analysis
SYSTEM SUMMARY

ZINC FERRITE SULFUR REMOVAL PROCESS
RESULTS GENERATED ON OCT 17, 1994 AT 17:37

** SO2 BALANCE (LB/HR) **

H2S IN RAWGAS 19720.09
H2S IN FUEL GAS 41.41
SO2 IN SO2 GAS 36991.65
PPMV OF H2S IN FUEL GAS 11.

** FLOW RATES (LB/HR) **

RAWGAS 0.26D+07 MW = 23.11
FUEL GAS 0.26D+07 MW = 23.08
SO2 GAS 548541.39 MW = 20.35
STEAM REQUIRED TO MAINTAIN
CO/CO+CO2 RATIO (SHFT-STM) 1.00
DILUENT REQUIRED TO MAINTAIN
REGENERATOR TEMPERATURE (GEN-STM) 408906.11
AIR FLOW TO REGENERATOR (REGENAIR) 133438.72

** GENERAL **

CO/(CO+CO2) RATIO: 0.281
SULFUR CAPTURE EFFICIENCY: 0.998
TEMPERATURE OF THE REGENERATED GAS: 1450.2
VOLUME PERCENT AIR IN REGENERATION GAS: 16.765

*** MS7000F GAS TURBINE CONDITIONS ***

FUEL FLOW RATE: 0.257043E+07 LB/HR
AIR FLOW RATE: 0.987479E+07 LB/HR
AIR DIVERTED TO GASIFIER: 0.138042E+07 LB/HR
AIR DIVERTED TO ZINC FER: 0.133436E+06 LB/HR
FUEL LHV (WET BASIS): 1726.5 BTU/LB, 105.5 BTU/SCF
FUEL HHV (WET BASIS): 1938.9 BTU/LB, 117.9 BTU/SCF
FIRING TEMPERATURE: 2350.0 F
COMBUSTOR EXIT TEMPERATURE: 2436.7 F
TURBINE EXHAUST TEMPERATURE: 1135.1 F
THERMAL EFFICIENCY (LHV): 0.4308
GENERATOR EFFICIENCY: 0.9850
GASIFIER AIR INTERCOOLER DUTY: 0.000000E+00 BTU/HR

Ratio of Inlet Air to Exhaust Gas: 0.901
Ratio of Boost Air to Inlet Air: 0.153
Ratio of Gasif Air to Inlet Air: 0.140
Ratio of Regen Air to Inlet Air: 0.014
Ratio of Fuel Gas to Inlet Air: 0.260
CO2 Emissions: 194.3 LB/MMBTU
CO Emissions: 0.32281 LB/MMBTU
CO Emissions: 186.60 ppmv
SO2 Emissions: 0.01217 LB/MMBTU
NOx Emissions: 0.54707 LB/MMBTU
NOx Emissions: 192.49 ppmv
Oxygen in Exhaust: 9.8814 percent
Vapor in Exhaust: 16.5883 percent
CO2 in Exhaust: 7.1489 percent

*** STEAM TURBINE CONDITIONS ***

STEAM DIVERTED TO GASIFIER: 0.690210E+06 LB/HR
STEAM DIVERTED TO ZINC FER: 0.408899E+06 LB/HR
STEAM TURBINE FLOW RATE: 0.129455E+07 LB/HR
SUPERHEATED STEAM TEMPERATURE: 1000.0 F
EXPANDED STEAM QUALITY: 0.9000
ISENTROPIC EFFICIENCY: 0.9227
GENERATOR EFFICIENCY: 0.9850

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

PLANT SECTION	FLOWSHEET PERFORMANCE PARAMETERS	VALUE UNITS
Coal Feed	Mass flow of coal to gasifier	562252.96 lb/hr
Limestone Feed	Mass flow of limestone (mixed)	0.00 lb/hr
	Mass flow of limestone (cisolid)	0.38 lb/hr
Oxidant Feed	Work to boost air compressor	0.138840E+05 Watts
	First precool inlet air temp.	750.33 F
	First precool outlet air temp.	679.51 F
	First precool inlet BFW temp.	505.54 F
	First precool outlet BFW temp.	605.29 F
	Heat transfer in first precool	0.277075E+08 BTU/hr
	Heat trans. to BFW from HRSG	0.577650E+09 BTU/hr
	Heat trans. to BFW from Regen.	0.181605E+09 BTU/hr
	Heat leaving from economizer	0.614260E+09 BTU/hr
Gasification	Gasifier output syngas pres.	307.50 psia
	Gasifier output syngas temp.	1100.00 F
	Gasifier output syngas density	0.183078E-01 lbmole/ft3
	Gasifier recycle fines flow rate	18694.91 lb/hr
Zinc Ferrite	Zinc Ferrite inlet syngas flow	111385.75 lbmole/hr
	Off-gas to sulfuric acid plant	26954.95 lbmole/hr
	SO2 to sulfuric acid plant	577.43 lbmole/hr
	Off-gas temp. to sul. acid plant	1233.44 lbmole/hr
	Zinc ferrite inlet H2S flow	578.64 lbmole/hr
	Zinc ferrite inlet COS flow	0.00 lbmole/hr
Sulfation	Gasifier ash removal	49221.66 lb/hr
	Gasifier ash removal	8817.64 lb/hr
	Gasifier fines removal	245.99 lb/hr
	Gasifier fines removal	737.96 lb/hr
BFW Treating	Raw water to power plant	1247373.27 lb/hr
	Steam turbine condensate	750596.08 lb/hr
Gas Turbine	Gas turbine inlet air flow	9874952.61 lb/hr
Steam Cycle	HRSG outlet HP steam pres.	1465.00 psia
	HRSG outlet HP steam flow	1294574.35 lb/hr
	Gas turbine net shaft work	-571417E+09 Watts
	Steam turbine net shaft work	-175968E+09 Watts
	Steam cycle auxiliary power	0.356242E+07 Watts
	Acid gas auxiliary power	0.000000E+00 Watts
Miscellaneous	Moisture in coal feed	0.12 wt %
	Ash in coal	0.10 wt %
	Temp. of ambient air	59.00 F
	Heating value of coal	12774.00 BTU/lb
SCR	SCR Inlet Gas Flow	390773.89 lbmole/hr
	SCR Inlet Gas Temperature	717.00 lbmole/hr
	SCR Inlet Gas Pressure	15.33 lbmole/hr

Environmental	SCR Inlet NO loading	357.29 lbmole/hr
	SCR Inlet NO2 loading	18.80 lbmole/hr
	SCR Ammonia Requirement	304.78 lbmole/hr
	SCR Dilution Steam Requirement	5790.83 lbmole/hr
	High Pressure Blowdown	26419.88 lbmole/hr
	Low Pressure Blowdown	26419.88 lbmole/hr
	CO2 from gas turbine	27936.17 lbmole/hr
	CO from gas turbine	72.92 lbmole/hr
	SO2 from gas turbine	1.20 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	397.41 lbmole/hr
	NO from gas turbine	71.46 lbmole/hr
	NO2 from gas turbine	3.76 lbmole/hr
Actual coal heating value		11248.73 BTU/lb
COST VAR WARNING ---- Variable MCFG1 value of		562252.961
in DCSS above the upper limit of		433000.000
COST VAR WARNING ---- Variable VSNZF1 value of		9218.250
in DCCY above the upper limit of		6000.000
COST VAR WARNING ---- Variable MRW value of		1247373.267
in DCBF above the upper limit of		614000.000

COST SUMMARY

Air Blown Lurgi-Based IGCC System with Hot Gas Cleanup

A. KEY INPUT ASSUMPTIONS

Performance Assumptions:

Gasifier Availability:	0.87	Sorb. Sulfur Loading:	0.168
Max. Desulf. Vessel Diameter:	12.50	Superficial Velocity:	2.00
L/D Ratio:	3.00	Absorption Cycle Time:	30.0
Maximum Space Velocity (1/hr):	2000.	Sorbent Bulk Density:	82.0
Sorb. Attrition Rate/80 cycles:	0.800	Generator Efficiency:	0.9850
Gasifier Coal Throughput:	1.01		

Economic Assumptions:

Cost Year:	January 1989	Inflation Rate:	0.000
Plant Cost Index:	351.5	Real Escalation Rate:	0.000
Chemicals Cost Index:	411.3	Plant Booklife:	30
Plant Capacity Factor:	0.65	Sales Tax Rate:	0.05
General Facilities Factor:	0.20	Real Return on Debt:	0.046
Indirect Construction Factor:	0.20	Real Ret. on Pref.:	0.052
Engr & Home Office Fees:	0.10	Real Ret. on Equity:	0.087
Project Contingency Factor:	0.17	Debt Ratio:	0.500
Byproduct Marketing Factor:	0.10	Pref. Stock Ratio:	0.150
Average Operating Labor Rate:	19.70	Fed. & State Taxes:	0.380
Number of Shifts:	4.25	Investment Tax Credit:	0.000
Construction Interest Rate:	0.10	Prop. Taxes & Insur.:	0.020
Years of Construction:	4		

Process Contingency and Maintenance Cost Factors:

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Limestone Handling	0.000	0.030
Oxidant Feed	0.100	0.020
Gasification	0.200	0.030
Coke, Ash, & Bent. Subsystems	0.050	0.020
High Temp. Cyclones	0.050	0.030
Zinc Ferrite	0.400	0.030
Sulfuric Acid Plant	0.100	0.020
Sulfation	0.600	0.045
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.250	0.020
Heat Recovery Steam Generator	0.025	0.015
Selective Catalytic Reduction	0.100	0.025
Steam Turbine	0.015	0.050
General Facilities	0.015	

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

Plant Section	Number of Units		Direct Operating Total Capital Cost	Process Contingency
	Operating	Total		
Coal Handling	1	1	22670.	1550.
Limestone Handling	0	0	0.	0.
Oxidant Feed	3	3	4861.	665.
Gasification	11	13	86034.	23531.
Coke, Ash, & Bent. Subsystems	11	13	15668.	1071.
High Temp. Cyclones	22	26	6421.	439.
Zinc Ferrite	11	26	12558.	6870.
Sulfuric Acid Plant	1	1	25127.	3436.
Sulfation	0	0	0.	0.
Boiler Feedwater Treatment	1	1	5937.	0.
Gas Turbine	3	3	96000.	32821.
Heat Recovery Steam Generator	3	3	28645.	979.
Selective Catalytic Reduction	3	3	10023.	1371.
Steam Turbine	1	1	27507.	940.
General Facilities	N/A	N/A	68290.	4669.

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

Description	Capital Cost
Total Direct Cost	409741.
Indirect Construction Cost	81948.
Sales Tax	16799.
Engineering and Home Office Fees	50849.
Environmental Permitting	1000.
Total Indirect Costs	150596.
Total Process Contingencies	78343.
Project Contingency	111769.
Total Plant Cost	750449.
AFDC	120259.
Total Plant Investment	870708.
Preproduction (Startup) Costs	24298.
Inventory Capital	20751.
Initial Catalysts and Chemicals	35511.
Land	2613.
TOTAL CAPITAL REQUIREMENT (\$1,000) ----->	962589.

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

Description	Annual Cost
Operating Labor	8533252.
Maintenance Costs	16325007.
Administration and Supervision	4518976.
TOTAL FIXED OPERATING COST (\$/year) ----->	29377235.

E. CALCULATED VARIABLE OPERATING COSTS -----

1. CONSUMABLES (\$/year)

Description	Assumed Unit Cost	Calc. Material Requirement	Calc. Annual Operating Cost
Sulfuric Acid:	110.00 \$/ton	2418.3 ton/yr	266017.
NaOH:	220.00 \$/ton	505.2 ton/yr	111137.
Na2 HPO4:	0.70 \$/lb	3001.7 lb/yr	2101.
Hydrazine:	3.20 \$/lb	14451.6 lb/yr	46245.
Morpholine:	1.30 \$/lb	13488.9 lb/yr	17536.
Lime:	80.00 \$/ton	576.0 ton/yr	46077.
Soda Ash:	160.00 \$/ton	638.5 ton/yr	102155.
Corrosion Inh.:	1.90 \$/lb	114670.5 lb/yr	217874.
Surfactant:	1.25 \$/lb	114670.5 lb/yr	143338.
Chlorine:	250.00 \$/ton	18.2 ton/yr	4543.
Biocide:	3.60 \$/lb	20201.2 lb/yr	72724.
SCR Catalyst:	250.00 \$/ft^3	5955.4 ft^3/yr	1488856.
Ammonia:	150.00 \$/ton	14777.1 ton/yr	2216566.
Zinc Fer Sorb:	3.00 \$/lb	2960533.4 lb/yr	8881600.
Limestone:	18.00 \$/ton	0.0 ton/yr	0.
Sul Acid. Cat.:	1.90 \$/l	29650.4 l/yr	56336.
Coke:	150.00 \$/ton	1562.8 ton/yr	234418.

Bentonite:	0.03 \$/lb	4577299.7 lb/yr	132742.
Fuel Oil:	42.00 \$/bbl	40245.9 bbl/yr	1690328.
Plant Air Ads.:	2.80 \$/lb	3018.4 lb/yr	8452.
Raw Water:	0.73 \$/Kgal	851394.6 Kgal/yr	621518.
Waste Water:	840.00 \$/gpm ww	0.0 lb/hr	0.
LPG - flare:	11.70 \$/bbl	3521.5 bbl/yr	41202.
TOTAL CONSUMABLES (\$/year)			16401766.

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

Coal:	1.61 \$/MMBtu	562253.0 lb/hr	65841846.
Ash Disposal:	10.00 \$/ton	708.3 ton/day	1680392.
Sulfuric Acid:	40.00 \$/ton	173231.5 ton/yr	6236336.)

TOTAL VARIABLE OPERATING COST (\$/year) -----> 77687668.

F. CALCULATED COST OF ELECTRICITY

Power Summary (MWe)		Auxiliary Loads (MWe)	
Gas Turbine Output	562.85	Coal Handling	1.06
Steam Turbine Output	173.33	Limestone Hdlg	0.00
Total Auxiliary Loads	26.71	Oxidant Feed	13.88
		Gasification	2.81
Net Electricity	709.46	Zinc Ferrite	0.00
		General Fac	3.51

Capital Cost: 1356.78 \$/kW

Fixed Operating Cost: 41.41 \$/(kW-yr)

Incremental Variable Costs: 4.48 mills/kWh

Byproduct Credit: 1.54 mills/kWh

Fuel Cost: 16.30 mills/kWh

Variable Operating Cost: 19.23 mills/kWh

COST OF ELECTRICITY -----> 51.14 mills/kWh

Fixed Charge Factor: 0.1034 Variable Cost Levelization Factor: 1.0000

The plant heatrate (HHV) is: 8915. BTU/kWh. Efficiency: 0.3831

G. ENVIRONMENTAL SUMMARY

INPUTS:	Coal	0.7925 lb/kWh
	Water	1.7582 lb/kWh
OUTPUTS:	Blowdown	0.0745 lb/kWh
	Ash	0.0832 lb/kWh
	Sulfuric Acid	0.0858 lb/kWh
	CO ₂ emissions	1.7326 lb/kWh
	CO emissions	0.0029 lb/kWh
	SO ₂ emissions	0.0414 lb/MMBtu
	COS emissions	0.0000 lb/MMBtu
	CH ₄ emissions	0.0000 lb/kWh
	H ₂ S emissions	0.0000 lb/MMBtu
	NO _x emissions	0.5471 lb/MMBtu

NOTE: NO_x emissions are based on fuel bound ammonia

I. SCR SUMMARY

Reference Removal Efficiency	0.80 fraction
Actual Removal Efficiency	0.80 fraction
Ammonia Slip	10.00 ppmv
Reference NO _x Concentration	42.00 ppmv
Minimum Relative Cat. Activ.	842.12 ratio
Activity Time Constant	12638.01
Replacement Interval	11388.00 hrs
Reference Temperature	717.00 F
Actual Temperature	717.00 F
Dilution Steam/NH ₃ Ratio	19.00 lbmole/lbmole
NH ₃ /NO _x Ratio	0.81 lbmole/lbmole
Fraction NO _x as NO	0.95
Reference Replacement Interv.	17082.00 hrs
SO ₂ Oxidation Rate	0.01 fraction
SCR Pressure Drop	0.14 psi
Reference Space Velocity	10000.00 1/hr
Calculated Space Velocity	4150.95 1/hr
No. Initial Catalyst Layers	3
No. Total Catalyst Layers	3

No. Layers Replaced at a time	1
Factor, efficiency	1.00
Factor, end mole ratio	0.3908
Factor, activity	1.0621
Factor, temperature	1.0000
Initial Catalyst Volume	35732.55 ft ³
Total Catalyst Volume	35732.55 ft ³
Reference Cat. Volume	14832.39 ft ³
Annual Layers Replaced	0.50
Annual Volume Replaced	5955.43 ft ³ /yr
Catalyst Life, yrs	6.00 yrs
Direct Capital Cost	10022.92
Ammonia Flow, tons/yr	304.78

Computer Model Output for One-Stage DSRP Analysis: Lurgi-based IGCC with One-Stage DSRP

Air-Blown Fixed-Bed Gasifier IGCC w/SCR: Deterministic Analysis
SYSTEM SUMMARY

*** GASIFIER CONDITIONS ***

COAL FLOW RATE: 0.663869E+06 LB/HR
SORBENT FLOW RATE: 0.236712E+01 LB/HR
AIR FLOW RATE: 0.162993E+07 LB/HR
REQUIRED STEAM: 0.814966E+06 LB/HR
SUPPLIED STEAM: 0.998983E+03 LB/HR
SO2 FLOW RATE: 0.999983E-06 LB/HR
ASH FLOW RATE: 58122.759878 LB/HR
GASIFIER PRESSURE: 1307.5 PSIA
GASIFIER TEMPERATURE: 1100.0 F

ZINC FERRITE SULFUR REMOVAL PROCESS
RESULTS GENERATED ON JUL 6, 1995 AT 22:19

** SO2 BALANCE (LB/HR) **

H2S IN RAWGAS 23286.81
H2S IN FUEL GAS 48.90
SO2 IN SO2 GAS 43682.21
PPMV OF H2S IN FUEL GAS 11.

** FLOW RATES (LB/HR) **

RAWGAS 0.30D+07 MW = 23.11
FUEL GAS 0.30D+07 MW = 23.08
SO2 GAS 647678.12 MW = 20.35
STEAM REQUIRED TO MAINTAIN
CO/CO+CO2 RATIO (SHIFT-STM) 1.00
DILUENT REQUIRED TO MAINTAIN
REGENERATOR TEMPERATURE (GEN-STM) 482787.44
AIR FLOW TO REGENERATOR (REGENAIR) 157573.38

** GENERAL **

CO/(CO+CO2) RATIO: 0.281
SULFUR CAPTURE EFFICIENCY: 0.998
TEMPERATURE OF THE REGENERATED GAS: 1450.2
VOLUME PERCENT AIR IN REGENERATION GAS: 16.768

DIRECT SULFUR RECOVERY PROCESS
RESULTS GENERATED ON JUL 6, 1995 AT 22:19

** DSRP MASS BALANCE (LB/HR) **

Description	DSRP GAS STREAMS		
	Mass Flow (lb/hr)	Temp. (deg F)	Pres. (psia)
HGD OFF-GAS	647678.12	1234.	305.50
COAL GAS	3031798.68	1206.	286.72
SLIP STREAM	136108.52	1206.	286.72
COOLER INLET	783786.64	1228.	286.72
SHIFT INLET	783786.64	1147.	285.72
REACTOR 1 INLET	783786.47	1160.	284.72
CONDENSER 1 INLET	763091.87	1292.	259.72
CONDENSER 1 OUTLET	763091.87	350.	249.72
REHEATER INLET	763091.87	350.	249.72
REACTOR 2 INLET	763091.87	350.	249.72
CONDENSER 2 INLET	763091.87	350.	249.72
CONDENSER 2 OUTLET	763091.87	350.	249.72

DSRP TAILGAS	763091.87	350.	249.72
COMPRESSED TAILGAS	763091.87	360.	286.72
FUEL TO GAS TURB.	3662061.73	801.	286.72

** DSRP SULFUR MASS BALANCE (LBMOLE/HR) **

INLET SULFUR:	SO2 IN OFF-GAS	681.86
	SO2 IN SLIP STREAM	0.00
	H2S IN SLIP STREAM	0.06
	COS IN SLIP STREAM	0.00
TOTAL INLET SULFUR----		681.93

OUTLET SULFUR:	CONDENSER NO. 1	645.42
	CONDENSER NO. 2	0.00
	SO2 IN TAILGAS	13.64
	H2S IN TAILGAS	18.48
TOTAL OUTLET SULFUR----		677.53

DSRP SULFUR CAPTURE EFFICIENCY: 94.65 PERCENT

DSRP GAS STREAMS

HGD Off-gas:	647678.12
Coal Gas:	3031798.68
Coal Gas Slip Stream:	136108.52
Gas Into Cooler:	783786.64
Gas Into Shift:	783786.64
Reactor No. 1 Inlet:	783786.47
Condenser No. 1 Inlet:	763091.87

*** MS7000F GAS TURBINE CONDITIONS ***

FUEL FLOW RATE:	0.366200E+07 LB/HR
AIR FLOW RATE:	0.874772E+07 LB/HR
AIR DIVERTED TO GASIFIER:	0.162993E+07 LB/HR
AIR DIVERTED TO ZINC FER:	0.157571E+06 LB/HR
FUEL LHV (WET BASIS):	1392.2 BTU/LB, 83.0 BTU/SCF
FUEL HHV (WET BASIS):	1563.0 BTU/LB, 92.7 BTU/SCF
FIKING TEMPERATURE:	2350.0 F
COMBUSTOR EXIT TEMPERATURE:	2421.5 F
TURBINE EXHAUST TEMPERATURE:	1176.3 F
THERMAL EFFICIENCY (LHV):	0.4272
GENERATOR EFFICIENCY:	0.9850
GASIFIER AIR INTERCOOLER DUTY:	0.000000E+00 BTU/HR

Ratio of Inlet Air to Exhaust Gas:	0.821
Ratio of Boost Air to Inlet Air:	0.204
Ratio of Gasif Air to Inlet Air:	0.186
Ratio of Regen Air to Inlet Air:	0.018
Ratio of Fuel Gas to Inlet Air:	0.419
CO2 Emissions:	194.2 LB/MMBTU
CO Emissions:	0.30849 LB/MMBTU
CO Emissions:	209.94 ppmv
SO2 Emissions:	0.32141 LB/MMBTU
NOx Emissions:	0.54071 LB/MMBTU
NOx Emissions:	223.98 ppmv
Oxygen in Exhaust:	6.4380 percent
Vapor in Exhaust:	26.0943 percent
CO2 in Exhaust:	8.4090 percent

*** STEAM TURBINE CONDITIONS ***

STEAM DIVERTED TO GASIFIER:	0.814966E+06 LB/HR
STEAM DIVERTED TO ZINC FER:	0.482779E+06 LB/HR
STEAM TURBINE FLOW RATE:	0.224906E+07 LB/HR
SUPERHEATED STEAM TEMPERATURE:	1000.0 F
EXPANDED STEAM QUALITY:	0.9000
ISENTROPIC EFFICIENCY:	0.9227
GENERATOR EFFICIENCY:	0.9850

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

PLANT SECTION	FLOWSHEET PERFORMANCE PARAMETERS	VALUE	UNITS
Coal Feed	Mass flow of coal to gasifier	663880.32	lb/hr
Limestone Feed	Mass flow of limestone (mixed)	0.00	lb/hr
	Mass flow of limestone (cisolid)	2.37	lb/hr
Oxidant Feed	Work to boost air compressor	0.163937E+05	Watts
	First precooler inlet air temp.	750.33	F
	First precooler outlet air temp.	679.51	F
	First precooler inlet BFW temp.	505.47	F
	First precooler outlet BFW temp.	605.29	F
	Heat transfer in first precooler	0.327159E+08	BTU/hr
	Heat trans. to BFW from HRSG	0.593697E+09	BTU/hr
	Heat trans. to BFW from Regen.	0.000000E+00	BTU/hr
	Heat leaving from economizer	0.326186E+09	BTU/hr
Gasification	Gasifier output syngas pres.	307.50	psia
	Gasifier output syngas temp.	1100.00	F
	Gasifier output syngas dehsity	0.183078E-01	lbmole/ft3
	Gasifier recycle fines flow rate	22074.02	lb/hr
Zinc Ferrite	Zinc Ferrite inlet syngas flow	131518.33	lbmole/hr
	Off-gas to sulfuric acid plant	31825.97	lbmole/hr
	SO2 to sulfuric acid plant	681.86	lbmole/hr
	Off-gas temp. to sul. acid plant	1233.52	lbmole/hr
	Zinc ferrite inlet H2S flow	683.30	lbmole/hr
	Zinc ferrite inlet COS flow	0.00	lbmole/hr
Sulfation	Gasifier ash removal	58123.74	lb/hr
	Gasifier ash removal	10407.15	lb/hr
	Gasifier fines removal	290.45	lb/hr
	Gasifier fines removal	871.34	lb/hr
BFW Treating	Raw water to power plant	1474450.03	lb/hr
	Steam turbine condensate	891234.78	lb/hr
Gas Turbine	Gas turbine inlet air flow	8747865.05	lb/hr
Steam Cycle	HRSG outlet HP steam pres.	1465.00	psia
	HRSG outlet HP steam flow	2249098.84	lb/hr
	Gas turbine net shaft work	-.651019E+09	Watts
	Steam turbine net shaft work	-.247120E+09	Watts
	Steam cycle auxiliary power	0.522513E+07	Watts
	Acid gas auxiliary power	0.000000E+00	Watts
Miscellaneous	Moisture in coal feed	0.12	wt %
	Ash in coal	0.10	wt %
	Temp. of ambient air	59.00	F
	Heating value of coal	12774.00	BTU/lb
DSRP	DSRP Sulfur Product	20694.59	lb/hr
	DSRP Gas Cooler Inlet	37723.93	lbmole/hr
	DSRP Reactor 1 Inlet	37723.93	lbmole/hr
	DSRP Condenser 1 Inlet	37055.70	lbmole/hr
	DSRP Reheater Inlet	37055.70	lbmole/hr
	DSRP Reactor 2 Inlet	37055.70	lbmole/hr
	DSRP Condenser 2 Inlet	37055.70	lbmole/hr
	DSRP Compressor Work	0.109463E+07	Watts
	DSRP Reactor 2 Inlet Pressure	284.72	psia
	DSRP Reactor 2 Inlet Temperature	1160.20	F
SCR	SCR Inlet Gas Flow	391914.61	lbmole/hr
	SCR Inlet Gas Temperature	717.00	lbmole/hr
	SCR Inlet Gas Pressure	15.33	lbmole/hr
	SCR Inlet NO loading	416.97	lbmole/hr
	SCR Inlet NO2 loading	21.95	lbmole/hr
	SCR Ammonia Requirement	355.05	lbmole/hr
	SCR Dilution Steam Requirement	6745.91	lbmole/hr
Environmental	High Pressure Blowdown	45899.98	lbmole/hr
	Low Pressure Blowdown	45899.98	lbmole/hr
	CO2 from gas turbine	32956.29	lbmole/hr
	CO from gas turbine	82.28	lbmole/hr
	SO2 from gas turbine	37.50	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	469.24	lbmole/hr
	NO from gas turbine	83.39	lbmole/hr
	NO2 from gas turbine	4.39	lbmole/hr
	Actual coal heating value	11248.73	BTU/lb
COST VAR WARNING	---- Variable MCFGI value of	663880.324	

	in DCSS	above the upper limit of	433000.000
COST VAR WARNING	---- Variable VSNZFI value of		9209.889
	in DCCY	above the upper limit of	6000.000
COST VAR WARNING	---- Variable MRW value of		1474450.031
	in DCBF	above the upper limit of	614000.000
COST VAR WARNING	---- Variable MSH/N value of		749699.614
	in DCHR	above the upper limit of	640000.000
COST VAR WARNING	---- Variable MPW value of		2365684.813
	in MSABFP	above the upper limit of	2200000.000
COST VAR WARNING	---- Variable MCWI value of		7707.543
	in M**CWI	above the upper limit of	7700.000

COST SUMMARY
Air Blown Lurgi-Based IGCC System with Hot Gas Cleanup
Cost Model Version 1.63 Rev. Date: June 24, 1995

A. KEY INPUT ASSUMPTIONS

Performance Assumptions:				
Gasifier Availability:	0.87	Sorb. Sulfur Loading:	0.168	
Max. Desulf. Vessel Diameter:	12.50	Superficial Velocity:	2.00	
	L/D Ratio:	3.00	Absorption Cycle Time:	30.0
Maximum Space Velocity (1/hr):	2000.	Sorbent Bulk Density:	82.0	
Sorb. Attrition Rate/80 cycles:	0.800	Generator Efficiency:	0.9850	
Gasifier Coal Throughput:	1.01			

Economic Assumptions:

Cost Year:	January 1989	Inflation Rate:	0.000
Plant Cost Index:	351.5	Real Escalation Rate:	0.000
Chemicals Cost Index:	411.3	Plant Booklife:	30
Plant Capacity Factor:	0.65	Sales Tax Rate:	0.05
General Facilities Factor:	0.20	Real Return on Debt:	0.046
Indirect Construction Factor:	0.20	Real Ret. on Pref.:	0.052
Engr & Home Office Fees:	0.10	Real Ret. on Equity:	0.087
Project Contingency Factor:	0.17	Debt Ratio:	0.500
Byproduct Marketing Factor:	0.10	Pref. Stock Ratio:	0.150
Average Operating Labor Rate:	19.70	Fed. & State Taxes:	0.380
Number of Shifts:	4.25	Investment Tax Credit:	0.000
Construction Interest Rate:	0.10	Prop. Taxes & Insur.:	0.020
Years of Construction:	4		

Process Contingency and Maintenance Cost Factors:

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Limestone Handling	0.000	0.030
Oxidant Feed	0.100	0.020
Gasification	0.200	0.030
Coke, Ash, & Bent. Subsystems	0.050	0.020
High Temp. Cyclones	0.050	0.030
Zinc Ferrite	0.400	0.030
Sulfuric Acid Plant	0.100	0.020
Sulfation	0.600	0.045
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.250	0.020
Heat Recovery Steam Generator	0.025	0.015
Direct Sulfur Recovery Process	0.250	0.020
Selective Catalytic Reduction	0.100	0.020
Steam Turbine	0.025	0.015
General Facilities	0.050	0.015

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

Coal Handling	1	1	24743.	1692.
Limestone Handling	0	0	0.	0.
Oxidant Feed	3	3	5284.	723.
Gasification	13	15	99270.	27147.
Coke, Ash, & Bent. Subsystems	13	15	18716.	1280.
High Temp. Cyclones	26	30	7406.	506.
Zinc Ferrite	13	30	14484.	7922.
Sulfuric Acid Plant	1	1	0.	0.
Sulfation	0	0	0.	0.
Boiler Feedwater Treatment	1	1	6727.	0.
Gas Turbine	3	3	96000.	32816.
Heat Recovery Steam Generator	3	3	33509.	1145.
Direct Sulfur Recovery Proces	2	2	14892.	5091.
Selective Catalytic Reduction	3	3	10337.	1413.
Steam Turbine	1	1	38630.	1321.
General Facilities	N/A	N/A	74000.	5059.

C. CALCULATED TOTAL CAPITAL REQUIREMENT (\$1,000)

Description	Capital Cost
Total Direct Cost	443998.
Indirect Construction Cost	88800.
Sales Tax	18204.
Engineering and Home Office Fees	55100.
Environmental Permitting	1000.
Total Indirect Costs	163104.
Total Process Contingencies	86115.
Project Contingency	121313.
Total Plant Cost	814529.
AFDC	130528.
Total Plant Investment	945058.
Preproduction (Startup) Costs	26787.
Inventory Capital	24591.
Initial Catalysts and Chemicals	41812.
Land	3203.
TOTAL CAPITAL REQUIREMENT (\$1,000)	1050901.

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

Description	Annual Cost
Operating Labor	9229844.
Maintenance Costs	17174910.
Administration and Supervision	4829942.
TOTAL FIXED OPERATING COST (\$/year)	31234697.

E. CALCULATED VARIABLE OPERATING COSTS

1. CONSUMABLES (\$/year)

Description	Assumed Unit Cost	Calc. Material Requirement	Calc. Annual Operating Cost
Sulfuric Acid:	110.00 \$/ton	2866.7 ton/yr	315340.
NaOH:	220.00 \$/ton	592.7 ton/yr	130401.
Na2 HPO4:	0.70 \$/lb	3534.5 lb/yr	2474.
Hydrazine:	3.20 \$/lb	17019.9 lb/yr	54464.
Morpholine:	1.30 \$/lb	15894.8 lb/yr	20663.
Lime:	80.00 \$/ton	699.5 ton/yr	55961.
Soda Ash:	160.00 \$/ton	771.5 ton/yr	123444.
Corrosion Inh.:	1.90 \$/lb	139553.2 lb/yr	265151.
Surfactant:	1.25 \$/lb	139553.2 lb/yr	174442.
Chlorine:	250.00 \$/ton	21.4 ton/yr	5351.
Biocide:	3.60 \$/lb	23760.8 lb/yr	85539.
DSRP Catalyst:	12.28 \$/lb	43034.4 lb/yr	528462.
SCR Catalyst:	250.00 \$/ft^3	6250.5 ft^3/yr	1562635.
Ammonia:	150.00 \$/ton	17214.3 ton/yr	2582142.
Zinc Fer Sorb:	3.00 \$/lb	3495996.1 lb/yr	10487988.
Limestone:	18.00 \$/ton	0.0 ton/yr	0.
Sul Acid. Cat.:	1.90 \$/l	35008.6 l/yr	66516.
Coke:	150.00 \$/ton	1875.1 ton/yr	281259.
Bentonite:	0.03 \$/lb	5404647.7 lb/yr	156735.
Fuel Oil:	42.00 \$/bbl	48287.7 bbl/yr	2028083.
Plant Air Ads.:	2.80 \$/lb	3621.6 lb/yr	10140.
Raw Water:	0.73 \$/Kgal	1006385.9 Kgal/yr	734662.

Waste Water:	840.00 \$/gpm ww	0.0 lb/hr	0.
LPG - Flare:	11.70 \$/bbl	4225.2 bbl/yr	49435.
TOTAL CONSUMABLES (\$/year)			19721288.

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

Coal:	1.61 \$/MMBtu	663880.3 lb/hr	77742777.
Ash Disposal:	10.00 \$/ton	836.3 ton/day	1984151.
Sulfuric Acid:	40.00 \$/ton	0.0 ton/yr	0.
Sulfur:	125.00 \$/ton	58917.5 ton/yr	6628217.

TOTAL VARIABLE OPERATING COST (\$/year) -----> 92819998.

F. CALCULATED COST OF ELECTRICITY

Power Summary (MWe)	Auxiliary Loads (MWe)
Gas Turbine Output	641.25
Steam Turbine Output	243.41
Total Auxiliary Loads	33.44
Net Electricity	851.23
Coal Handling	1.25
Limestone Hdlg	0.00
Oxidant Feed	16.39
Gasification	3.32
Zinc Ferrite	0.00
Sulfation	0.00
DSRP	1.09
Steam Cycle	5.23
SCR	0.21
General Fac	3.94

Capital Cost:	1234.58 \$/kW
Fixed Operating Cost:	36.69 \$/(kW-yr)
Incremental Variable Costs:	4.48 mills/kWh
Byproduct Credit:	1.37 mills/kWh
Fuel Cost:	16.04 mills/kWh
Variable Operating Cost:	19.15 mills/kWh
COST OF ELECTRICITY	48.01 mills/kWh
Fixed Charge Factor: 0.1034	Variable Cost Levelization Factor: 1.0000

The plant heatrate (HHV) is: 8773. BTU/kWh. Efficiency: 0.3893

G. ENVIRONMENTAL SUMMARY

INPUTS:	Coal	0.7799 lb/kWh
	Water	1.7322 lb/kWh
OUTPUTS:	Blowdown	0.1078 lb/kWh
	Ash	0.0819 lb/kWh
	Sulfuric Acid	0.0000 lb/kWh
	CO2 emissions	1.7035 lb/kWh
	CO emissions	0.0027 lb/kWh
	SO2 emissions	0.3506 lb/MMBtu
	COS emissions	0.0000 lb/MMBtu
	CH4 emissions	0.0000 lb/kWh
	H2S emissions	0.0000 lb/MMBtu
	NOx emissions	0.5407 lb/MMBtu

NOTE: NOx emissions are based on fuel bound ammonia

I. SCR SUMMARY

Reference Removal Efficiency	0.80 fraction
Actual Removal Efficiency	0.80 fraction
Ammonia Slip	10.00 ppmv
Reference NOx Concentration	42.00 ppmv
Minimum Relative Cat. Activ.	994.34 ratio
Activity Time Constant	12638.01
Replacement Interval	11388.00 hrs
Reference Temperature	717.00 F
Actual Temperature	717.00 F
Dilution Steam/NH3 Ratio	19.00 lbmole/lbmole
NH3/NOx Ratio	0.81 lbmole/lbmole
Fraction NOx as NO	0.95
Reference Replacement Interv.	17082.00 hrs
SO2 Oxidation Rate	0.01 fraction
SCR Pressure Drop	0.14 psi
Reference Space Velocity	10000.00 1/hr
Calculated Space Velocity	3966.51 1/hr
No. Initial Catalyst Layers	3
No. Total Catalyst Layers	3
No. Layers Replaced at a time	1
Factor, efficiency	1.00
Factor, end mole ratio	0.3734

Factor, activity	1.0622
Factor, temperature	1.0000
Initial Catalyst Volume	37503.25 ft ³
Total Catalyst Volume	37503.25 ft ³
Reference Cat. Volume	14875.69 ft ³
Annual Layers Replaced	0.50
Annual Volume Replaced	6250.54 ft ³ /yr
Catalyst Life, yrs	6.00 yrs
Direct Capital Cost	10336.52
Ammonia Flow, tons/yr	355.05

J. DSRP SUMMARY

Stage 1 Space Velocity	5000.00 scc/(cc-h)
Stage 2 Space Velocity	5000.00 scc/(cc-h)
DSRP Work, MW	1.0946
Number of total trains	2
Number of oper. trains	2
Rxr 1 Inlet Volume Flow Rate	2303599.8741 ft ³ /hr
Rxr 1 Inlet Temperature	1160.20 F
Rxr 1 Inlet Pressure	284.72 psia
Cost Index	351.50
Pressure Correction Factor 1	0.84 \$1,000
Rxr 2 Inlet Volume Flow Rate	0.0000 ft ³ /hr
Rxr 2 Inlet Temperature	350.00 F
Rxr 2 Inlet Pressure	249.72 psia
Pressure Correction Factor 2	0.00
Heat Duty: Precooler	0.291470E+08 BTU/hr
Heat Duty: Condenser 1	0.644246E+09 BTU/hr
Heat Duty: Reheater	-.132215E-02 BTU/hr
Heat Duty: Condenser 2	-.132215E-02 BTU/hr
Direct Cost, gas handling 1	14526.79 \$1,000
Direct Cost, gas handling 2	0.00 \$1,000
Direct Cost, sulfur hand, 1	365.11 \$1,000
Direct Cost, sulfur hand, 2	0.00 \$1,000
Direct Cost, total	14891.90 \$1,000
Total Initial Catalyst	129103.21 lb
Annual Catalyst	43034.40 lb/yr
Sulfur Byproduct	58917.49 tons/year

Computer Model Output for Two-Stage DSRP Analysis: Lurgi-based IGCC with Two-Stage DSRP

Air-Blown Fixed-Bed Gasifier IGCC w/SCR: Deterministic Analysis
SYSTEM SUMMARY

*** GASIFIER CONDITIONS ***

COAL FLOW RATE: 0.653009E+06 LB/HR
SORBENT FLOW RATE: 0.568699E+00 LB/HR
AIR FLOW RATE: 0.160327E+07 LB/HR
REQUIRED STEAM: 0.801634E+06 LB/HR
SUPPLIED STEAM: 0.998983E+03 LB/HR
SO2 FLOW RATE: 0.999983E+06 LB/HR
ASH FLOW RATE: 57172.308874 LB/HR
GASIFIER PRESSURE: 1307.5 PSIA
GASIFIER TEMPERATURE: 1100.0 F

ZINC FERRITE SULFUR REMOVAL PROCESS
RESULTS GENERATED ON JUL 6, 1995 AT 21:43

** SO2 BALANCE (LB/HR) **

H2S IN RAWGAS 22905.91
H2S IN FUELGAS 48.10
SO2 IN SO2GAS 42967.71
PPMV OF H2S IN FUELGAS 11.

** FLOW RATES (LB/HR) **

RAWGAS 0.30D+07 MW = 23.11
FUELGAS 0.30D+07 MW = 23.08
SO2GAS 637315.78 MW = 20.35
STEAM REQUIRED TO MAINTAIN
CO/CO+CO2 RATIO (SHFT-STM) 1.00
DILUENT REQUIRED TO MAINTAIN
REGENERATOR TEMPERATURE (GEN-STM) 475122.18
AIR FLOW TO REGENERATOR (REGENAIR) 154995.98

** GENERAL **

CO/(CO+CO2) RATIO: 0.281
SULFUR CAPTURE EFFICIENCY: 0.998
TEMPERATURE OF THE REGENERATED GAS: 1450.0
VOLUME PERCENT AIR IN REGENERATION GAS: 16.761

DIRECT SULFUR RECOVERY PROCESS
RESULTS GENERATED ON JUL 6, 1995 AT 21:43

** DSRP MASS BALANCE (LB/HR) **

Description	DSRP GAS STREAMS		
	Mass Flow (lb/hr)	Temp. (deg F)	Pres. (psia)
HGD OFF-GAS	637315.78	1233.	305.50
COAL GAS	2982200.85	1206.	286.72
SLIP STREAM	133882.63	1206.	286.72
COOLER INLET	771198.41	1228.	286.72
SHIFT INLET	771198.41	1148.	285.72
REACTOR 1 INLET	771198.24	1161.	284.72
CONDENSER 1 INLET	752777.28	1292.	259.72
CONDENSER 1 OUTLET	752777.28	350.	249.72
REHEATER INLET	752777.28	350.	249.72
REACTOR 2 INLET	752777.28	550.	248.72
CONDENSER 2 INLET	749874.02	550.	223.72
CONDENSER 2 OUTLET	749874.02	350.	213.72

DSRP TAILGAS	749874.02	350.	213.72
COMPRESSED TAILGAS	749874.02	372.	286.72
FUEL TO GAS TURB.	3601418.30	852.	286.72

** DSRP SULFUR MASS BALANCE (LBMOLE/HR) **

INLET SULFUR: SO2 IN OFF-GAS 670.71
SO2 IN SLIP STREAM 0.00
H2S IN SLIP STREAM 0.06
COS IN SLIP STREAM 0.00
TOTAL INLET SULFUR----> 670.77

OUTLET SULFUR: CONDENSER NO. 1 574.51
CONDENSER NO. 2 90.55
SO2 IN TAILGAS 3.35
H2S IN TAILGAS 2.37
TOTAL OUTLET SULFUR----> 670.77

DSRP SULFUR CAPTURE EFFICIENCY: 99.15 PERCENT

Air-Blown Fixed-Bed Gasifier IGCC w/SCR: Deterministic Analysis
SYSTEM SUMMARY

*** MS7000F GAS TURBINE CONDITIONS ***

FUEL FLOW RATE: 0.360136E+07 LB/HR
AIR FLOW RATE: 0.878523E+07 LB/HR
AIR DIVERTED TO GASIFIER: 0.160327E+07 LB/HR
AIR DIVERTED TO ZINC FER: 0.154993E+06 LB/HR
FUEL LHV (WET BASIS): 1391.2 BTU/LB, 82.9 BTU/SCF
FUEL HHV (WET BASIS): 1562.0 BTU/LB, 92.6 BTU/SCF
FIRING TEMPERATURE: 2350.0 F
COMBUSTOR EXIT TEMPERATURE: 2422.1 F
TURBINE EXHAUST TEMPERATURE: 1174.1 F
THERMAL EFFICIENCY (LHV): 0.4326
GENERATOR EFFICIENCY: 0.9850
GASIFIER AIR INTERCOOLER DUTY: 0.000000E+00 BTU/HR

Ratio of Inlet Air to Exhaust Gas: 0.824
Ratio of Boost Air to Inlet Air: 0.200
Ratio of Gasif Air to Inlet Air: 0.182
Ratio of Regen Air to Inlet Air: 0.018
Ratio of Fuel Gas to Inlet Air: 0.410
CO2 Emissions: 194.2 LB/MMBTU
CO Emissions: 0.30849 LB/MMBTU
CO Emissions: 206.56 ppmv
SO2 Emissions: 0.06098 LB/MMBTU
NOx Emissions: 0.54114 LB/MMBTU
NOx Emissions: 220.56 ppmv
Oxygen in Exhaust: 6.6749 percent
Vapor in Exhaust: 25.6962 percent
CO2 in Exhaust: 8.2744 percent

*** STEAM TURBINE CONDITIONS ***

STEAM DIVERTED TO GASIFIER: 0.801634E+06 LB/HR
STEAM DIVERTED TO ZINC FER: 0.475114E+06 LB/HR
STEAM TURBINE FLOW RATE: 0.224309E+07 LB/HR
SUPERHEATED STEAM TEMPERATURE: 1000.0 F
EXPANDED STEAM QUALITY: 0.9000
ISENTROPIC EFFICIENCY: 0.9227
GENERATOR EFFICIENCY: 0.9850

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

PLANT SECTION	FLOWSHEET PERFORMANCE PARAMETERS	VALUE UNITS
-----	-----	-----

Coal Feed	Mass flow of coal to gasifier	653019.87 lb/hr
Limestone Feed	Mass flow of limestone (mixed)	0.00 lb/hr
	Mass flow of limestone (cisolid)	0.57 lb/hr
Oxidant Feed	Work to boost air compressor	0.161255E+05 Watts
	First precool inlet air temp.	750.33 F
	First precool outlet air temp.	679.51 F
	First precool inlet BFW temp.	505.47 F
	First precool outlet BFW temp.	605.29 F
	Heat transfer in first precool	0.321807E+08 BTU/hr
	Heat trans. to BFW from HRSG	0.592591E+09 BTU/hr
	Heat trans. to BFW from Regen.	0.000000E+00 BTU/hr
	Heat leaving from economizer	0.325341E+09 BTU/hr
Gasification	Gasifier output syngas pres.	307.50 psia
	Gasifier output syngas temp.	1100.00 F
	Gasifier output syngas density	0.183078E-01 lbmole/ft3
	Gasifier recycle fines flow rate	21712.91 lb/hr
Zinc Ferrite	Zinc Ferrite inlet syngas flow	129366.78 lbmole/hr
	Off-gas to sulfuric acid plant	31318.26 lbmole/hr
	SO2 to sulfuric acid plant	670.71 lbmole/hr
	Off-gas temp. to sul. acid plant	1233.29 lbmole/hr
	Zinc ferrite inlet H2S flow	672.12 lbmole/hr
	Zinc ferrite inlet COS flow	0.00 lbmole/hr
Sulfation	Gasifier ash removal	57173.27 lb/hr
	Gasifier ash removal	10239.03 lb/hr
	Gasifier fines removal	285.70 lb/hr
	Gasifier fines removal	857.09 lb/hr
BFW Treating	Raw water to power plant	1449628.01 lb/hr
	Steam turbine condensate	853466.30 lb/hr
Gas Turbine	Gas turbine inlet air flow	8785379.11 lb/hr
Steam Cycle	HRSG outlet HP steam pres.	1465.00 psia
	HRSG outlet HP steam flow	2243131.11 lb/hr
	Gas turbine net shaft work	-647921E+09 Watts
	Steam turbine net shaft work	-238149E+09 Watts
	Steam cycle auxiliary power	0.513474E+07 Watts
	Acid gas auxiliary power	0.000000E+00 Watts
Miscellaneous	Moisture in coal feed	0.12 wt %
	Ash in coal	0.10 wt %
	Temp. of ambient air	59.00 F
	Heating value of coal	12774.00 BTU/lb
DSRP	DSRP Sulfur Product	21324.21 lb/hr
	DSRP Gas Cooler Inlet	37119.76 lbmole/hr
	DSRP Reactor 1 Inlet	37119.76 lbmole/hr
	DSRP Condenser 1 Inlet	36482.59 lbmole/hr
	DSRP Reheater Inlet	36482.59 lbmole/hr
	DSRP Reactor 2 Inlet	36482.59 lbmole/hr
	DSRP Condenser 2 Inlet	36452.40 lbmole/hr
	DSRP Compressor Work	0.298293E+07 Watts
	DSRP Reactor 2 Inlet Pressure	284.72 psia
	DSRP Reactor 2 Inlet Temperature	1161.29 F
SCR	SCR Inlet Gas Flow	391800.48 lbmole/hr
	SCR Inlet Gas Temperature	717.00 lbmole/hr
	SCR Inlet Gas Pressure	15.33 lbmole/hr
	SCR Inlet NO loading	410.47 lbmole/hr
	SCR Inlet NO2 loading	21.60 lbmole/hr
	SCR Ammonia Requirement	349.58 lbmole/hr
	SCR Dilution Steam Requirement	6641.96 lbmole/hr
Environmental	High Pressure Blowdown	45778.19 lbmole/hr
	Low Pressure Blowdown	45778.19 lbmole/hr
	CO2 from gas turbine	32419.12 lbmole/hr
	CO from gas turbine	80.93 lbmole/hr
	SO2 from gas turbine	7.00 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	461.56 lbmole/hr
	NO from gas turbine	82.09 lbmole/hr
	NO2 from gas turbine	4.32 lbmole/hr
	Actual coal heating value	11248.73 BTU/lb
COST VAR WARNING	Variable MCFGI value of	653019.868
	in DCSS above the upper limit of	433000.000
COST VAR WARNING	Variable VSNZFI value of	9059.221

	in DCCY	above the upper limit of	6000.000
COST VAR WARNING	Variable MRW	value of	1449628.005
	in DCFB	above the upper limit of	614000.000
COST VAR WARNING	Variable MSH/N	value of	747710.372
	in DCHR	above the upper limit of	640000.000
COST VAR WARNING	Variable MPW	value of	2303094.303
	in MSABFP	above the upper limit of	2200000.000

COST SUMMARY
Air Blown Lurgi-Based IGCC System with Hot Gas Cleanup
Cost Model Version 1.63 Rev. Date: June 24, 1995

A. KEY INPUT ASSUMPTIONS

Performance Assumptions:

Gasifier Availability:	0.87	Sorb. Sulfur Loading:	0.168
Max. Desulf. Vessel Diameter:	12.50	Superficial Velocity:	2.00
L/D Ratio:	3.00	Absorption Cycle Time:	30.0
Maximum Space Velocity (1/hr):	2000.	Sorbent Bulk Density:	82.0
Sorb. Attrition Rate/80 cycles:	0.800	Generator Efficiency:	0.9850
Gasifier Coal Throughput:	1.01		

Economic Assumptions:

Cost Year:	January 1989	Inflation Rate:	0.000
Plant Cost Index:	351.5	Real Escalation Rate:	0.000
Chemicals Cost Index:	411.3	Plant Booklife:	30
Plant Capacity Factor:	0.65	Sales Tax Rate:	0.05
General Facilities Factor:	0.20	Real Return on Debt:	0.046
Indirect Construction Factor:	0.20	Real Ret. on Pref.:	0.052
Engr & Home Office Fees:	0.10	Real Ret. on Equity:	0.087
Project Contingency Factor:	0.17	Debt Ratio:	0.500
Byproduct Marketing Factor:	0.10	Pref. Stock Ratio:	0.150
Average Operating Labor Rate:	19.70	Fed. & State Taxes:	0.380
Number of Shifts:	4.25	Investment Tax Credit:	0.000
Construction Interest Rate:	0.10	Prop. Taxes & Insur.:	0.020
Years of Construction:	4		

Process Contingency and Maintenance Cost Factors:

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Limestone Handling	0.000	0.030
Oxidant Feed	0.100	0.020
Gasification	0.200	0.030
Coke, Ash, & Bent. Subsystems	0.050	0.020
High Temp. Cyclones	0.050	0.030
Zinc Ferrite	0.400	0.030
Sulfuric Acid Plant	0.100	0.020
Sulfation	0.600	0.045
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.250	0.020
Heat Recovery Steam Generator	0.025	0.015
Direct Sulfur Recovery Proces	0.250	0.020
Selective Catalytic Reduction	0.100	0.020
Steam Turbine	0.025	0.015
General Facilities	0.050	0.015

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

Plant Section	Number of Units	Direct Operating Total	Capital Cost	Process Contingency
Coal Handling	1	1	24522.	1676.
Limestone Handling	0	0	0.	0.
Oxidant Feed	3	3	5240.	717.
Gasification	13	15	99270.	27147.
Coke, Ash, & Bent. Subsystems	13	15	18391.	1257.
High Temp. Cyclones	26	30	7353.	503.

Zinc Ferrite	13	30	14364.	7856.
Sulfuric Acid Plant	1	1	0.	0.
Sulfation	0	0	0.	0.
Boiler Feedwater Treatment	1	1	6614.	0.
Gas Turbine	3	3	96000.	32816.
Heat Recovery Steam Generator	3	3	33484.	1145.
Direct Sulfur Recovery Proces	2	2	23500.	8033.
Selective Catalytic Reduction	3	3	10304.	1409.
Steam Turbine	1	1	37227.	1273.
General Facilities	N/A	N/A	75254.	5145.

C. CALCULATED TOTAL CAPITAL REQUIREMENT (\$1,000)

Description	Capital Cost
Total Direct Cost	451523.
Indirect Construction Cost	90305.
Sales Tax	18512.
Engineering and Home Office Fees	56034.
Environmental Permitting	1000.
Total Indirect Costs	165851.
Total Process Contingencies	88975.
Project Contingency	123611.
Total Plant Cost	829960.
AFDC	133001.
Total Plant Investment	962961.
Preproduction (Startup) Costs	27124.
Inventory Capital	24323.
Initial Catalysts and Chemicals	42773.
Land	3140.
TOTAL CAPITAL REQUIREMENT (\$1,000)	1069951.

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

Description	Annual Cost
Operating Labor	9229844.
Maintenance Costs	17137179.
Administration and Supervision	4825415.
TOTAL FIXED OPERATING COST (\$/year)	31192437.

E. CALCULATED VARIABLE OPERATING COSTS

1. CONSUMABLES (\$/year)

Description	Assumed Unit Cost	Calc. Material Requirement	Calc. Annual Operating Cost
Sulfuric Acid:	110.00 \$/ton	2818.9 ton/yr	310082.
NaOH:	220.00 \$/ton	581.6 ton/yr	127953.
Na2 HPO4:	0.70 \$/lb	3476.3 lb/yr	2433.
Hydrazine:	3.20 \$/lb	16739.1 lb/yr	53565.
Morpholine:	1.30 \$/lb	15631.8 lb/yr	20321.
Lime:	80.00 \$/ton	688.0 ton/yr	55036.
Soda Ash:	160.00 \$/ton	759.1 ton/yr	121451.
Corrosion Inh.:	1.90 \$/lb	137224.3 lb/yr	260726.
Surfactant:	1.25 \$/lb	137224.3 lb/yr	171530.
Chlorine:	250.00 \$/ton	21.1 ton/yr	5275.
Biocide:	3.60 \$/lb	23427.6 lb/yr	84339.
DSRP Catalyst:	12.28 \$/lb	83963.5 lb/yr	1031072.
SCR Catalyst:	250.00 \$/ft^3	6219.9 ft^3/yr	1554977.
Ammonia:	150.00 \$/ton	16949.0 ton/yr	2542352.
Zinc Fer Sorb:	3.00 \$/lb	3438812.7 lb/yr	10316438.
Limestone:	18.00 \$/ton	0.0 ton/yr	0.
Sul Acid. Cat.:	1.90 \$/l	34450.1 l/yr	65455.
Coke:	150.00 \$/ton	1845.8 ton/yr	276875.
Bentonite:	0.03 \$/lb	5316232.8 lb/yr	154171.
Fuel Oil:	42.00 \$/bbl	47535.0 bbl/yr	1996470.
Plant Air Ads.:	2.80 \$/lb	3565.1 lb/yr	9982.
Raw Water:	0.73 \$/Kgal	989443.6 Kgal/yr	722294.
Waste Water:	840.00 \$/gpm ww	0.0 lb/hr	0.
LPG - Flare:	11.70 \$/bbl	4159.3 bbl/yr	48664.
TOTAL CONSUMABLES (\$/year)			19931464.

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

Coal:	1.61 \$/MMBtu	653019.9 lb/hr	76470978.
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Ash Disposal:	10.00 \$/ton	822.7 ton/day	1951763.
Sulfuric Acid:	40.00 \$/ton	0.0 ton/yr	0.
Sulfur:	125.00 \$/ton	60710.0 ton/yr	6829877.

TOTAL VARIABLE OPERATING COST (\$/year) -----> 91524328.

F. CALCULATED COST OF ELECTRICITY

Power Summary (MWe)	Auxiliary Loads (MWe)
Gas Turbine Output	638.20
Coal Handling	1.23
Steam Turbine Output	234.58
Limestone Hdlg	0.00
Total Auxiliary Loads	34.82
Oxidant Feed	16.13
Gasification	3.27
Net Electricity	837.96
Zinc Ferrite	0.00
General Fac	3.90

Capital Cost:	1276.86 \$/kW
Fixed Operating Cost:	37.22 \$/(kW-yr)
Incremental Variable Costs:	4.59 mills/kWh
Byproduct Credit:	1.43 mills/kWh
Fuel Cost:	16.03 mills/kWh
Variable Operating Cost:	19.18 mills/kWh

COST OF ELECTRICITY -----> 48.91 mills/kWh
Fixed Charge Factor: 0.1034 Variable Cost Levelization Factor: 1.0000

The plant heatrate (HHV) is: 8766. BTU/kWh. Efficiency: 0.3896

G. ENVIRONMENTAL SUMMARY

INPUTS:	Coal	0.7793 lb/kWh
	Water	1.7300 lb/kWh
OUTPUTS:	Blowdown	0.1093 lb/kWh
	Ash	0.0818 lb/kWh
	Sulfuric Acid	0.0000 lb/kWh
	CO2 emissions	1.7023 lb/kWh
	CO emissions	0.0027 lb/kWh
	SO2 emissions	0.0902 lb/MMBtu
	COS emissions	0.0000 lb/MMBtu
	CH4 emissions	0.0000 lb/kWh
	H2S emissions	0.0000 lb/MMBtu
	NOx emissions	0.5411 lb/MMBtu

NOTE: NOx emissions are based on fuel bound ammonia

I. SCR SUMMARY

Reference Removal Efficiency	0.80 fraction
Actual Removal Efficiency	0.80 fraction
Ammonia Slip	10.00 ppmv
Reference NOx Concentration	42.00 ppmv
Minimum Relative Cat. Activ.	978.07 ratio
Activity Time Constant	12638.01
Replacement Interval	11388.00 hrs
Reference Temperature	717.00 F
Actual Temperature	717.00 F
Dilution Steam/NH3 Ratio	19.00 lbmole/lbmole
NH3/NOx Ratio	0.81 lbmole/lbmole
Fraction NOx as NO	0.95
Reference Replacement Interv.	17082.00 hrs
SO2 Oxidation Rate	0.01 fraction
SCR Pressure Drop	0.14 psi
Reference Space Velocity	10000.00 1/hr
Calculated Space Velocity	3984.88 1/hr
No. Initial Catalyst Layers	3
No. Total Catalyst Layers	3
No. Layers Replaced at a time	1
Factor, efficiency	1.00
Factor, end mole ratio	0.3752
Factor, activity	1.0622
Factor, temperature	1.0000
Initial Catalyst Volume	37319.45 ft^3
Total Catalyst Volume	37319.45 ft^3
Reference Cat. Volume	14871.36 ft^3
Annual Layers Replaced	0.50

Annual Volume Replaced	6219.91 ft ³ /yr
Catalyst Life, yrs	6.00 yrs
Direct Capital Cost	10304.22
Ammonia Flow, tons/yr	349.58

J. DSRP SUMMARY

Stage 1 Space Velocity	5000.00 scc/(cc-h)
Stage 2 Space Velocity	5000.00 scc/(cc-h)
DSRP Work, MW	2.9829
Number of total trains	2
Number of oper. trains	2
Rxr 1 Inlet Volume Flow Rate	2268228.8773 ft ³ /hr
Rxr 1 Inlet Temperature	1161.29 F
Rxr 1 Inlet Pressure	284.72 psia
Cost Index	351.50
Pressure Correction Factor 1	0.84 \$1,000
Rxr 2 Inlet Volume Flow Rate	1589777.9452 ft ³ /hr
Rxr 2 Inlet Temperature	550.00 F
Rxr 2 Inlet Pressure	248.72 psia
Pressure Correction Factor 2	0.80
Heat Duty: Precooler	0.282255E+08 BTU/hr
Heat Duty: Condenser 1	0.631346E+09 BTU/hr
Heat Duty: Reheater	-1.372399E+09 BTU/hr
Heat Duty: Condenser 2	0.308328E+09 BTU/hr
Direct Cost, gas handling 1	14331.69 \$1,000
Direct Cost, gas handling 2	8236.84 \$1,000
Direct Cost, sulfur hand, 1	372.52 \$1,000
Direct Cost, sulfur hand, 2	558.77 \$1,000
Direct Cost, total	23499.82 \$1,000
Total Initial Catalyst	251890.49 lb
Annual Catalyst	83963.50 lb/yr
Sulfur Byproduct	60710.01 tons/year

This cover stock is 30% post-consumer waste
and 30% pre-consumer waste, and is recyclable.