**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

Final Report

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ABSTRACT

This document assesses the characteristics of advanced gasification-based power plants with respect to capital costs, electricity costs, thermal performance, and environmental performance. The study was sponsored by the U.S. Department of Energy (DOE) Morgantown Energy Technology Center (METC) and Southern Company Services, Inc., and was designed to be as specific as possible for The Southern Company, one of the nation's largest electric utility systems. Georgia Power Company's Plant Wansley, which currently has two 865-MW coal-fired steam units and is a candidate for future expansion, was selected as the site for the study.

The study team was led by Southern Company Services. Other participants were The M.W. Kellogg Company, Tennessee Technological University Center for Electric Power, and General Electric Company.

The two gasification processes used for the study were the KRW fluidized-bed gasifier and the transport gasifier. The transport technology is a high-velocity circulating fluidized-bed reactor currently under development by The M.W. Kellogg Company. The combustion turbines are the GE MS7001F gas turbine for IGCC plants and the GE LM8000 intercooled steam-injected gas turbine for ISTIG plants.

Seven design concepts or cases were identified for the study. For six cases, a process design was developed; major equipment items were identified; estimates of capital cost, operation and maintenance cost, and cost of electricity were developed; reliability was predicted; and development issues were identified. The seventh case, consisting of an IGCC unit combined with mild gasification, was limited to a feasibility assessment, only. All seven cases used air as the gasification oxidant and employed hot gas cleanup. Air-blown IGCC with hot gas cleanup was compared with oxygen-blown IGCC with cold gas cleanup. Detailed Second-Law and thermoeconomic analysis was performed; design changes and further optimization were recommended.

Possible implementation for The Southern Company is discussed in terms of advantages and disadvantages relative to other baseload options. Development issues, engineering and demonstration needs, and commercialization plans are addressed.

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EXECUTIVE SUMMARY

Integrated gasification combined-cycle (IGCC) electric generating plants are currently under development to provide a clean, highly efficient coal fueled generating option. IGCC offers baseload operation by coupling a coal-gasifier with a combined-cycle power generating block. The gasifier system converts coal into a clean combustible gas that fuels a combustion turbine. The combustion turbine exhaust heat is used to produce steam, which drives a turbine generator. The primary area of integration involves the recovery and transfer of heat from the gasifier syngas coolers to the combined-cycle block. In addition to the heat recovery system, the gasifier section also contains a raw gas cleanup section that includes particulate and sulfur removal systems. Another gasification-based power plant under consideration relies on the Intercooled Steam-Injected Gas Turbine (ISTIG) concept. This design eliminates the steam turbine system by reinjecting all steam produced into the gas turbine.

Through a joint site-specific study with the U.S. Department of Energy (DOE) Morgantown Energy Technology Center (METC), Southern Company Services, Inc., (SCS) conducted an evaluation to determine the characteristics of gasification power plants with respect to capital costs, electricity costs, thermal performance, and environmental performance. This evaluation was conducted to be as specific as possible to The Southern Company. The Southern Company is one of the nation's largest investor-owned electric utility systems, supplying energy through five operating electric utilities to approximately 10 million people across the Southeastern United States. The present generation expansion plan for The Southern Company estimates that the next baseload generation will be required in about the year 2006. If IGCC technology has been developed to commercial readiness, these plants could be installed as an alternative to baseload pulverized-coal-fired units currently called for in the expansion plan.

The study team was led by SCS, the engineering and research subsidiary of The Southern Company. Other participants were The M. W. Kellogg Company (MWK); Tennessee Technological University (TTU) Center for Electric Power; and General Electric Company (GE). SCS Research & Environmental Affairs was responsible for overall project management, environmental assessment, plant integration, and overall cost estimates. SCS engineering departments were responsible for the design and costing of the power island, the materials handling systems, and the balance of plant, as well as reliability-availability-maintainability (RAM) analysis and system analysis. MWK was responsible for the gasifier island designs and costs. TTU performed Second Law and thermoeconomic analyses, and GE provided combustion turbine performance and cost data.

The major objectives of the joint SCS/DOE study of air-blown gasification power plants with hot gas cleanup were to:

- Evaluate various power plant configurations to determine if an air-blown gasification-based power plant with hot gas cleanup can compete against pulverized coal with flue gas desulfurization for baseload expansion.
- Determine if air-blown gasification with hot gas cleanup is more cost effective than oxygen-blown IGCC with cold gas cleanup.
- Perform Second Law and thermoeconomic analyses of air-blown IGCC with hot gas cleanup and oxygen-blown IGCC with cold gas cleanup to understand the underlying performance differences and identify design improvements.
- Compare the cost, performance, and reliability of industrial turbine combined-cycle versus ISTIG power island configurations.
- Compare the cost, performance, and reliability of large (400 MW) versus small (100 to 200 MW) gasification power plants.
- Compare the cost, performance, and reliability of air-blown gasification power plants using fluidized-bed reactors to air-blown IGCC using transport reactors.

The two gasification processes selected for the study were the KRW fluidized-bed gasifier and the transport gasifier. The transport technology is a high-velocity circulating fluidized-bed reactor currently under development by MWK. The two combustion turbines selected for the study were a GE MS7001F gas turbine for the IGCC plants and a GE LM8000 intercooled steam-injected gas turbine for the ISTIG designs.

Seven design concepts were selected to accomplish the overall objectives of the study. For each case, a process design was developed; major equipment items were identified; estimates of capital cost, O&M cost, and cost of electricity were developed; reliability was predicted; and development issues were identified. All the cases used air as the gasification oxidant and employed hot gas cleanup. The design fuel was an Illinois No. 6 bituminous coal.

PROJECT DESCRIPTION

The study was based on the Plant Wansley site of Georgia Power Company, which is one of several candidate sites for baseload expansion in The Southern Company. The site is approximately 40 miles southwest of Atlanta, Georgia, and is located on the Chattahoochee River. Plant Wansley currently has two operating 865-MW coal-fired steam units with the major infrastructure required for a four-unit plant. By locating the IGCC plant at an existing electric utility site originally planned for expansion, it is possible to reduce the capital cost by sharing facilities such as land, site development, water supply, coal receiving, ash disposal, security facilities, and some shops and office space.

originally planned for expansion, it is possible to reduce the capital cost by sharing facilities such as land, site development, water supply, coal receiving, ash disposal, security facilities, and some shops and office space.

The study criteria were selected to match the needs of The Southern Company. Phased or staged construction was not considered an advantage in The Southern Company because projected expansion plans show the need for combustion turbines, combined-cycle units, and large coal-fired units, if units were converted to IGCC, additional combustion turbine and combined-cycle units would be required. Smaller utilities may find advantages with staged construction.

Because New Source Performance Standards (NSPS) do not exist for IGCC at this time, air emission limits for coal-fired plants were used as a guideline. The plants specified in the study also conform to the National Ambient Air Quality Standards (NAAQS) for NO_x , SO_2 , and particulates. Mechanical draft cooling towers are used and no hot water is assumed to be discharged.

PLANT DESIGNS

Table 1 compares the primary parameters and system components for the seven study cases. The IGCC system configurations for Case 1, 3, 5, and 6 are similar in that all four cases employ the same combined-cycle power island configuration. The four cases differ in either plant size or the design of the gasification island. The differences in the gas island design are in the desulfurization method or the type of gasifier. KRW gasification is a fluidized-bed system that has been developed through extensive pilot-scale testing. The Kellogg transport reactor is in the initial stages of development but appears to have significant advantages over other fluidized-bed systems.

Cases 2 and 4 evaluate a power cycle referred to as the Intercooled Steam-Injected Gas Turbine (ISTIG) design. The ISTIG takes advantage of the higher efficiencies and reliability of gas turbines used in modern jet aircraft and eliminates the steam bottoming cycle from the gas turbine power plant in order to reduce capital cost, simplify the design, and produce a highly modular plant. The ISTIG considered in this project is a study concept which has not been configured in hardware. Commercialization would require a substantial development effort, consisting primarily of component addition and packaging to convert an aircraft engine design to a stationary power plant.

Case 7 involves a transport pyrolyzer for mild gasification to produce char feed for the transport gasifier and a liquid product for peaking turbine fuel. The yield and quality of coal liquids depends on the coal properties and the conditions of the pyrolysis.

Table 1

Primary System Components and Parameters for
the Study Cases

CASE	GASIFIER TYPE	NOMINAL PLANT SIZE(MW)	CYCLE	SULFUR CLEANUP		FEATURES
				IN-BED	ZINC FERRITE	
1	KRW	400	IGCC ¹	YES	POLISHING	
2	KRW	100	ISTIG ²	YES	POLISHING	LM8000 CT ³
3	KRW	200	IGCC	YES	POLISHING	
4	KRW	400	ISTIG	YES	POLISHING	LM8000 CT
5	KRW	400	IGCC	NO	BULK	H ₂ SO ₄ PLANT
6	TRANSPORT	400	IGCC	YES	NONE	HYBRID
7	TRANSPORT	400	IGCC	YES	NONE	MILD GASIFICATION

1. Integrated gasification combined-cycle
2. Integrated steam injected gas turbine
3. Combustion turbine

PLANT PERFORMANCE AND COST RESULTS

Table 2 is a summary of the features and performance of the cases addressed in the study. Also included for comparison is an updated oxygen-blown case with conventional cold gas cleanup. The oxygen-blown case is based on a prior site-specific study conducted by SCS with EPRI co-funding (EPRI GS-6876, July 1990). Among the DOE cases, the best performance was achieved by the transport IGCC (Case 6), which uses a cycle that includes both gasification and pressurized combustion, called a hybrid cycle. The heat rate of 8153 Btu/kWh (HHV) for this case would be better than virtually any other coal-fired plant (generating only electric power) currently in existence or planned for the near future. This performance compares to heat rates ranging from 8595 to 9357 Btu/kWh for the other IGCC cases, and 9039 Btu/kWh for the oxygen-blown design. The substantial performance advantage of this case over the KRW-based designs is attributable in large measure to the high cold gas efficiency predicted for the transport gasifier (82.1 percent vs. 70 percent), which is due in turn to the good gasification kinetics and the low reactor temperature. While the transport

Table 2

Summary of Gasification Power Plant Performance

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	OXYGEN BLOWN *
NOMINAL PLANT SIZE (MW)	400	100	200	400	400	400	400
GASIFIER TYPE	KRW	KRW	KRW	KRW	KRW	TRANS.	KRW
CYCLE TYPE	IGCC	ISTIG	IGCC	ISTIG	IGCC	IGCC	IGCC
BULK SULFUR REMOVAL	IN-BED	IN-BED	IN-BED	IN-BED	ZnFe	IN-BED	SELEXOL
<u>GASIFIER FEEDS:</u>							
COAL (T/DAY) (as prepared)	3,792	892	1,896	3,567	3,783	3,213	3,845
LIMESTONE (T/DAY)	1,053	248	526	990	---	431	---
AIR (T/DAY)	12,888	2,984	6,444	11,936	13,274	11,334	---
OXYGEN (T/DAY)	---	---	---	---	---	---	2,347
<u>GASIFICATION ISLAND PRODUCTS:</u>							
FUEL GAS (LHV, MMBTU/DAY)	63,336	15,096	31,668	60,384	62,184	59,757	73,653
STEAM TO POWER ISLAND (MLB/DAY)							
1600 psia (sat.) (1500 psia for Oxy. Blown)	12,447	---	6,224	---	12,824	6,160	10,555
850 psia (sat.)	---	949	---	3,795	---	---	---
SOLID WASTE (T/DAY)	1,231	289	615	1,158	540	797	492
SULFURIC ACID (T/DAY)	---	---	---	---	280	---	---
SULFUR (T/DAY)	---	---	---	---	---	---	78
<u>GASIFICATION ISLAND PERFORMANCE:</u>							
COLD GAS EFFICIENCY - LHV (%)	69.7	70.6	69.7	70.6	68.6	77.6	82.1
CARBON CONVERSION (%)							
GASIFIER ONLY	96.5	97.1	96.5	97.1	95.2	98.2	95.8
GASIFICATION ISLAND	99.9	99.9	99.9	99.9	95.2	99.9	95.8
SULFUR REMOVAL (%)							
GASIFIER ONLY	86.5	86.5	86.5	86.5	----	----	---
GASIFICATION ISLAND	99.4	99.4	99.4	99.4	99.4	99.2	96.4
<u>R-A-M ANALYSIS:</u>							
TOTAL PLANT EQ. AVAIL. FACTOR, % ***	84.4	84.2	83.0	86.1	84.0	85.3	85.2
GAS ISLAND EQ. AVAIL. FACTOR, % ****	94.8	93.4	93.5	95.2	94.4	96.0	N/A
<u>OVERALL PLANT PERFORMANCE:</u>							
COAL ENERGY INPUT (MMBTU/HR)	3,940	927	1,970	3,706	3,932	3,339	3,737
GROSS POWER OUTPUT (MW)							
COMBUSTION TURBINE(S)	311.6	104.7	155.8	419.0	305.6	263.2	298.8
STEAM TURBINE(S)	182.2	---	94.0	---	153.4	178.4	159.4
TOTAL POWER	493.8	104.7	249.8	419.0	459.0	441.6	458.2
STATION SERVICE (MW)	35.4	5.2	18.2	19.9	38.8	32.1	44.7
NET POWER OUTPUT (MW)	458.4	99.5	231.7	399.1	420.2	409.5	413.5
NET HEAT RATE - HHV (BTU/KWH)	8595	9310	8504	9287	9357	8153	9039
THERMAL EFFICIENCY - HHV (%)	39.7	36.6	40.1	36.7	36.5	41.8	37.7

* Updated design based on SCS study of KRW-based IGCC, Electric Power Research Institute Research Project 2773-5

** 350 psig steam supply pressure is adequate for Case 6.

*** Includes planned overhaul outage of 412 hours/year.

**** Does not include planned overhaul outage.

Although the transport is conceptual in status, the potential reductions in costs and heat rate justify further testing at a proof-of-concept scale to demonstrate the technology.

Among the KRW-based IGCC cases, Case 1 has the best performance, with an efficiency advantage of approximately 5 percent over the oxygen-blown reference case. This improvement is due in part to the better energy utilization of a hot gas cleanup system. However, as pointed out in the Second Law analysis of Section 10, the reference case efficiency is adversely affected by the lower total carbon conversion of that design, which does not have a sulfator as in the hot gas cleanup cases. The Second Law analysis indicated that both cases could be improved by further optimization.

With regard to the relative costs of the IGCC cases (Table 3), transport IGCC again has a substantial advantage due to both high efficiency and low capital cost. The capital cost advantage, like the high efficiency, stems from superior gasification and desulfurization performance assumed for the transport reactor, which reduces reactor size and eliminates the need for external desulfurization. Comparing Case 1 with the reference case, air-blown gasification with hot gas cleanup is projected to have a 9-percent advantage in cost of electricity over oxygen-blown IGCC. An important conclusion of this study is that the major advantage of the Case 1 plant may be a substantial reduction in capital cost, amounting to 13 percent versus the oxygen-blown plant. As discussed in Section 11, a true capital cost savings would probably be of greater importance in The Southern Company than operating cost advantages. In particular, operating cost advantages arising from byproduct sales, as in the oxygen-blown reference case and DOE Case 5, would not be viewed as favorably as capital-derived savings because of market risks and added operating complexity.

Based on the current study, full hot gas cleanup using fixed beds of zinc ferrite sorbent coupled with sulfuric acid production (Case 5) would not appear to be an attractive IGCC alternative. Complete reliance on a zinc ferrite desulfurization system would not be acceptable to The Southern Company at the current level of development. Moreover, the higher capital cost and overall plant complexity for this case do not fit well with The Southern Company system. Potential improvements in this technology, some of which are outlined in Appendix B, must be pursued if this plant design is to gain widespread acceptance.

With regard to scale economies, Cases 1 and 3 provide an excellent indication of how plant scale affects both capital and operating costs, since Case 3 is essentially a half-scale version of Case 1. The conclusion from this analysis is that the capital cost scale economies over this size range are rather low (proportional to size in mega watts raised to the power of 0.8), which is a reasonable conclusion for highly modular technologies. Nonetheless, smaller plants would not be of

Table 3
Summary of Gasification Power Plant Costs

	ALL COSTS IN MID-1980 DOLLARS							OXYGEN BLOWN
	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6		
	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW
NET POWER OUTPUT, MW	458.4	99.6	231.7	399.1	420.2	409.5	413.5	
<u>CAPITAL COST</u>								
PLANT CONSTRUCTION COST	378 825	120 1206	219 944	356 893	412 962	329 803	405 960	
PROCESS CONTINGENCY	20 43	6 61	10 45	22 55	24 56	17 42	15 36	
PROJECT CONTINGENCY	80 174	25 254	46 197	76 190	87 208	69 169	84 203	
TOTAL PLANT COST	477 1042	151 1521	275 1188	454 1138	524 1247	415 1014	504 1218	
AFUDC	94 204	29 290	54 226	89 223	103 245	82 199	99 239	
TOTAL PLANT INVESTMENT	571 1246	180 1811	329 1413	543 1362	627 1492	497 1213	602 1457	
OWNER'S COSTS	44 96	14 136	25 107	40 99	50 120	27 66	37 90	
TOTAL CAPITAL REQ'T	615 1342	194 1947	354 1520	583 1461	677 1612	524 1279	640 1547	
<u>O&M COST*</u>								
FIXED O&M COSTS, \$/kW-YR	38.3	77.8	47.7	42.2	45.9	37.2	43.9	
VARIABLE O&M, mills/kWh	2.41	2.63	2.43	2.45	3.55	1.05	0.99	
FUEL COSTS, mills/kWh	12.9	14.0	12.8	13.9	14.0	12.2	13.6	
<u>COST OF ELECTRICITY**</u>								
CURRENT \$, mills/kWh	67.4	95.5	74.5	73.1	79.9	62.8	73.8	
CONSTANT \$, mills/kWh	49.1	69.4	54.2	53.2	53.1	45.6	53.4	

* FIRST-YEAR O&M COSTS

** TEN-YEAR LEVELIZED COST AT 85 % CAPACITY FACTOR.

significant interest for The Southern Company because of the large growth increments of the 30,000-MW system.

The ISTIG cases developed in the project were of more interest to DOE than to SCS. The advantages cited for this cycle include higher efficiencies and reduced capital costs due to elimination of the steam turbine cycle. These advantages were not seen in the current study, as noted in Tables 2 and 3. The ISTIG cases were less efficient than expected because of a limitation in the ability of the LM-8000 aeroderivative gas turbine to accommodate the full steam raising potential of the gasification island. It is not clear how this limitation could be overcome with the equipment specified in the study.

From the standpoint of SCS, the ISTIG plant concept is also unattractive at the present time because of lack of development support from the private sector. The LM-8000-based ISTIG is no longer supported by GE, which is now proposing a new ISTIG concept. In any case, GE has indicated in the past that it will not pursue development of these systems unless a threshold customer base is assured. SCS Research and Environmental Affairs does not intend to pursue further studies of the ISTIG concept unless additional progress toward commercialization is made. A potentially more attractive development along these lines is the Humid Air Turbine (HAT) cycle currently being studied under EPRI sponsorship.

A simple sensitivity analysis was conducted to evaluate uncertainties in the economic and financial analysis of the various case options. The major categories of uncertainty in the cost of electricity are:

- Cost of construction
- Fuel cost
- Capacity factor
- Efficiency of plant operation
- O&M costs

Most of the economic sensitivity relationships are approximately linear over the ranges evaluated, and relative risks can be illustrated graphically as shown in Figure 1. The capacity factor is clearly the most important variable and is influenced by design reliability and by system dispatching characteristics. After capacity factor, uncertainties in capital cost and the combined effect of thermal efficiency factors (carbon conversion, cold gas efficiency, auxiliary power consumption, and stack temperature) pose roughly equal risks. Coal cost and other O&M costs represent a relatively lesser magnitude of risk.

Figure 1

Relative Risks - Case 1
Increase in Cost of Electricity for Maximum Probable Uncertainty

ENVIRONMENTAL PERFORMANCE

Air Emissions

All IGCC plants are designed to efficiently reduce air emissions and effluents to regulatory limits or below. Environmental permits will, however, be required for air, water, and solid waste disposal and will be applied for at least 2 years prior to startup.

Air emissions are primarily from three sources: fines lockhopper vent gas, flue gas from the heat recovery steam generators (HRSGs), and flue gas from the sulfator. The lockhopper vent gas contains fine particulate from the fines removal lockhopper. The flue gas from the HRSGs contain SO_2 and NO_x .

The hot particulate removal step removes virtually all the particulate from the gas stream that goes to the HRSGs. The flue gas from the sulfator contains SO_2 , due to the oxidation of the sulfides to sulfate; NO_x and particulate are also present. SO_2 emissions are controlled by in-bed desulfurization in the gasifier and by removal of most of the remaining sulfur by a zinc ferrite sorbent, except for Case 5. In Case 5, zinc ferrite is used exclusively for desulfurization. For all cases, sulfur removal efficiencies greater than 99 percent are achieved.

Pilot plant measurements by MWK indicate that NH_3 and HCN in the fuel gas from KRW gasification result in NO_x emissions ranging from 0.05 to 0.2 pounds of NO_x per MMBtu when limestone is used for in-bed desulfurization. The cases in this study (except Case 5) were assumed to emit NO_x at rates ranging from 0.17 to 0.2 lb/per MMBtu. These rates are near the NSPS limit for a simple-cycle combustion turbine (75 ppmv at 15-percent O_2). Selective catalytic reduction (SCR) is needed in Case 5 (no limestone) to reduce the NO_x emissions to a level below the NSPS for combustion turbines. This is because chemical components containing fuel-bound nitrogen are substantially higher in the fuel gas when gasification occurs without limestone in the bed. All the cases would be required to add SCR if substantially lower NO_x emissions are mandated. The current trend with respect to oil- and gas-fired combined cycles is toward the imposition of NO_x limits stricter than the NSPS requirements for stand-alone combustion turbines.

Near absolute particulate removal is achieved in the fuel gas by the use of a cyclone and a ceramic candle gas filter. Pilot plant data from MWK indicate that particulate concentrations in the fuel gas are well below 10 ppm. However, the long-term reliability and operability of hot particulate filters is a major development issue which must be addressed.

SO_2 and NO_x emissions from the sulfator are controlled by limiting the temperature in the sulfator. Sulfur released from the char in the sulfator is

captured by the limestone and converted to calcium sulfate. In Cases 1 through 4, the sulfator flue gas is used for coal drying. The flue gas from the drying step is filtered in order to meet particulate emission standards prior to release.

Disposal of Solid Wastes

The major solid wastes, including spent limestone and ash, are disposed of in a dry landfill at the plant site. The limestone containing ash (LASH) discharged in Cases 1, 2, 3, 4, and 6 is assumed to be nonhazardous by current standards. Samples of LASH material from MWK pilot plant in-bed desulfurizations tests, both before and after sulfation treatment, were tested and in every case the EP-toxicity test results showed compliance with the Resource Conservation and Recovery Act of 1976 (RCRA) requirements. The solid waste discharged in Case 5, which has no limestone-derived solids, has also been determined to be nonhazardous by current standards.

The landfill used for disposal of solid wastes is lined with a 3-ft thick clay liner and all leachate and surface runoff from the landfill is routed to a sedimentation pond (also lined with clay) for removal of entrained solids prior to discharge at an existing water discharge point. Groundwater monitoring wells are located around the landfill. When ash is stacked to the maximum design height for the landfill, the material will be covered with a 2-ft layer of soil, and grass will be planted.

Water Discharges

Unlike conventional IGCC, most of the water streams to be discharged from gasification power plants with hot gas cleanup are the same as streams in conventional power plants. Water streams treated by conventional techniques include coal pile runoff and structure runoff. Runoff from the limestone storage area is routed to an existing ash pond along with the coal pile runoff to allow sufficient time for solids to settle prior to discharge.

In Case 5, a weak acid stream from the sulfuric acid plant must be neutralized. A sodium hydroxide solution is used to neutralize the stream prior to discharge.

COMPARISON OF AIR-BLOWN IGCC WITH HOT GAS CLEANUP VERSUS OXYGEN-BLOWN IGCC WITH COLD GAS CLEANUP

An air-blown, hot gas cleanup IGCC system differs in many respects from the more conventional oxygen-blown gasifier, cold gas cleanup system. The areas of most interest for comparison are:

- Reliability and Operability
- Environmental Effects

- Economics
- Overall Efficiency

The two plant types were compared in these three areas based on the air-blown IGCC with hot gas cleanup design for Case 1 and the updated oxygen-blown IGCC case with cold gas cleanup based on the aforementioned EPRI site-specific study. The features of these two cases can be compared in Tables 2 and 3 (shown previously).

Reliability and Operability

Basically, there is little difference in reliability or operability of the gasifiers themselves. The oxygen-blown case requires an oxygen plant with added complexity, but the air-blown system requires more gasifiers. Case 1 includes a sulfator and a limestone handling and feeding system. Beyond these considerations, the systems diverge with the oxygen-blown, cold gas system becoming essentially a chemical processing plant, including the Selexol™ acid gas removal process, the Claus process for recovering sulfur, and the SCOT process for treating tailgas. These processes are considered complex and unfamiliar to many traditional power plant operators.

For hot gas cleanup, components must be proven through demonstration runs. The performance of the ceramic candle filter element is a major concern. The lifetime of the zinc ferrite sorbent is also a major consideration. One key development item is the fuel gas control valve. The operation of the combustion turbine on low-Btu gas is a second key development item.

Emissions

Both cases can easily meet NSPS for SO₂ from coal-fired power plants. Both IGCC systems have similar capabilities for the removal of sulfur from the flue gas. However, the oxygen-blown design would require the addition of a COS hydrolysis step to match the removal level predicted for Case 1.

NO_x emissions for the two cases are both much lower than typical emissions of pulverized coal plants, and both meet the NSPS limit for standalone combustion turbines of 75 ppm in the turbine exhaust. However, for the oxygen-blown IGCC system, almost all of the fuel-derived nitrogen compounds formed in the gasifier are scrubbed out and appear in the sour condensate water. Thus, the only NO_x in the turbine exhaust is thermal NO_x derived from the air. This probably amounts to less than 25 ppm.

For the hot gas cleanup case, the presence of limestone in the gasifier accelerates cracking reactions and reduces the amount of fuel-bound nitrogen compounds

(chiefly ammonia) in the gasifier product gas. However, these compounds do remain and contribute to an NO_x exhaust level of about 60 ppm. If the NSPS limit is lowered to 25 ppm, some NO_x control procedures, such as SCR, would have to be employed for systems with hot gas cleanup.

One issue which may be of concern in the future is emissions of trace compounds referred to as air toxics. In this area, IGCC with cold gas scrubbing would have an advantage over systems based on hot gas cleanup. Coal-fired power plants are not currently regulated with respect to air toxics, and insufficient information is currently available to fully quantify this advantage.

Cost Comparisons

In terms of costs, the estimated cost of electricity for Case 1 is 67 mills/kWh (49 mills in constant dollars) versus 74 mills/kWh (53 mills in constant dollars) for the oxygen-blown case. Air-blown gasification with hot gas cleanup has significantly lower capital costs (13 percent) and a better heat rate (5 percent) than the oxygen-blown case used in this study.

SYSTEM STUDY OF AIR-BLOWN IGCC WITH HOT GAS CLEANUP VERSUS A PULVERIZED COAL UNIT WITH FGD

A production cost analysis for The Southern Company was performed to compare the difference in system revenue requirements associated with the installation at Plant Wansley of 811 MW of IGCC capacity based on the Case 1 design versus an 811-MW pulverized coal (PC) unit with flue gas desulfurization (FGD). Either plant would be commissioned in the year 2006 in keeping with the system integrated resource plan. Cost differentials included costs due to both capital investment and operating expenses over a 45-year plant life.

This analysis showed 1990 present worth savings of over \$600 million if the air-blown IGCC unit were installed rather than the PC unit. Most of the savings were attributed to reduced system operating expenses, primarily due to lower fuel costs resulting from the higher efficiency of the IGCC plant. However, approximately \$150 million (1990 present worth) in capital cost savings were realized because of the lower overnight capital cost and shorter construction time of the IGCC plant.

The conclusion from this analysis is that air-blown gasification with hot gas cleanup appears to have a large economic advantage over a PC plant with FGD. The predicted economic advantage depends primarily on the thermal efficiency of the IGCC plant, which can be predicted with greater confidence than relative capital costs. The capital costs would appear to be roughly the same for both plant types, within the limits of estimating error. IGCC based on the transport

reactor in a hybrid system (Case 6) would offer even greater cost savings than Case 1, although a detailed system revenue requirements analysis was not done for this case.

SECOND-LAW AND THERMOECONOMIC ANALYSIS

An important issue in the study of air-blown IGCC with hot gas cleanup is the potential for further improvements in plant design through design optimization. The Second Law (exergy analysis) and the thermoeconomic analysis applied in this study identify and evaluate the real energy waste and the real cost sources in IGCC design options. With the aid of these analyses, conclusions can be drawn without redesigning the total plant.

A detailed Second-Law analysis and comparison were conducted for two 400-MW IGCC designs: the air-blown system with hot gas cleanup (Case 1) versus the oxygen-blown system with cold gas cleanup (modified reference case). This analysis indicated that although the oxygen-blown KRW gasifier has a higher overall efficiency, this advantage is offset by lower efficiency in the gas turbine, compared to Case 1. Both cases were found to be in need of further optimization. Several possible design changes were identified for the air-blown IGCC plant to improve thermal efficiency in a cost-effective way. These design changes include:

- Operating the gasifier in the temperature range of 1850 to 1870°F versus 1900°F as designed.
- Redesigning the recycle gas cooler, IP economizer, and IP drum to avoid heat transfer through the pinch point.
- Studying the economic feasibility of developing a gas turbine system with a reheat stage.
- Exploring further the option of utilizing heat recovered in the ash sulfation step for power production (steam generation) rather than coal drying.

SCS recommends that these and other design improvements be investigated further to determine their feasibility and impact on the capital and operating costs for the air-blown IGCC plant with hot gas cleanup.

IMPLEMENTATION OF IGCC FOR THE SOUTHERN COMPANY

Based on the needs and system characteristics of The Southern Company, air-blown IGCC with hot gas cleanup would appear to have many of the features that would favor implementation of this technology, including lower capital costs, inherently high thermal efficiency, short lead times, and emphasis on dry cleanup and solids handling over refinery-type gas cleaning operations. Advantages often cited for conventional IGCC plants such as lower space requirements, reduced solid waste volume, byproduct production, and highest environmental performance would not appear to outweigh the advantages of IGCC with hot gas cleanup in The Southern Company. However, special advantages of oxygen-blown IGCC with cold cleanup, such as compatibility with liquid fuels production and integration with certain industrial facilities, could favor this type of system.

The implementation of a new baseload technology such as IGCC by 2006 requires an aggressive schedule of R&D, technology transfer, and preliminary engineering activities in the immediate future. It is essential that IGCC with hot gas cleanup be demonstrated at utility scale within the 1990s to allow this technology to be considered for the next baseload unit in The Southern Company.

There are a number of development issues for air-blown gasification power plants with hot gas cleanup, as summarized in Table 4. Barrier issues should be addressed in near-term development and commercialization plans.

Table 4

**Major Development Issues For Air-Blown Systems
With Hot Gas Cleanup**

KRW GASIFICATION

- Scaleup to commercial size
- Gasifier throughput

TRANSPORT GASIFICATION

- Hydrodynamic performance
- Kinetics of reactions
- Sulfur capture efficiency
- Carbon conversion
- Detailed design and scaleup
- Capital costs

HOT GAS CLEANUP

- Reliability, performance, and cost of hot particulate filters
- Performance, costs, and system effects of zinc ferrite; improved sorbents
- Fate of fuel-bound nitrogen in fuel gas and effects on NO_x emissions
- Carryover of volatile alkali compounds to the combustion turbine
- Emissions of trace compounds (potential air toxics), notably mercury and lead

COMBUSTION TURBINE (CT)

- Combustion characteristics of low-Btu gas
- Extraction of gasifier air from the CT compressor
- Cooling of the CT components and blades
- Fuel control valve for high temperature fuel gas
- Control issues relating to integration between CT and the gasification island

OTHER ISSUES

- Characteristics of sulfide-containing waste solids; potential disposal costs and liability
- Performance of materials and components under exposure to high temperatures and pressures (e.g., valves, piping, heat exchangers)

1.0 INTRODUCTION

1.1 BACKGROUND

The Southern Company is one of the nation's largest investor-owned electric utility systems, supplying energy to approximately 10 million people across the Southeastern United States. The Southern Company is the parent firm of five electric utilities: Alabama Power Company, Georgia Power Company, Gulf Power Company, Mississippi Power Company, and Savannah Electric and Power Company. Southern Company Services, Inc., (SCS) provides engineering, research, and technical support for the five utilities. The companies of The Southern Company represent over \$18 billion in assets and over 30,000 MW of generating capacity, of which approximately 80 percent is coal-fired.

The present expansion plan for The Southern Company estimates that the next baseload generation will be added in about 2006. Between 1995 and 2006, more than 6000 MW of peaking, storage, and intermediate-capacity generation is projected to be added. However, the recent fluctuations in petroleum prices have caused uncertainties in the system expansion plan in that large changes in energy prices can greatly affect the economy and hence the rate of growth in electric energy usage. Currently, system expansion models predict that oil- or gas-fired, simple-cycle combustion turbines will be the most economical technology for peaking generation, based on projected oil and gas prices. The models also predict that oil- or gas-fired combined-cycle plants will be the most economical choice for intermediate-load capacity.

The Southern Company currently operates 1120 MW of combustion turbines, primarily for peaking service. One plant in Mississippi Power Company's service area is operated as a cogeneration plant with two 13,750-kW General Electric (GE) combustion turbines with heat recovery steam generators. This plant has a capacity factor of 95 percent. The Southern Company experience with this unit has shown that combustion turbines can be extremely reliable for baseload operation.

However, the large uncertainty in the availability and prices of oil and natural gas is a major concern. Because of previous experience with shortages of oil and gas, there is reluctance to commit to an expansion plan based heavily on these fuels. Contingency options being considered to supply peaking fuels in the event of a major oil shortage include stockpiling petroleum and direct production of synthetic fuels from coal. Methanol coproduction may also be an attractive option for the supply of peaking fuels, although this requires an oxygen-blown gasifier.

The increasing emphasis on combustion turbines in The Southern Company generation expansion plan, along with concerns about their fuel supply, have focused attention on coal gasification as a means of deriving premium fuels from coal. Moreover, integrated gasification combined cycle (IGCC) is viewed as a potentially attractive technology for baseload expansion. IGCC plants are highly efficient, have a relatively low operating cost, and have the best environmental performance of any coal-fired plants. IGCC plants also have the potential to produce methanol for use as a peaking fuel.

A site-specific study conducted by SCS for the Electric Power Research Institute (EPRI) indicated that an IGCC plant based on an oxygen-blown KRW gasifier with Selexol™/Claus/SCOT sulfur cleanup (referred to in this report as cold gas cleanup) is a competitive baseload power generation technology compared to a baseload pulverized-coal-fired unit currently called for in The Southern Company generation expansion plan. The IGCC plant was assumed to be added at Georgia Power Company's Plant Wansley near Atlanta, Georgia.

As a result of the study, SCS identified a number of cycle changes which could potentially improve the cost, performance, and operability of gasification-based power plants. One of the cycles identified was air-blown IGCC with hot gas cleanup. The use of air-blown IGCC plants with hot gas cleanup offers the potential for lower capital cost and higher cycle efficiencies than first generation IGCC plants with cold gas cleanup. The capital cost of the air-blown IGCC plant with hot gas cleanup should be lower due to elimination of the oxygen plant and low-temperature gas scrubbing train. Cycle efficiencies should be higher for a system utilizing hot gas cleanup because the fuel gas is fed to the turbine at a much higher temperature than in a cold gas cleanup system. Air-blown IGCC plants with hot gas cleanup are simpler than oxygen-blown plants and may be more compatible with the operating practices of utilities based primarily on pulverized-coal technology.

To investigate the effects of air-blown IGCC power plants with hot gas cleanup, SCS entered into a follow-on site-specific study with the U.S. Department of Energy Morgantown Energy Technology Center (DOE/METC). The focus of the study was to evaluate various power plant configurations to determine if air-blown gasification with hot gas cleanup is competitive against either oxygen-blown IGCC with cold gas cleanup or a conventional pulverized-coal-fired plant. The study included investigating the effects that different power island configurations, plant sizes, methods of sulfur removal, and gasifier types had on the design and costs of the air-blown gasification-based power plant with hot gas cleanup. As in the EPRI IGCC study, the air-blown gasification power plants were assumed to be located at Plant Wansley.

2.0 PROJECT DESCRIPTION

2.1 OBJECTIVES

2.1.1 Overall Objectives

The major objectives of the joint SCS/DOE study of air-blown gasification power plants with hot gas cleanup are to:

- Evaluate various power plant configurations to determine if an air-blown gasification-based power plant with hot gas cleanup can compete against pulverized coal with flue gas desulfurization for baseload expansion at Georgia Power Company's Plant Wansley.
- Determine if air-blown gasification with hot gas cleanup is more cost effective than oxygen-blown IGCC with cold gas cleanup.
- Perform Second-Law/Thermoeconomic Analysis of air-blown IGCC with hot gas cleanup and oxygen-blown IGCC with cold gas cleanup.
- Compare cost, performance, and reliability of IGCC based on industrial gas turbines and ISTIG power island configurations based on aeroderivative gas turbines.
- Compare cost, performance, and reliability of large (400 MW) and small (100 to 200 MW) gasification power plants.
- Compare cost, performance, and reliability of air-blown gasification power plants using fluidized-bed gasifiers to air-blown IGCC using transport gasification and pressurized combustion.

Seven design concepts were selected for the study to accomplish the overall objectives. The primary system components and parameters for the seven cases are compared in Table 2-1.

2.2 PROJECT ORGANIZATION

The overall structure for the project is illustrated in Figure 2-1. The project team consisted of SCS, the M.W. Kellogg Company (MWK), Tennessee Technological University (TTU) Center for Electric Power, and GE. SCS Research and Environmental Affairs was responsible for overall project management, environmental assessment, plant integration, and overall cost estimates. SCS engineering departments were responsible for the design and costing of the power island, the materials handling systems, and the balance of plant, as well as reliability-availability-maintainability (RAM) analysis and system analysis.

Table 2-1

Primary System Components and Parameters
For the Study Cases

CASE	GASIFIER TYPE	NOMINAL PLANT SIZE (MW)	CYCLE	SULFUR CLEANUP			FEATURES
				IN-BED	ZINC FERRITE		
1	KRW	400	IGCC ¹	✓	POLISHING		
2	KRW	100	ISTIG ²	✓	POLISHING		LM-8000 CT ³
3	KRW	200	IGCC	✓	POLISHING		
4	KRW	400	ISTIG	✓	POLISHING		LM-8000 CT
5	KRW	400	IGCC	—	BULK		H ₂ SO ₄ PLANT
6	TRANSPORT	400	IGCC	✓	—		HYBRID
7	TRANSPORT PYROLIZER	400	IGCC	✓	—		MILD GASIFICATION

Notes:

1. Integrated gasification combined-cycle
2. Integrated steam injected gas turbine
3. Combustion turbine

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

Figure 2-1 Project Organization and Responsibilities

MWK was responsible for the gasifier island designs and costs. TTU performed the Second-Law and thermoeconomic analyses. GE provided combustion turbine performance and cost data.

2.3 STUDY CASES

Following is a summary description of the seven plant system configurations addressed in the study. Case 1 is the base case for system comparisons.

2.3.1 Case 1: 400-MW Air-Blown IGCC With Hot Gas Cleanup

This system configuration represents the base case design and consists of a nominally 400-MW power plant with air-blown, KRW fluidized-bed gasifiers employing in-bed desulfurization. Figure 2-2 presents a simplified schematic of the system configuration. The system includes two heavy-duty GE MS7001F industrial gas turbines (2300°F firing temperature) coupled with one reheat steam turbine bottoming cycle. Spent limestone and ash from the gasifier are oxidized in an external sulfator prior to disposal.

The hot gas cleanup system consists of hot particulate removal followed by hot gas desulfurization. Hot particulate removal consists of a cyclone followed by an absolute, ceramic-candle-type filter. Gas temperatures are maintained at approximately 1000°F. Solids collected in the particulate-capture equipment are sent to the external sulfator prior to disposal. Supplemental hot gas desulfurization is accomplished by means of a fixed-bed zinc ferrite system for sulfur polishing, or removal of a final increment of sulfur not easily removed in the fluidized bed. Regeneration off-gas from the polishing desulfurization system is recycled to the gasifier for in-bed sulfur capture.

2.3.2 Case 2: 100-MW ISTIG

The gasifier island portion of this system is similar in configuration to the system described in Case 1. However, in this case, only one train of gasification is employed, so that the total power output is reduced to the range of 100 MW. The power island is based on a GE aeroderivative intercooled steam-injected gas turbine (ISTIG) rather than the industrial turbine used in Case 1. Due to the use of an ISTIG, the steam turbine cycle is eliminated and steam from the HRSG is used as a reactant in the gasifier and for power generation in the gas turbine. Additionally, the high-pressure steam system is eliminated in this case. A simplified block flow diagram of the system configuration is given in Figure 2-3.

MWK

Figure 2-2 Simplified Schematic of the System Configuration for IGCC Cases 1 and 3

CASES 2 & 4 100-MW and 400-MW ISTIG

Figure 2-3 Simplified Schematic of the System Configuration for ISTIG Cases 2 and 4

2.3.3 Case 3: 200-MW Air-Blown IGCC With Hot Gas Cleanup

The configuration for Case 3 is identical to that in Case 1 with the exception of the number of trains of gasification, cleanup, and gas turbine power generation. The primary objective of examining this case is to evaluate the effect of plant size on the cost of air-blown gasification with hot gas cleanup. This system is based on a single advanced GE MS7001F gas turbine, and thus is the minimum size IGCC configuration possible with this turbine. Despite the small size of this plant, a reheat steam cycle is available and has been used. The gasification configuration is identical to that in Case 1.

2.3.4 Case 4: 400-MW ISTIG

For this case, four GE ISTIG trains are powered by a fluidized-bed gasification system configured as in Case 2. The total power output of this plant will nominally be 400 MW, and therefore this case will be closely comparable to the base case. This configuration is a modular scaleup of Case 2 with respect to the gasification and combustion turbine systems, with a total of four trains each equivalent to the single train of Case 2.

2.3.5 Case 5: 400-MW IGCC with Bulk External Desulfurization

The configuration for this case is similar to that in the base case except that desulfurization with limestone in the KRW fluidized-bed gasifier is eliminated. Bulk fuel gas desulfurization is done externally by zinc ferrite. In this design, a sulfuric acid plant is added to treat the zinc ferrite regeneration gas, producing sulfuric acid as a saleable byproduct. Because of the elimination of in-bed desulfurization with limestone, no sulfator vessel is required. The power island design for this case is essentially the same as that in the base case. Figure 2-4 illustrates the overall system configuration for Case 5.

2.3.6 Case 6: 400-MW IGCC with Transport Gasifier and Sulfator

Case 6 is similar to Case 1 except that the KRW fluidized-bed gasifier is replaced with a Kellogg circulating-bed gasifier. This gasifier operates at gas velocities which extend beyond those necessary for fluidization into the regime of particle transport or pneumatic conveying, and thus is termed a "transport" reactor. The system configuration illustrated in Figure 2-5 includes a pressurized transport sulfator which is fed with char/ash from the gasifier. The transport gasifier operates with in-bed desulfurization by limestone. Because of the small coal and limestone particle sizes, sulfur capture is very rapid in the gasifier and high levels of sulfur removal are obtained. Therefore, no external zinc ferrite system is required for this case, and consequentially chloride guard beds have been eliminated. With respect to the power island, the flowsheet for Case 6 is similar to that described for Case 1.

CASE 5
400-MW IGCC

DESIGN SCOPE

Figure 2-4 Simplified Schematic of the System Configuration for IGCC Case 5

DESIGN SCOPE

Figure 2-5 Simplified Schematic of the System Configuration for IGCC Case 6

2.3.7 Case 7: 400-MW IGCC with Mild Gasification, Transport Reactors

The final case is similar to Case 6 with the addition of a mild gasification upstream of the char gasifier, as shown in the simplified system configuration for Case 7 in Figure 2-6. The char produced in the mild gasifier is used as the gasifier feed and the liquid product is stored for peaking turbine fuel. As in Case 6, the gasifier is a Kellogg transport design rather than the KRW fluidized-bed type. A transport sulfator is included to oxidize the ash produced by the gasifier. Again, no external zinc ferrite system is required for final sulfur polishing. The power island portion of this plant is based on the GE MS7001F advanced industrial combustion turbine and a reheat steam bottoming cycle, as in the base case.

2.4 TECHNICAL APPROACH

2.4.1 Overall Approach and Technology Selection

The IGCC systems study assesses a number of different power generation designs based on coal gasification encompassing two different gasifier types, two combustion turbine configurations, two approaches to fuel gas desulfurization, two basic plant sizes, and a mild gasification configuration. All of these cases use air rather than oxygen as the gasification oxidant, and all employ hot gas cleanup rather than conventional cold cleanup. All the systems were based on the Plant Wansley site of Georgia Power Company, which is one of several candidate sites for baseload expansion in The Southern Company. For each case, process designs were developed, major equipment items were identified, and estimates of capital and O&M costs were developed. The latest results available from research projects and studies were used to estimate performance and costs. Two basic cycles were considered: combined cycles and ISTIG cycle. Combined-cycle units are closer to commercial readiness and represent incremental extensions of demonstrated technology. ISTIG units reduce cycle complexity by eliminating the steam turbine cycles and also offer the potential for higher thermal efficiencies. However, the ISTIG turbine is not commercially available nor is there a firm commitment to produce such a machine.

Two types of gasifier technologies were selected for the study: the KRW fluidized-bed gasifier and the M.W. Kellogg transport gasifier. The KRW gasification process was selected because of a number of important advantages over other technologies:

- It is an economic and efficient process capable of processing a wide variety of coals with significantly different properties. The process operates at temperatures which provide excellent chemical and thermal efficiency.

- It is safe and operable as demonstrated by an extensive data base of over 13,000 hours of pilot plant experience. The system features simplicity of control, load following capability, and safe shutdowns in the event of steam or oxidant interruptions.
- It is environmentally acceptable, yielding a solid ash that is nontoxic by current U.S. Environmental Protection Agency standards, and a product gas essentially free of hydrocarbons.

Case 6 is based on the transport gasifier/sulfator concept currently under development by The M.W. Kellogg Company. The transport concept, which is based on Kellogg's fluidized-bed catalytic cracking designs, utilizes highly reactive pulverized coal and limestone transported at relatively high velocities through a pressure vessel. The solids are removed from the gas stream by a series of cyclones and are recycled to the gasifier. The transport reactor was selected for the current study because of three major advantages over other systems:

- The transport gasifier is projected to operate at 1600°F or higher, and thus would not produce significant tars as does a fixed-bed gasifier. In contrast to entrained-flow systems, the transport gasifier does not require control of liquid slag for stable operation and is better matched to hot gas cleanup in terms of temperature.
- The transport reactor uses finer coal and limestone than the KRW fluidized-bed gasifier (or other fluidized-bed units), and thus should achieve higher rates of reaction. This feature improves limestone and carbon utilization and minimizes reactor size.
- The transport reactor is itself a pressure vessel, and is therefore simpler to construct and maintain than other circulating pressurized-fluidized-bed combustion (PFBC) and gasification systems that consist of a reactor inside a large pressurized shell.

The study estimated the benefits of using a hot gas cleanup/desulfurization process so that fuel gas does not have to be cooled to near ambient temperature to remove particulates and sulfur. The concept being utilized in Cases 1 through 5 has been demonstrated at pilot scale at KRW's Waltz Mill site in Madison, Pennsylvania (Reference 1).

Despite the fact that certain plant systems are not yet commercially available, the plant designs were developed under the assumption of technological maturity. In other words, each design is for an "nth" plant, with several demonstration or early commercial plants assumed to have preceded the plant under consideration. The cost estimates do not, however, reflect widespread penetration of the technology in the utility market (e.g., 20 or more plants).

Reference Case: SCS/EPRI Site-Specific IGCC Study

An important aspect of the technical approach to this study was to develop the plant designs and costs on a consistent basis with the previous SCS/EPRI study of an IGCC power plant at Georgia Power's Plant Wansley (Reference 2). This IGCC power plant consists of oxygen-blown gasification, heat recovery, fines removal, gas cooling, recycle gas compression, Selexol™ acid gas removal, Claus sulfur recovery, and SCOT tailgas treatment. The gasification area consists of three KRW fluidized-bed reactors each designed at 42-percent throughput to achieve high plant availability. The power island consists of two GE MS7001F combustion turbines operating in a combined-cycle system containing two heat recovery steam generators and a single reheat steam turbine. A simplified block flow diagram of the IGCC plant is shown in Figure 2-7 along with a conceptual plot plan in Figure 2-8.

The cases addressed in the present study employ the same design basis as that used in the EPRI case to facilitate a comparison of results obtained from the EPRI-sponsored and DOE-sponsored studies. The design approach, costing methodology, and economic evaluations are on a common basis. The EPRI case is closer to commercialization than the cases addressed here, thus assumptions made about the commercialization path of air-blown gasification with hot gas cleanup are addressed in subsequent sections of this report.

Because of changes in technology and engineering experience since the EPRI study, it was necessary to update the EPRI study design and costs in certain areas to maintain comparability. This study update was conducted by SCS in partnership with EPRI. The major impetus for updating the EPRI design was the significant revision of the GE MS7001F gas turbine performance rating by GE.

2.4.2 Gasifier Island Design Methodology

The gas-island design was performed through the use of Kellogg's proprietary design models in a logical stepwise manner that integrates these advanced technologies with the power island design. The following steps were executed for the plant design in each case, as illustrated in Figure 2-9.

- Based on the specified coal composition and site conditions, a design basis document was prepared which contains important design-specific information for major plant sections of the gas island. This document also identifies the interfaces/conditions between the gas and power island.
- Kellogg's proprietary computer programs were used to estimate the gasifier performance for each of the cases.

Figure 2-7 Simplified Block Flow Diagram of the SCS/EPRI Oxygen-Blown IGCC Plant With Cold-Gas Cleanup (Reference Case)

Figure 2-8 Conceptual Plot Plan for the 400-MW SCS/EPRI
IGCC Reference-Case Plant

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Figure 2-9 Plant Design Steps

- An ASPEN simulation model was used to calculate heat and material balances in the plant sections downstream of the gasifier.
- The resulting fuel gas compositions were then provided to the combustion turbine designers to obtain performance of both turbine types.
- Based on the fuel gas required for each gas turbine, the number of gasifiers for the gas island was calculated. This determination included adjustments in spare capacity to achieve plant availability.
- The material balance for the complete plant was then calculated and performance tables were prepared based on the number of gasification trains.

2.4.3 Power Island Design Methodology

The design of the power island portion of the different gasification power plants was accomplished through these steps:

- Interface points between the power island and gasifier island were established based on the needs of the gasifier island and the ability of the power island to meet these needs.
- Preliminary fuel composition for the gasifier fuel stream was provided by Kellogg based on the overall process design of the gasification island.
- Gas turbine performance based on the preliminary fuel composition was determined, consisting primarily of fuel flow requirements and gas turbine compressor discharge air conditions.
- After final sizing of the gasifier by Kellogg, interface flow requirements for the power island were finalized.
- Coal and limestone handling designs and gas and steam turbine performance were determined based on the finalized gasifier interface flow requirements.
- The balance of plant and general facilities were designed to meet the requirements of each plant design.

The methodology for final equipment design within the power island itself consisted of these steps:

- From the final heat and material balances, thermal design parameters were established for the major power island equipment.
- Equipment manufacturers were consulted for the major power island equipment regarding cost, equipment arrangement, auxiliary requirements, and internal efficiencies.
- Final cost and performance data for the integrated IGCC unit were developed.

All thermal modeling of the power island cycles was done using the SYNTHA thermal simulation computer program.

2.4.4 Second-Law/Thermoeconomic Analysis Technical Approach

The Second-Law Analysis compared from a thermodynamic standpoint the design of the air-blown IGCC power plant with hot gas cleanup in Case 1 to the design of the oxygen-blown IGCC power plant with cold gas cleanup in the reference case. The thermodynamic evaluation included a comparison of the real energy waste in the corresponding areas of each design option. The exergy destruction in each plant component and the exergy losses associated with material streams rejected to the environment provide the basis for comparing the overall plant performance as well as the performance of oxygen-blown versus air-blown KRW gasifiers and cold gas versus hot gas cleanup.

A thermoeconomic evaluation of Case 1 assessed the performance of the most important plant components or group of components from a combined thermodynamic and economic viewpoint using design parameter changes to estimate the potential for improving the overall plant cost effectiveness. Design inefficiencies were also identified with recommendations for improving the cost effectiveness of the plant design.

The formal procedures for the Second Law/Thermoeconomic analyses are based on the application of methods which are highly specialized and are still undergoing development. These methods are discussed in greater detail in Section 10.2 of the report.

2.5 ORGANIZATION OF THE REPORT

The Contents of the report provide a detailed outline of the report material. The study results are presented logically in the following sequence: background material (Sections 1 and 2), design basis information (Section 3), detailed plant descriptions (Sections 4 through 6), cost estimates (Section 7), economic and financial analyses (Section 8), and various additional analyses and conclusions (Sections 9 through 12). With respect to the detailed descriptions of the various study cases, it will be noted that the cases for which full designs were developed are divided into two major groups by type of thermodynamic cycle. The IGCC cases (1, 3, 5, and 6) are discussed in Section 4, and the ISTIG cases (2 and 4) are treated separately in Section 5. The mild-gasification IGCC plant (Case 7) is discussed in Section 6. The cost and economic analyses treat all completed plant designs (Cases 1 through 6) collectively.

3.0 STUDY CRITERIA AND DESIGN BASIS

The study criteria were selected to match the needs and requirements of The Southern Company. The gasification power plant is assumed to be sited at Plant Wansley of Georgia Power Company. The Plant Wansley site is one of several sites around the system that is available for possible expansion of baseload generation. By locating the gasification power plant at an existing site with the capability for expansion, permitting would be streamlined and shared facilities and services could be utilized. This should ultimately lower the capital and O&M cost for the gasification-based plant compared to a "greenfield" gasification power plant.

3.1 THE SOUTHERN COMPANY GENERATION EXPANSION PLAN

Electricity generation options in The Southern Company include nuclear, coal, oil, and gas in conventional steam units; combustion turbines for peaking service; and hydroelectric plants. The current capacity mix for The Southern Company is shown in Table 3-1. For 1990, The Southern Company is forecast to have a peak demand of approximately 28,000 MW. This demand is projected to grow at an average compound growth rate of 2.3 percent per year over the next 10 years. The Southern Company currently has ample baseload capacity and the need for more is not indicated until 2006. Approximately 79 percent of capacity is baseload capacity (coal plus nuclear), while the long-range plan suggests that approximately 59 percent should be baseload capacity. The projected mix for 2006 is also shown in Table 3-1.

Table 3-1

The Southern Company Capacity Mix

	1990		2006	
	MW	%	MW	%
Nuclear	5,385	16.1	5,385	11.4
Coal ^a	21,325	63.8	24,699	52.3
Oil/Gas Steam	1,187	3.6	505	1.1
Hydro/Pumped Storage	4,124	12.3	5,122	10.8
Combustion Turbines	1,413	4.2	11,531	24.4
TOTAL	33,434	100.0	47,242	100.0

a. Includes cycling coal, but excludes firm off-system capacity sales.

The projected expansion plan assumed for this study is shown in Table 3-2. Generating units presently committed are expected to supply adequate capacity through 1994. The next new capacity in The Southern Company is projected to be peaking capacity, followed by intermediate and baseload capacity. The 1990 base expansion plan shows combustion turbines being added as peaking capacity beginning in 1995, combined-cycle units as intermediate capacity beginning in 2002, and coal-fired units as baseload capacity beginning in 2006.

The study assumes that the gasification power plant would be completed and placed in service in the year 2006. The timing of the gasification power plant is based on the projected schedule for adding baseload pulverized-coal-fired capacity on the system. This assumption was made since the gasification power plant has capital and operating costs similar to those of the pulverized-coal-fired unit, has similar operating characteristics, and would be commercial in the timeframe under consideration.

The study assumes that the gasification power plant would not be constructed using the phased-construction approach, in which combustion turbines or combined-cycle units are initially constructed and gasifiers are added later. There were no significant benefits identified for using phased construction in the base implementation plan. Projected capacity additions show the need for combustion turbines, combined-cycle units, and large coal-fired units. Therefore, if the combustion turbines and combined-cycle units were converted to gasification power plants, they would have to be replaced with new combustion turbines and combined-cycle units. In addition, there are no advantages in staging the capacity additions to match load growth since projected load growth is sufficient to justify adding the large capacity increments of combined-cycle or coal-fired units. Smaller utilities may gain advantages from staged construction if the increase in load is smaller than the gasification power plant because this would allow them to bring capacity online closer to the time when it is needed. This is not predicted to be the situation for The Southern Company.

3.2 STUDY CRITERIA

3.2.1 Site Characteristics

The Plant Wansley site is approximately 40 miles southwest of Atlanta, Georgia, and is located on the Chattahoochee River (Figure 3-1). Georgia Power Company (GPC) owns this 5262-acre site and the two existing 865-MW pulverized-coal-fired steam units. The site was originally developed for up to four

Table 3-2

**Generation Expansion Plan Used in the
Plant Wansley IGCC Study
(Major Unit Additions)**

<u>Year</u>	<u>Type of Addition</u>	<u>MW Added</u>	<u>Year</u>	<u>Type of Addition</u>	<u>MW Added</u>
1991	Coal	667	2003	Peaking	1260
	Pumped Hydro	75			
1992	Pumped Hydro	75	2004	Intermed. Peaking	1288 504
1993			2005	Intermed. Peaking	644 378
1994	Peaking	948	2006	Peaking Coal	252 811
1995	Peaking	1134	2007	Peaking Coal	378 811
1996	Peaking	252	2008	Peaking Coal	126 811
	Pumped Hydro	848			
1997	Peaking	1009	2009	Peaking Coal	504 811
1998	Peaking	882	2010	Peaking	504
1999	Peaking	882			
2000	Peaking	1386			
2001	Peaking	1008			
2002	Peaking	1512

Figure 3-1 Plant Wansley Site Location Map

865-MW coal-fired steam units. Only two units were actually built because of concern that two additional units might adversely impact ambient air quality in the Atlanta area unless flue gas desulfurization were added. Figure 3-2 is an aerial view of the site. The site has 324 surface acres for ash disposal and a large lake to supply makeup and cooling water. Ambient conditions for the site are listed in Table 3-3.

Mechanical draft cooling towers are used to cool the plant circulating water. No high-temperature water is discharged to the lake or river. The water supply is sufficient for more than 3500 MW of conventional pulverized-coal-fired steam generation, with only 1730 MW currently installed. The characteristics of makeup water from the water supply at Plant Wansley are listed in Table 3-4. Since gasification-based power plants generally require less water than conventional pulverized-coal-fired generation, there is sufficient water for the proposed unit and several additional units if needed.

In addition to the ash disposal and water supply for makeup and cooling water at Plant Wansley, other facilities can also be shared with the gasification power plant. For this study, the major additional areas in which sharing of capital facilities is assumed are as follows: basic site development, including certain roads and main rail facilities; coal receiving, unloading, and long-term storage; some portions of administrative and service buildings, including such items as plant management, security facilities, and other general site building space. However, it should be noted that the gasification power plant is not located immediately adjacent to the existing pulverized coal units. The adjacent space is reserved for two new pulverized coal units, even though the gasification power plant would probably take the place of one such unit if selected. The decision to separate the gasification-based plant from the existing coal units reduced somewhat the potential for sharing of general facilities.

In terms of operating costs, it was assumed that the manpower requirements of the gasification-based power plant would be reduced slightly by locating the plant at an existing site. Manpower sharing would apply primarily to management and administrative staff. Specific assumptions regarding plant staffing are described in Section 8 of the report.

3.2.2 Coal and Limestone Characteristics

The base coal is an Illinois No. 6 bituminous coal from the Old Ben No. 26 mine. Table 3-5 shows a typical analysis for this coal. The coal would be delivered by barge to the GPC barge transloader in north Alabama. From there, the coal would be loaded on unit trains and shipped to the Wansley site. Estimated

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

Figure 3-2 Aerial View Of The Plant Wansley Site

Table 3-3

Site Data Assumed for Plant Wansley

Location:	Near Roopville, Georgia
Ambient Pressure:	14.4 psia
Ambient Temperatures:	
Dry Bulb	90°F
Wet Bulb	70°F
Relative Humidity:	60%

Table 3-4

Makeup Water Composition

All units, except pH, in mg/L

Ionic Concentrations

<u>Cations</u>		<u>Anions</u>	
Ca ⁺²	14	HCO ⁻	20
Mg ⁺²	7	SO ⁻	9
Na ⁺¹	<u>24</u>	Cl ⁻	<u>16</u>
	45		45

Other Characteristics

pH	7.1
Alkalinity (as CO ₃ ⁻²)	3
Silica (as SiO ₂)	9.5
Iron (as Fe)	2.0
Color (APHA ^a)	10 to 140 (< 30 for 50% of the time)
Turbidity (JTU ^b)	10 to 150 (< 40 for 75% of the time)

- a. American Public Health Association
 b. Jackson turbidity units

Table 3-5

Base Coal Analysis - Illinois No. 6 Seam, Old Ben No. 26 Mine

<u>Proximate Analysis</u>	<u>As-Received (wt%)</u>	<u>Dry Basis (wt%)</u>
Moisture	11.12	-
Ash	9.70	10.91
Volatile Matter	34.99	39.37
Fixed Carbon	<u>44.19</u>	<u>49.72</u>
TOTAL	100.00	100.00
HHV (Btu/lb)*	11,666	13,126
<u>Ultimate Analysis</u>	<u>As-Received (wt%)</u>	<u>Dry Basis (wt%)</u>
Moisture	11.12	-
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen (by difference)	<u>6.88</u>	<u>7.75</u>
TOTAL	100.00	100.00
<u>Ash Fusion Temperature</u>	<u>Reducing (°F)</u>	
Initial Deformation	2195	
Softening (H=W)	2260	
Softening (H=1/2W)	2345	
Fluid	2415	
<u>Mineral Analysis of Ash</u>	<u>Ignited Basis (wt%)</u>	
Silica, SiO ₂	48.31	
Alumina, Al ₂ O ₃	19.48	
Titania, TiO ₂	0.92	
Ferric oxide, Fe ₂ O ₃	16.90	
Lime, CaO	5.13	
Magnesia, MgO	0.85	
Potassium oxide, K ₂ O	2.18	
Sodium oxide, Na ₂ O	1.23	
Sulfur trioxide, SO ₃	2.88	
Phosphorus pentoxide, P ₂ O ₅	0.28	
Strontium oxide, SrO	0.01	
Barium oxide, BaO	0.12	
Manganese oxide, Mn ₂ O ₄	0.05	
Undetermined	<u>1.66</u>	
<u>Fouling Index</u>	0.47	
<u>Slagging Index</u>	0.97	

*NOTE: Heating value calculated using a correlation based on the coal ultimate analysis.

delivery cost for this coal is \$35 per ton (1990 dollars). The design limestone used in the study is a high calcium limestone from a quarry at Rome, Georgia. This is considered to be the most economical source of limestone for Plant Wansley. Dolomite is not readily available within an economic distance of the plant. An analysis of the Rome limestone is listed in Table 3-6.

3.2.3 Environmental Performance Criteria

Air Emissions

New Source Performance Standards (NSPS) for gasification power plants do not exist at the present time. However, it is anticipated that the EPA would use NSPS for coal-fired plants as a guideline for air emission limits for a gasification power plant. The NSPS air emission limits for these plants are summarized in Table 3-7.

The goals of the designs addressed in this study are to reduce particulates to well below 10 ppm and to remove greater than 95 percent of the sulfur, which will surpass the guidelines for both particulates and SO₂. Also, the plants are designed not to exceed the combustion turbine NSPS limit for NO_x of 75 ppmv at 15-percent oxygen. The NSPS limit of 75 ppm NO_x must now be considered an upper limit for NO_x levels for gas turbine power plants. State limits are often much more stringent than the NSPS requirement, and the trend is toward tighter control of NO_x emissions. At present, the baseline NO_x for combustion turbines in the state of Georgia is NSPS or uncontrolled emissions, whichever is less. From this baseline, emissions levels for specific plants would be considered on a case-by-case basis. Because of the lack of test cases at this time, it is not possible to predict NO_x emission requirements for a gasification-based power plant installed at Plant Wansley in future years.

The plants in the study were also specified to conform with the National Ambient Air Quality Standards (NAAQS) for NO_x, SO₂, and particulates. The Industrial Source Complex Short Term (ISCST) model was used to predict ambient air quality concentrations and help determine the impact of the gasification power plants on air quality. The monitoring significant impact levels, shown in Table 3-7, serve to determine whether preconstruction air monitoring for a period of one year is required at the plant site. This monitoring would be used to determine the existing air quality at the plant site.

The study criteria for air emissions are summarized in Table 3-8.

Table 3-6

Limestone Analysis

Source	Rome, Georgia, Quarry
Type	Florida Rock
<u>Composition, wt% dry basis</u>	
CaCO ₃	93.5
MgCO ₃	4.2
Inerts	2.3
	<u>100.0</u>
Bulk Density, lb/cu ft	90
Moisture, wt% max	3.0

Table 3-7

Air Emission Limits

NSPS Standards for Coal-Fired Plants

<u>Component</u>	<u>Limit</u>
SO ₂	1.2 lb/MMBtu
NO _x *	0.6 lb/MMBtu
Particulate	0.03 lb/MMBtu

* NOTE: NSPS NO_x limit for stationary combustion turbines is 75 ppmv at 15% O₂.

NAAQS Air Emission Limits (mg/m³)

	<u>Average Time</u>	<u>Significance Level</u>	<u>NAAQS</u>
SO ₂	3 hour	—	1300
	24 hour	13	365
	annual	—	80
NO _x	annual	14	100
Particulate	annual	10	150

Study Criteria for Air Emissions

SO ₂	> 95% Removal
NO _x	75 ppmv at 15% O ₂
Particulate	< 0.03 lb/MMBtu
All	Not to exceed NAAQS limits

Water Discharge

Discharge standards for the treated water are also based on New Source Performance Standards (NSPS) for coal-fired units, since standards for gasification-based plants are not formulated. The NSPS effluent discharge limitations for coal-fired plants are listed in Table 3-8.

Solid Waste

The solid wastes from the gasification island, chiefly ash with spent limestone, would be disposed of in a dry landfill at the plant site. A dry disposal system is required for the ash due to the presence of lime in the ash-laden material. Use of the onsite disposal facility is based upon all of the materials being nonhazardous as determined by the Resource Conservation and Recovery Act. The requirements of the Georgia Department of Natural Resources for the waste disposal facility are anticipated to include a liner, a leachate collection system, and a groundwater monitoring system.

Additional solids which would have to be disposed of for certain study cases are spent zinc ferrite used for flue gas desulfurization, and nahcolite used for removal of chlorides. Because of limited long-term experience with these materials in service, it is not known whether the spent solids would be classified as hazardous or nonhazardous for purposes of disposal. However, the amounts of these materials to be disposed of are relatively small, and the cost impact of disposal and handling methods is not considered significant for the present study. Therefore, the study assumes that these materials could be disposed of in the dry landfill along with the ash.

3.2.4 Other Plant Design Criteria

In other areas, the plant design basis for the study was to use as much as possible the same major design conditions and guidelines as were applied in the oxygen-blown IGCC reference case. The equivalent availability for the 400-MW-size plants was to be no less than 85 percent. For the smaller size plants of similar design, the availabilities were to be calculated and reported but were not to be less than 65 percent. The design point for the gasification islands was set to fully load the gas turbines at an ambient air temperature of 90°F. Whenever possible, existing facilities at the site were used as long as their use would not disturb the locations of future Units 3 and 4 at Plant Wansley.

Table 3-8

Effluent Discharge Limitations (NSPS)
All units in mg/L unless otherwise indicated

Parameter	Average <u>Monthly</u>	Monthly <u>Maximum</u>
pH units	6-9	--
Ammonia (N)	25	85
Cyanide (Total)	5	10
Sulfide	0.2	0.4
Chemical Oxygen Demand (COD)	150	--
Biological Oxygen Demand (BOD)	30	--
Oil and Grease	10	20
Total Suspended Solids (TSS)	25	50
Phenol	0.05	0.1
Benzene	--	0.05
Naphthalene	--	0.05

The design basis for the present study differed in several important ways from that used in the EPRI study. The EPRI IGCC plant was designed and evaluated with oil firing capability in order to provide supplemental power production at low ambient temperatures and improve overall plant availability. For the present study, the GE MS7001F gas turbine used as the basis for the IGCC plant designs was rated by GE according to the so-called "flat rating" concept, which has been developed by GE since the EPRI site-specific IGCC study was completed. Under this concept, there is very little change in gas turbine output with ambient temperature, and thus minimal need for supplemental firing with fuel oil. Moreover, system dispatch analyses conducted for the EPRI study indicated that very little oil backup fuel would be used due to the higher price of fuel oil relative to coal, and thus oil capability did not greatly increase plant availability. For these reasons, fuel oil firing capability was not provided for the cases evaluated in this study, although oil is used as a startup fuel.

With regard to sulfur removal, the EPRI Plant Wansley IGCC study assumed that there was no advantage in obtaining sulfur removals much above the NSPS limit of 90 percent, and therefore, a maximum sulfur removal of 95 percent was established. The addition of a COS hydrolysis step would be required in order to obtain removals beyond this level. For the present study, sulfur removals above 95 percent were readily attainable with the basic process flowsheets, and the criterion for sulfur removal was in the range of 95 to 99 percent.

3.3 Gasification Island Design Basis

The rationales for determination or selection of the gasification island design basis parameters for the study cases are discussed in the following paragraphs. Case-specific design basis information for gasification systems is provided in Table 3-9 for Cases 1 through 6. It should be noted that Case 7 was not carried to completion.

3.3.1 KRW Gasification/In-Bed Desulfurization

The principal design constraints for the gasifier design are vessel diameter, freeboard velocity, and temperature. The vessel diameter is limited to approximately 14 feet. This diameter allows road shipment of a shop-fabricated vessel and is within the limits of extrapolation of cold model fluidization data for a single-feed-tube gasifier. The freeboard velocity is limited to a value that will

Table 3-9 (Sheet 1 of 4)

Gasification Island Design Basis

Case Number	1	2	3	4	5	6
Nominal Plant Size, MW	400	100	200	400	400	400
<u>GASIFICATION</u>						
Number of Gasifiers	4	1	2	4	6	3
Operating Conditions:						
Pressure, psia	400	750	400	750	400	300
Temperature, °F	1900	1900	1900	1900	1850	1455
Ratios (wt/wt):						
Air/Coal	3.40	3.35	3.40	3.35	3.51	3.06
Steam/Coal	0.18	0.18	0.18	0.18	0.18	0.34
Recycle Gas/Coal	0.23	0.24	0.23	0.24	0.54	---
Transport Gas						
Coal	Air	Air	Air	Air	Air	Air
Limestone	Air	Air	Air	Air	N/A	Air
LASH*	R.G.	R.G.	R.G.	R.G.	R.G.	N/A
Coal Feedsize to Gasifier	-1/4 in.	-1/4 in.	-1/4 in.	-1/4 in.	-1/4 in.	50 µm
Limestone Feedsize,						
MMD, µm	65	650	650	650	650	60
Moisture in Coal, wt. %	5.0	5.0	5.0	5.0	5.0	5.0
Ca/S molar ratio	3.1	3.1	3.1	3.1	N/A	1.5
Freeboard Velocity, ft/s	2.61	1.4	2.61	1.4	1.6	N/A
Fluidized-bed density, lb/cu ft	39.0	39.0	39.0	39.0	17.1	N/A
Carbon in LASH, wt%						
(Sorbent-free/coal ash basis)	4.2	4.2	4.2	4.2	5.0	8.1
Number of Cyclones	4	1	2	4	6	6
Cyclone Efficiency, %						
Char	95	95	95	95	95	98
Sorbent	98	98	98	98	N/A	99

* LASH = Limestone-containing ash
R.G. = Recycle gas

Table 3-9 (Sheet 2 of 4)
Gasification Island Design Basis

Case Number	1	2	3	4	5	6
Nominal Plant Size, MW	400	100	200	400	400	400
<u>HEAT RECOVERY SYSTEM</u>						
Type	Fire-tube	None	Fire-tube	None	Fire-tube	Fire-tube
Number Required	4	N/A	2	N/A	6	3
Gas Inlet Temp., °F	1900	N/A	1900	N/A	1850	1455
Gas Outlet Temp., °F	1244	N/A	1244	N/A	1300*	1006
<u>GAS FILTERS</u>						
Type	Candle	Candle	Candle	Candle	Candle	Candle
Number Required	4	1	2	4	6	3
Operating Temp., °F	1015	1000	1015	1000	1000	1000
Operating Pressure, psia	380	735	380	735	380	290
Inlet Solids Loading, lb/lb gas	0.0008	0.0007	0.0008	0.0007	0.0010	0.0011
<u>CHLORIDE GUARDS</u>						
Sorbent Type	Nahcolite	Nahcolite	Nahcolite	Nahcolite	Nahcolite	N/A
Number Trains	4	1	2	4	6	N/A
Number Required	8	2	4	8	12	N/A
Operating Temperature	1007	1008	1007	1008	1003	N/A
Operating Pressure	365	715	365	715	365	N/A
Design Outlet Chloride - conc., in ppm	3	3	3	3	3	N/A

* Convective boiler upstream of recycle cyclone

Table 3-9 (Sheet 3 of 4)

Gasification Island Design Basis

Case Number	1	2	3	4	5	6
Nominal Plant Size, MW	400	100	200	400	400	400
<u>EXTERNAL DESULFURIZATION</u>						
No. of Trains Required	4	1	2	4	6	N/A
No. of Reactors	8	2	4	8	12	N/A
Sorbent Type	ZnFe	ZnFe	ZnFe	ZnFe	ZnFe	N/A
Inlet Temp., °F	1000	1000	1000	1000	1000	N/A
Approach to CO Shift						
Equilibrium, °F	0	0	0	0	0	N/A
Regeneration:						
Gas Type	Air-steam	Air-steam	Air-steam	Air-steam	Air-steam	N/A
Temperature, °F	950	840	950	840	950	N/A
Pressure, psia	450	850	450	850	450	N/A
Sulfur Loading, lb/lb sorbent	0.14	0.14	0.14	0.14	0.14	N/A
Space Velocity, hr ⁻¹	2000	1997	2000	1997	1315	N/A
No. of Cycles/Sorbent Batch	100	100	100	100	100	N/A
<u>RECYCLE GAS SYSTEM</u>						
Compressor Type*	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	N/A
Number Required (spare)	4 (2)	2 (1)	2 (1)	4 (2)	6 (2)	N/A
Compressor Efficiency, % (poly)	75	75	75	75	75	N/A
Operating Conditions:						
Inlet Temp., °F	330	360	330	360	330	N/A
Inlet Pressure, psia	305	675	305	675	300	N/A
Outlet Pressure, psia	450	780	450	780	430	N/A

* A single compressor (two-stage) is utilized for the recycle gas and the blowback gas systems. There will be an intercooler between the first and second compression stages (Cases 1 through 5).

Table 3-9 (Sheet 4 of 4)

Gasification Island Design Basis

Case Number	1	2	3	4	5	6
Nominal Plant Size, MW	400	100	200	400	400	400
<u>FILTER BLOW-BACK SYSTEM</u>						
Compressor Type	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal
Number Required (spare)	4 (2)	1 (1)	2 (1)	4 (2)	6 (2)	3 (1)
Operating Conditions:						
Inlet Temp, °F	330	360	330	360	330	300
Inlet Pressure, psia	448	778	448	778	428	280
Outlet Pressure, psia	750	1430	750	1430	750	700
<u>SULFATOR</u>						
Number Required	3	1	2	2	None	3
Operating Pressure, psia	Atmosph	Atmosph	Atmosph	Atmosph	N/A	290
CaS Conversion, %	90	90	90	90	N/A	99
Fines Combustion eff., %	98	98	98	98	N/A	100
Operating Temperature, °F	1600	1600	1600	1600	N/A	1650
Steam Pressure, psia	1600	1600	1600	1600	N/A	1600
Flue gas Temp.,* °F	1400	1400	1400	1400	N/A	1650

* Hot flue gas is routed to the coal drying section for Cases 1 through 4.

result in an optimal fines elutriation and recycle rate. The allowable freeboard velocity is higher for a gasifier utilizing limestone for sulfur removal. The presence of limestone results in a denser bed which suppresses elutriation and enhances the rate of fines reactivity. The temperature is limited by the ash softening characteristic of the coal and by the heat dissipation from the combustion jet. Since heat dissipation is enhanced by the denser bed, a higher gasifier temperature is allowed when limestone is used. The higher gasifier temperature enhances the gasifications rate.

For systems which use bulk in-bed desulfurization plus external zinc ferrite sulfur polishing, the percent sulfur removal per pass in the gasifier should be optimized to minimize system costs. The percent sulfur removal can be increased by increasing the limestone feed rate to raise the calcium-to-sulfur (Ca/S) ratio; however, a practical limit results from the cost of limestone supply, the capacity for Limestone Containing Ash (LASH) sulfation, and the cost of waste disposal. As the percent in-bed desulfurization is reduced, the duty of zinc ferrite polishing increases, requiring more sorbent, possibly a greater number of sorbent reactors, and a larger recycle stream of SO₂ and steam to the gasifier. Excessive amounts of SO₂ and steam require an increase in Ca/S ratio and cause a reduction in the gasifier coal throughput.

The relative impacts of these parameters on cost of electricity were estimated as illustrated in Figure 3-3. This optimization does not account for the effects of added gas moisturization to protect the zinc ferrite sorbent. Optimization of these parameters results in an optimum Ca/S ratio of 3:1 for the Rome limestone available near the Plant Wansley site. At these conditions, the per pass in-bed desulfurization is 88 percent. The strong interrelationship between the gasifier capacity, the in-bed desulfurization, and the zinc ferrite desulfurization results in economic incentive to use the highest reactivity sorbent available. Dolomitic limestone or dolomite may be more reactive than the calcium-rich limestone selected for the study. However, dolomitic limestone or dolomite is not available within an economic distance of Plant Wansley. Moreover, the use of dolomite requires the disposal of substantial quantities of magnesium compounds, which have relatively high solubilities in water and thus could have an adverse effect on water discharges from the landfill area.

For Case 5 an acid plant was specified to evaluate the tradeoff between in-bed desulfurization combined with sulfation versus a greater zinc ferrite desulfurization load combined with sulfur recovery as sulfuric acid byproduct.

Note: This optimization did not include the effects of zinc ferrite sorbent on the moisturization requirements of the fuel gas. This should be considered in future studies.

Figure 3-3 Optimization of Gasifier Sulfur Removal For KRW Gasification With Zinc Ferrite Sulfur Polishing

3.3.2 Transport Gasification With In-Bed Sulfur Removal

A high level of carbon conversion is projected for the transport gasifier. Gasifier conversion is controlled by maintaining the proper circulation rate in the system, thus controlling the gasifier bed density. The limestone sorbent completely calcines in the gasifier and removes most of the H_2S produced by the gasification reactions. The gas/solids mixture flows up through the gasifier at relatively high velocity and, after approximately 3-seconds residence time, exits the gasifier at the top.

Solids from the gasifier, consisting of char and partially sulfided sorbent, are fed directly to the pressurized circulating-bed sulfator. The gas/solids mixture flows up through the riser at high velocity and, as the char burns, sulfur is released as SO_2 . This is captured by the sorbent and converted to $CaSO_4$. Virtually complete SO_2 removal should approach 100 percent. The sulfided sorbent in the feed from the gasifier is also converted to $CaSO_4$. The exothermic heat of sulfation reaction and residual carbon combustion results in a relatively small increase in gas/solids temperature leaving the sulfator. The solids recirculation rate is set to control this temperature rise. Combustion of char in the sulfator is essentially complete. The air for the sulfator is also extracted from the booster air compressor. Since the sulfator operates at nearly the same pressure as the gasifier, flue gas produced in the sulfator is routed to the gas turbine for increased cycle efficiency.

3.3.3 Hot Particulate Removal

The particulate cleanup train, consisting of a nonrecycle cyclone and a gas filter, were selected to provide hot clean gas to the gas turbine. The gas filter selected is ceramic-candle type rather than a sintered-metal type. This selection was made based on pilot plant experience as well as various evaluations by Argonne National Laboratories and Westinghouse. The two major problems in using a sintered metal filter at the required 1000°F temperature are: (1) rapid sulfidation of the filter elements, which destroys the base metal, and (2) blockage of filter passages due to the formation of a liquid phase Ni/S eutectic.

For the KRW designs the particulate removal system includes a nonrecycle cyclone ahead of a ceramic filter for two reasons; (1) to accommodate typical fluctuations that occur in solids rates exiting the fines recycle system; and, (2) to reduce the solids loading to the filter to 1000 ppmw or less. The LASH removed from the bottom of the gasifier is mixed with the fine solids captured by the filter. Although not essential to operability, this mixing of solids in the lower portion of the filter vessel increases solids flowability and decreases the solids removal temperature, thus improving the system reliability. The solids from both the cyclone and the filter are depressurized by a lockhopper system with hot (1000°F) valves.

The hot gas filter was selected to be a ceramic-candle type consisting of 1.5-meter-long elements and designed for a face velocity of 6 ft/min. These criteria were based upon 650 hours of pilot plant testing (Reference 1).

3.3.4 Zinc Ferrite Desulfurization

Zinc ferrite was selected as the external bed sorbent for high-temperature coal gas desulfurization because of its effectiveness and capability for both sulfur absorption and sorbent regeneration. Based on significant bench- and pilot-scale testing, this sorbent appears to be the best choice at the present time, although current research indicates that better sorbents may be on the horizon.

Design of the zinc ferrite removal system is based on an expected average sulfur loading of the sorbent of 14 pounds of sulfur per 100 pounds of sorbent with the theoretical maximum being about 35 pounds. The expected life of the sorbent is 100 absorption and 100 regeneration cycles. The sizing of the sorbent bed is based upon a nominal 12-ft-diameter reactor (13 feet for Cases 1 and 3) with an initial space velocity of 2000 (standard cubic feet per hour of gas per cubic foot of zinc ferrite online in desulfurization mode). For Case 5, in which the zinc ferrite bed is used in the bulk sulfur removal mode, a lower initial space velocity of 1315 was used to yield a reasonable sorbent life. Support plates and appropriate charging and removal systems above them will be provided where the sorbent bed height exceeds 30 feet, which is the sorbent crush strength limit. All designs include two desulfurization vessels per gasifier train with desulfurization occurring in one vessel and regeneration being done alternately in the second vessel. The design data are summarized in Table 3-10. The above criteria are based upon experimental results from the pilot plant (Reference 3) and from other DOE-sponsored programs.

3.3.5 Chloride Removal

Turbine Protection

The turbine inlet specification for chloride (as HCl) as provided by General Electric is 30 ppm. The raw fuel gas is estimated to contain 60 ppm Cl resulting from a 15-percent release of the coal chloride to the fuel gas (Section 5.4). This concentration in the raw fuel gas will be diluted to approximately 23 ppm at the turbine inlet and as such would be acceptable to the turbine without chloride removal.

Table 3-10
Zinc Ferrite Design Data

SULFUR ABSORPTION CAPACITY = 14 WT%

<u>SULFIDATION:</u>	<u>CASES 1&3</u>	<u>CASES 2&4</u>	<u>CASE 5</u>
1. SPACE VELOCITY, (1/HR)	2000	1997	1315
2. SUPERFICIAL VELOCITY, FT/SEC. (OUTLET BASIS)	3.05	1.14	1.87
3. CYCLE TIME/BED,HOURS	172	170	38
<u>REGENERATION:</u>			
1. SPACE VELOCITY, (1/HR)	76	74	477
2. SUPERFICIAL VELOCITY, FT/SEC. (OUTLET BASIS)	0.08	0.04	0.52
3. CYCLE TIME/BED,HOURS	152	156	24
SORBENT LIFE, YEARS	3.9	3.9	0.9
SORBENT LIFE, YEARS @ 85% S.F.	4.6	4.6	1.0
NUMBER OF VESSELS/TRAIN	2	2	2
BED DIA. X LENGTH, FT. X FT.	13X36	12X30	12X36

Steam Generator Protection

Chloride in the fuel gas may react with the zinc ferrite sorbent and volatilization of zinc chloride would be likely to occur. The zinc chloride may be carried downstream and could deposit on cool metal surfaces of the heat recovery steam generator, leading to chloride stress corrosion cracking. Despite this potential problem, steam generator protection alone probably does not justify the added cost of a chloride guard.

Zinc Ferrite Protection

The primary reason to remove the chloride from the fuel gas is to prevent the likely degradation of the zinc ferrite sorbent by the reaction of chloride with zinc. This would significantly reduce the sulfur capture capacity of the sorbent thus requiring an additional \$40 MM to \$50 MM in zinc ferrite costs over the 40-year plant life. The cost of this incremental sorbent requirement is estimated to be two to three times the capital and operating cost of the chloride guard system. This cost tradeoff is based upon the expectation that 20 percent of the chloride in the fuel gas would react with the zinc ferrite. Figure 3-4 shows a sensitivity of chloride reaction to the associated zinc consumption; and Table 3-11 identifies the basis of the cost tradeoff.

Current or future experimentation may provide a basis for modifying this tradeoff. However, at present it is essential that all designs that include zinc ferrite also include a chloride guard.

3.3.6 Sulfation of Gasifier Solids

KRW coal gasification with in-bed desulfurization produces a solid waste for eventual landfill disposal. This material consists of a mixture of ash agglomerates, spent sorbent, and partially used sorbent. As discussed previously, when limestone is used as the sorbent, the waste material is called LASH. The calcium sulfide that is formed in the gasifier creates a concern because it can release hydrogen sulfide gas if it is exposed to an acid solution.

Investigations have been performed to explore various methods for converting the calcium sulfide to calcium sulfate, the intent being to achieve the greatest degree of conversion and thereby produce a solid waste meeting present and future landfill requirements.

The primary method identified for conversion of calcium sulfide to calcium sulfate is by roasting in air at a temperature high enough for rapid conversion but without excessive emissions of sulfur dioxide. This process is termed "sulfation" and is included in the gas island.

Figure 3-4 Zinc Ferrite Consumption Due to Chloride

Table 3-11

Chloride Guard Cost Tradeoff

ASSUMPTIONS

- 15% Cl⁻ in coal present as HCl in fuel gas
- 20% HCl reacts with Zn in ZnFe
- Plant capacity of 400 MWe (Case 1) with capacity factor of 85%
- ZnFe cost of \$6000/ton and Nahcolite cost of \$250/ton
- 10% discount rate to obtain present worth of replacement sorbent
- ZnFe and Nahcolite disposal cost of \$40/ton

<u>COMPARISON</u> (Millions of dollars)	<u>With</u> <u>Cl⁻ Guard</u>	<u>Without</u> <u>Cl⁻ Guard</u>
Chloride Removal Equipment Cost	9.5	--
Contingencies and Preoperational Cost	2.9	--
Allowance for Funds Used During Construction	2.3	--
Nahcolite Sorbent and Disposal Cost	4.0	--
ZnFe Consumption and Disposal Cost	<u>0</u>	<u>46.8</u>
	18.7	46.8

Pilot plant tests (performed for MWK by Dorr Oliver) using a continuous sulfation reactor showed a 20- to 30-percent conversion of the sulfide in LASH, a lower value than expected but sufficient to meet requirements for this nonhazardous waste.

Improvements in this roasting process are expected to yield a 90-percent conversion of sulfide in the LASH as shown in the design basis table. A likely evolution in sulfation process improvement is to progress from the atmospheric fluidized-bed sulfator (as is now the reference design for the KRW technology) to an atmospheric circulating bed and finally to a pressurized transport bed as included in Case 6.

Additionally, sodium carbonate additive has been successfully used in the laboratory to effect a 90-percent conversion of the sulfide in the LASH.

3.4 POWER ISLAND/BALANCE OF PLANT DESIGN BASIS

Two different power island configurations were analyzed as part of this study: (1) combined-cycle plants and (2) intercooled steam injected gas turbine (ISTIG) plants. In both cases, the power island designs were highly integrated with the gasifier island designs in order to maximize efficiencies and minimize costs. The gas turbines were to be fully loaded at an ambient air temperature of 90°F.

The power island and balance-of-plant systems for each plant design were designed to accept the fuel gas from the gasification island and to interchange boiler feedwater, steam, and auxiliary flows with the gasification island in the most efficient manner possible. Therefore, the design basis for the power island consisted primarily of interface parameters generated iteratively from gasification island performance data.

Basic power island configuration data for the first six study cases are given in Table 3-12. Following are brief descriptions of the major technologies selected for each type of power island design.

3.4.1 Combined-Cycle Plants

A combined-cycle unit utilizes a gas turbine in combination with a steam turbine to produce electrical power. A heat recovery steam generator (HRSG) is used to recover the heat from the gas turbine exhaust and produce steam that is used in the steam turbine. Balance-of-plant equipment is required to support the operation of the gas turbine, HRSG, and steam turbine. The balance-of-plant equipment includes a condenser/cooling tower system, deaeration system, HRSG feedpumps, water treatment system, electrical distribution system, and miscellaneous mechanical and electrical support systems.

Table 3-12

Power Island Major Equipment Design Basis

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Combustion Turbines						
Type (all GE)	MS7001F	LM8000	MS7001F	LM8000	MS7001F	MS7001F
Number	2	1	1	4	2	2
Firing Temp., °F	2300	2500	2300	2500	2300	2300
Inlet Press., psia	295	650	295	650	290	280
HRSG's						
Number	2	1	1	4	2	2
Drum Pressure Levels						
High, psia	1550	900	1550	900	1550	1550
Intermed., psia	350	250	350	250	350	350
Deaerator Press., psia	25	25	25	25	25	25
Finch Point, °F	35	35	35	35	35	35
Flue Gas Stack Temp., °F	280	*	280	*	280	280
Steam Turbines						
Type						
Number	1	N/A	1	N/A	1	1
Throttle Press., psia	1450	N/A	1450	N/A	1450	1450
Throttle Temp., °F	1000	N/A	1000	N/A	1000	1000
Reheat Press., psia	350	N/A	350	N/A	350	350
Reheat Temp., °F	1000	N/A	1000	N/A	1000	1000
Exhaust Press., in. Hg	3.5	N/A	3.5	N/A	3.5	3.5

* Minimum flue gas stack temperature of 280°F was specified, but actual exhaust temperature was much higher due to cycle limitations.
See Section 5.

The GE MS7001F advanced gas turbine was selected as the gas turbine model to based on its potential for high efficiency operation and its advanced state of commercial development.

The equipment and configuration selected for the combined-cycle plant designs represent generally accepted engineering practice for combined-cycle plants, adapted as appropriate for integration with the gasification island. One assumption which should be pointed out is the use of a separate HRSG for each combustion turbine. This approach allows the use of smaller, shop-fabricated HRSGs with a consequent reduction in cost. On the other hand, a single large HRSG is often specified because of concerns about accurately splitting and recombining steam flows to and from a single steam turbine. SCS has determined that the control of steam flows in such circumstances has been successfully demonstrated and can be utilized with minimum risk in combined-cycle installations. Therefore, this assumption has been used in all study cases where applicable.

3.4.2 ISTIG Plants

The ISTIG unit utilizes an aircraft-derivative gas turbine that has the capability to accept large quantities of steam within the turbine section to augment power generation. A heat recovery steam generator (HRSG) is used to recover the heat from the gas turbine exhaust and produce the steam for injection into the gas turbine at various pressure levels. Balance-of-plant support equipment for the ISTIG configuration includes HRSG feedpumps, a deaeration system, an intercooling system for the gas turbine, water treatment facilities, and miscellaneous mechanical and electrical support equipment.

The proposed General Electric LM-8000 ISTIG design was chosen as the center of the ISTIG power island design based on its estimated ability to provide the required 100 MW power level, on its anticipated high cycle efficiency, and on past studies that have used the LM-8000 as the basis of a coal-gas-fueled power unit. The LM-8000 ISTIG is a theoretical design only as opposed to the GE MS7001F design used in cases 1, 3, 5, and 6, which is based on an actual operating design. Therefore, the ISTIG designs and performance developed in this study should be considered more conceptual in nature.

4.0 PLANT DESCRIPTIONS -- IGCC CASES 1, 3, 5, and 6

4.1 INTRODUCTION

The IGCC system configurations for Cases 1, 3, 5, and 6 are similar in that all four cases have the same combined-cycle power island configuration. The four cases differ in either plant size or design of the gas island. The differences in the gas island design are in the desulfurization method or the type of gasifier used for the plant design. Cases 1, 3, and 5 use the KRW gasifier and Case 6 uses the transport gasifier.

Section 4 provides detailed descriptions of the design and performance of the four IGCC study cases, including:

- Process descriptions
- Material balances
- Plant layouts and equipment arrangements
- RAM analysis results
- Plant performance

Cost and economic data for the designs are presented in subsequent sections.

4.2 AIR-BLOWN IGCC WITH KRW GASIFICATION, BULK SULFUR REMOVAL IN THE GASIFIER (CASES 1 AND 3)

4.2.1 Process Description

Cases 1 and 3 both utilize bulk sulfur removal with limestone in the KRW gasifier, with a zinc ferrite polishing step for additional sulfur removal. The main difference between the two cases is the plant size. Case 1 was specified as a 400-MW plant and Case 3 is half that size at nominally 200 MW. Both cases have essentially the same process steps with only the number of process trains varying for each case. Both cases are thus grouped together in the discussion that follows.

4.2.1.1 Overall Process Arrangement

The block flow diagrams for Cases 1 and 3 are shown in Figures 4-1 and 4-2, respectively. Case 1 has four separate processing trains for the gas island with two combustion turbines and heat recovery steam generators (HRSGs) and one steam turbine for the power island. Case 3, at half the size, has only two processing trains for the gas island and a single combustion turbine, HRSG, and steam turbine power island train. The corresponding stream flows for the stream numbers indicated on Figures 4-1 and 4-2 are listed in Table 4-1 and 4-2, respectively.

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-1 Case 1 Block Flow Diagram

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-2 Case 3 Block Flow Diagram

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 4-1 (Sheet 1 of 4) Case 1 Process Stream Flows
Note: All Streams Not Shown Are Proprietary to M.W. Kellogg

STREAM NUMBER	1	2	4	4A	10	14
DESCRIPTION	PREPARED COAL FEEDSTOCK	LIMESTONE FEEDSTOCK	TOTAL AIR FROM POWER ISLAND	AIR TO GT COMPRESSOR	STEAM TO GASIFIER	HIGH TEMP. BW FROM POWER ISLAND
	WT%	WT%	VOL%	VOL%	VOL%	WT%
GASES, LB/HR						
CO	28,010		0.00	0.00		0
H ₂	2,016		0.00	0.00		0
CO ₂	44,010		0.00	0.00		0
CH ₄	10,043		0.00	0.00		0
N ₂	28,013		0.00	3,702,002		0
Ar	30,048		0.00	64,000		0
O ₂	31,000		0.00	1,121,000		0
NH ₃	17,031		0.00	0.00		0
H ₂ S	34,000		0.00	0.00		0
CO _S	60,075		0.00	0.00		0
SO ₂	64,003		0.00	0.00		0
H ₂ O	18,015		0.00	0.00	10,200	0
HCl	30,461		0.00	0.00	0	0
TOTAL GASES			100.00	4,883,002	100.00	10,200
GAS FLOW, LB MOLE/HR*						
MOLECULAR WEIGHT			37,084	168,127	185,540	
GAS VOLUME, ACFM			23,306	1,150,857	220	
GAS VOLUME, BCFM			40,418	1,011,943	3,015	
LIQUIDS, LB/HR			234,511			812,273
WATER	18,015					
SOLIDS, LB/HR						
COAL	80.00	233,311				
LIMESTONE	10.34	32,004				
ASH	100.00	315,975				
TOTAL SOLIDS			1,073,955	4,804,002	185,540	812,273
TOTAL FLOW, LB/HR						
TEMPERATURE, F	200	90	700	90	900	900
PRESSURE, PSIA	14.0	14.0	200.0	14.7	450.0	1,800

Table 4-1 (Sheet 2 of 4) Case 1 Process Stream Flows

STREAM NUMBER	16	24	30	32	38	44	47
DESCRIPTION	GASIFIER STEAM EXPORT	SOLIDS TO DISPOSAL	STEAM TO ZnF ₂ REACTORS	PRODUCT GAS TO GAS TURBINE	PRODUCT GAS QUENCH STEAM	FINES LOCKHOPPER VENT GAS	SULFATOR STEAM TO POWER ISLAND
	VOL% LB/MOHR	WT%	VOL%	VOL%	VOL%	VOL%	VOL%
GASES, LB/MR							
CO	28,010			7.77	173,304	7.77	118
H ₂	2,016			20.63	33,126	20.64	23
CO ₂	44,010			14.39	804,416	14.39	348
CH ₄	16,043			0.64	8,241	0.64	6
N ₂	28,013			28.87	818,345	28.87	683
Ar	39,848			0.45	14,178	0.45	10
O ₂	31,808			0.00	0	0.00	0
NH ₃	17,031			0.01	203	0.01	0
H ₂ S	34,080			0.00	66	0.00	0
CO _S	60,075			0.00	0	0.00	0
SO ₂	64,063			0.00	0	0.00	0
H ₂ O	18,015		100.00	19.44	279,038	19.43	103
HCl	38,481			0.00	0	0.00	0
TOTAL GASES	100.00 883,853		100.00	100.00	389,708	100.00	153,315
GAS FLOW, LB MOL/HR*							
MOLECULAR WEIGHT	49,088			70,974	21,832	86	153,315
GAS VOLUME, ACFM	18,02			18,02	18,02	22,88	18,02
GAS VOLUME, SCFM	839			1,442	8,245	870	1,012
GAS VOLUME, SCFM	310,286			16,228	136,788	347	53,818
LIQUIDS, LB/MR							
WATER	18,015						
SOLIDS, LB/MR							
COAL		1.0					
LIMESTONE		87.2					
ASH		31.8					
TOTAL SOLIDS		100.0					
TOTAL FLOW, LB/MR	883,853	102,578	46,389	1,830,918	389,707	1,261	153,315
TEMPERATURE, F	805	150	950	1,000	500	524	805
PRESSURE, PSIA	1,800.0	14.0	450.0	295.0	450.0	14.0	1,800.0

*Final fuel gas composition based on 98 percent removal of HCl across chloride guard and 95 percent removal of H₂S across zinc ferrite polishing system.

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Cleanup Based Advanced Gas Turbine Systems

Table 4-1 (Sheet 3 of 4) Case 1 Process Stream Flows

STREAM NUMBER	48	50	51	52	53	54	55
DESCRIPTION	SULFATOR FLUE GAS	PREHEATED BFW TO POWER ISLAND	TOTAL FLOW DOWN TO POWER ISLAND	EXHAUST GAS TO HRSG	TOTAL DEAER BFW FROM POWER ISLAND	FLUE GAS TO MAIN STACK	BFW TO HRSG
	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %
GASES, LB/HR							
CO	28,010	0					
H ₂	2,018	0					
CO ₂	44,010	28,979					
CH ₄	18,043	0					
N ₂	28,013	147,884					
Ar	28,943	1,01	2,585				
O ₂	31,988	3,08	7,530				
NH ₃	17,031	0	0				
H ₂ S	34,080	0	0				
CO ₈	60,075	0	0				
SO ₂	64,063	0	0				
H ₂ O	18,015	3,02	3,472				
HCl	38,461	0	0				
TOTAL GASES	188,548			100.00	9,735,801	100.00	9,735,111
GAS FLOW, LB MOL/HR							
MOLECULAR WEIGHT	6.387						
GAS VOLUME, ACFM	28.82						
GAS VOLUME, SCFM	147,405						
LIQUIDS, LB/HR							
WATER	18,015	280,719	10,373.00		486,088		1,518,828
SOLIDS, LB/HR							
COAL	0.86						
LIMESTONE	67.32						
ASH	31.70						
TOTAL SOLIDS	100.00	205					
TOTAL FLOW, LB/HR	188,753	280,719	10,373.00	9,735,801	486,088	9,735,111	1,518,828
TEMPERATURE, F	1,400	425	605	1,008	240	290	121
PRESSURE, PSIA	14.0	1,600	1,600	16	1,600	15.0	1.7

Table 4-1 (Sheet 4 of 4) Case 1 Process Stream Flows

STREAM NUMBER	56	57A	57B	58	59	60	61
DESCRIPTION	BPW MAKEUP	HP BOILER BLOWDOWN	IP BOILER BLOWDOWN	COOLING TOWER MAKEUP	COOLING TOWER BLOWDOWN EVAP. DRIFT	DILUTION AIR	FLUE GAS TO LOCAL STACK
	VOL%	VOL%	VOL%	VOL%	VOL%	VOL%	VOL%
GASES, LB/MR							
CO	28.010						
H ₂	2.016						
CO ₂	44.010						
CH ₄	18.043						
N ₂	28.013						
Ar	39.948						
O ₂	31.899						
NH ₃	17.931						
H ₂ S	34.080						
CO ₈	60.078						
SO ₂	64.063						
H ₂ O	18.015						
HCl	36.461						
TOTAL GASES						100.00	100.00
GAS FLOW, LB/MOL/MR						218,005	210,511
MOLECULAR WEIGHT							
GAS VOLUME, ACFM							
GAS VOLUME, SCFM							
LIQUIDS, LB/MR							
WATER	480.924	13,258	1,309	3,485,164	3,485,164		
SOLIDS, LB/MR							
COAL							
LIMESTONE							
ASH							
TOTAL SOLIDS							
TOTAL FLOW, LB/MR	480,924	13,258	1,309	3,485,164	3,485,164	218,005	210,511
TEMPERATURE, F	80	605	429	80			290
PRESSURE, PSIA	14.7	1,600	339.5	14.7	14.7		15.5

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 4-2 (Sheet 1 of 4) Case 3 Process Stream Flows
Note: All Streams Not Shown Are Proprietary to M.W. Kellogg

STREAM NUMBER	1	2	4	4A	10	14
DESCRIPTION	PREPARED COAL FEEDSTOCK	LIMESTONE FEEDSTOCK	TOTAL AIR FROM POWER ISLAND	AIR TO GT COMPRESSOR	STEAM TO GASIFIER	HIGH TEMP. BFW FROM POWER ISLAND
	WT%	WT%	VOL%	VOL%	VOL%	WT%
GASES, LB/HR						
CO	28,010		2.00	0.00	0	
H ₂	2,816		0.00	0.00	0	
CO ₂	44,010		0.00	0.00	0	
CH ₄	10,043		0.00	0.00	0	
N ₂	28,013		78.14	406,331		
Ar	30,948		0.00	7,004		
O ₂	31,900		20.00	124,004		
NH ₃	17,031		0.00	0.00	0	
H ₂ S	34,000		0.00	0.00	0	
CO _S	60,075		0.00	0.00	0	
SO ₂	64,063		0.00	0.00	0	
H ₂ O	18,015		0.00	0.00	100.00	5,150
HCl	36,461		0.00	0.00	0	
TOTAL GASES			100.00	536,979	100.00	5,150
GAS FLOW, LB MOL/HR			18,542	188,127	238	
MOLECULAR WEIGHT			28.96	28.96	18.02	
GAS VOLUME, ACFM			20,208	1,130,367	100	
GAS VOLUME, SCFM			117,358	1,011,943	1,808	
LIQUIDS, LB/HR						
WATER	18,015					100.00
SOLIDS, LB/HR						406,137
COAL	88.66	141,858				
LIMESTONE	10.34	16,332				
ASH	100.00	78,994				
TOTAL SOLIDS						
	157,988	43,867	636,979	4,804,842	5,150	406,137
TOTAL FLOW, LB/HR						
TEMPERATURE, F	200	90	700	90	950	800
PRESSURE, PSIA	14.0	14.0	200	14.7	450	1,650.0

Table 4-2 (Sheet 2 of 4) Case 3 Process Stream Flows

STREAM NUMBER	16	24	30	32	38	44	47
DESCRIPTION	GASIFIER STEAM EXPORT	SOLIDS TO DISPOSAL	STEAM TO ZnF ₂ REACTOR	PRODUCT GAS TO GAS TURBINE	PRODUCT GAS QUENCH STEA	FINES LOCKHOPPER VENT GAS	SULFATOR STEAM TO POWER ISLAND
GASES, LB/HR							
CO	25,010			7.77	55,652	7.77	60
H ₂	2,016			20.63	16,567	20.64	11
CO ₂	44,010			14.36	252,208	14.36	174
CH ₄	16,043			0.84	4,121	0.84	3
N ₂	28,013			36.67	408,172	36.67	252
Ar	38,943			0.46	7,063	0.46	5
O ₂	31,899			0.00	0	0.00	0
NH ₃	17,931			0.91	101	0.91	0
H ₂ S	34,080			0.00	27	0.00	0
CO ₂	62,075			0.00	6	0.00	0
SO ₂	94,063			0.00	0	0.00	0
H ₂ O	18,015		100.00	19.44	139,519	19.43	98
HCl	36,461			0.00	0	0.00	0
TOTAL GASES	100.00		100.00	100.00	100.00	100.00	100.00
GAS FLOW, LB MOL/HR	441,978		23,300	915,462	194,854	632	78,653
MOLECULAR WEIGHT	24,534		1,258	39,537	10,816	28	4,254
GAS VOLUME, ACFM	18,92		18,02	22,96	18,02	22.96	18,02
GAS VOLUME, SCFM	639		721	35,227	4,123	335	508
LIQUIDS, LB/HR	155,148		9,144	251,907	68,400	174	28,908
WATER							
SOLIDS, LB/HR							
COAL		1.0					
LIMESTONE		87.2					
ASH		31.8					
TOTAL SOLIDS		100.0					
TOTAL FLOW, LB/HR	441,977	51,268	23,300	915,462	194,854	631	78,653
TEMPERATURE, F	605	150	950	1,000	500	524	605
PRESSURE, PSIA	1,600.0	14.0	450.0	295	450	14.0	1,600

* Final fuel gas composition based on 92 percent removal of HCl across chloride guard and 95 percent removal of H₂S across zinc ferrite polishing.

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Table 4-2 (Sheet 3 of 4) Case 3 Process Stream Flows

STREAM NUMBER	40	50	51	52	53	54	55
DESCRIPTION	SULFATOR FLUE GAS	PREHEATED BFW TO POWER ISLAND	TOTAL BLOWDOWN TO POWER ISLAND	EXHAUST GAS TO HRSG	TOTAL DEASER BFW FROM POWER ISLAND	FLUE GAS TO MAIN STACK	BFW TO HRSG
	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %
GASES, LB/HR							
CO	28.010	0.00	0.00	0.00	0.00	0.00	0.00
H ₂	2.918	0.00	0.00	0.00	0.00	0.00	0.00
CO ₂	44.010	0.00	0.00	0.00	0.00	0.00	0.00
CH ₄	18.043	0.00	0.00	0.00	0.00	0.00	0.00
N ₂	28.013	82.67	73.042	82.67	73.042	82.67	73.042
Ar	39.948	1.91	1.282	1.91	1.282	1.91	1.282
O ₂	31.999	3.68	2.788	3.68	2.788	3.68	2.788
NH ₃	17.031	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ S	34.080	0.00	0.00	0.00	0.00	0.00	0.00
CO ₂	60.075	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	64.063	0.02	0.02	0.02	0.02	0.02	0.02
H ₂ O	18.015	3.62	1.729	3.62	1.729	3.62	1.729
HCl	36.461	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL GASES	84.573	100.00	84.573	100.00	3,367.801	100.00	3,367.801
GAS FLOW, LB MOL/HR	3,184						
MOLECULAR WEIGHT	29.52						
GAS VOLUME, ACFM	73.763						
GAS VOLUME, SCFM	20,194						
LIQUIDS, LB/HR		100.00	125.360	100.00	5,106.00		763.512
WATER	18.015						
SOLIDS, LB/HR							
COAL	0.00						
LMESTONE	67.32						
ASH	31.70						
TOTAL SOLIDS	100.00	100.00	100.00	100.00	243.044	100.00	763.512
TOTAL FLOW, LB/HR	84.577	125.360	5,106.00	3,367.801	243.044	3,367.801	763.512
TEMPERATURE, F	1,400	425	605.00	1,000	240	201	121
PRESSURE, PSIA	14.6	1,800.0	1,800	10	1,800	15.0	1.7

Table 4-2 (Sheet 4 of 4) Case 3 Process Stream Flows

STREAM NUMBER	66	67A	67B	68	69	60	61
DESCRIPTION	BFW MAKEUP	HP BOILER BLOWDOWN	HP BOILER BLOWDOWN	COOLING TOWER MAKEUP	COOLING TOWER BLOWDOWN EVAP. DRAFT	DILUTION AIR	FLUE GAS TO LOCAL STACK
	VOL%	VOL%	VOL%	VOL%	VOL%	VOL%	VOL%
GASES, LB/HR							
CO	23,010						
H ₂	2,018						
CO ₂	44,010						
CH ₄	18,043						
N ₂	23,013						
Ar	20,942						
O ₂	31,909						
NH ₃	17,031						
H ₂ S	34,000						
CO ₂	60,575						
SO ₂	64,063						
H ₂ O	18,015						
HCl	30,461						
TOTAL GASES							
GAS FLOW, LB MOL/HR							
MOLECULAR WEIGHT							
GAS VOLUME, ACFM							
GAS VOLUME, SCFM							
LIQUIDS, LB/HR	100.00 230,482	100.00 6,721	100.00 503	100.00 1,742,582	100.00 1,742,582	100.00 100,224	100.00 100,224
WATER							
SOLIDS, LB/HR							
COAL							
LIMESTONE							
ASH							
TOTAL SOLIDS							
TOTAL FLOW, LB/HR	230,482	6,721	503	1,742,582	1,742,582	100,224	100,224
TEMPERATURE, F	80	805	539	80	14.7		60
PRESSURE, PSIA	14.7	1,000	330.5	14.7	14.7		15.5

For both Cases 1 and 3, the coal and limestone are sized and dried prior to being fed into the KRW gasifier. Flue gas from the sulfator is used to dry the feed coal, whereas an oil-fired heater is used to dry the limestone. The product gas from the gasifier is cooled to nominally 1,000°F prior to entering a cyclone, which removes the coarse particulates from the product gas, and a high-temperature ceramic filter, which removes fine particulates. Both the coarse and fine particulates are fed to the sulfator for the conversion of calcium sulfide to calcium sulfate in the limestone and ash (LASH) from the gasifier. The limestone in the gasifier serves as the bulk desulfurization step of the process.

After hot gas filtration of particulates, the product gas passes through a chloride guard bed prior to entering the zinc ferrite polishing system for removal of additional sulfur from the product gas. The chloride guard is essential for removing chloride from the fuel gas, since chloride can severely affect the structural integrity of the zinc ferrite sorbent. After the exothermic zinc ferrite step, the clean product gas is cooled to 1,000°F prior to entering the power island.

The cleaned product gas from the gas island is combusted in either one (Case 3) or two (Case 1) GE MS7001F gas turbines to produce the majority of the power for the power island. Air for the gasifier is initially compressed to 200 psia in the gas turbine compressor, followed by additional pressurization in a booster compressor to deliver air at the proper pressure for the gasifier. The 1,100°F flue gas from the gas turbine is sent to a HRSG (one for each gas turbine) to produce steam for additional power produced in a single steam turbine, as well as steam for the gas island. The flue gas from the HRSG is sent to the main power plant stack.

4.2.1.2 Coal and Limestone Receiving, Handling, and Preparation

The coal handling systems for both Cases 1 and 3 use the existing unit train coal unloading system at the Plant Wansley site. Figure 4-3 and Figure 4-4 are process flow diagrams of the coal and limestone preparation systems for Cases 1 and 3. Twenty-eight new vibrating feeders are installed in the existing reclaim tunnel under the unloading trestle. All equipment beyond the vibrating feeders is dedicated to the IGCC plant. After unloading, the coal is passed by a magnetic separator as it is conveyed by six conveyors to four coal storage silos for Case 1 and three coal storage silos for Case 3.

Next, the coal is simultaneously crushed and dried in fluidized-bed roller mills. Case 1 has three mills with a fourth serving as a spare. Case 3 has a total of two operating pulverizers with one spare. The heat for coal drying is provided by flue gas from the sulfator in the gasifier. The flue gas at 1,400°F is first cooled to a temperature suitable for coal drying by mixing with recycle gas from the pulverizer/dryer along with a quantity of air, resulting in a drying gas with a

Figure 4-3 Case 1 Coal and Limestone Preparation Process Flow Diagram

maximum of 12-percent oxygen by volume at a temperature of 500°F. For startup, an oil-fired heater provides the heat for coal drying.

After preparation, the coal is transported to coal and limestone bunkers and then fed with weigh feeders into bins where the coal and limestone are combined. Pneumatic conveying systems are then used to transport the coal/limestone mixture to surge bins located in the gasifier structures. In order to increase reliability/availability, each of the weighing and pneumatic systems is capable of filling any of the surge bins. Case 1 has a total of four pneumatic conveying systems, one of which is a spare. Case 3 has a total of two operating pneumatic conveying systems for the coal/limestone mixture, one serving as a spare.

The limestone system for both cases uses an existing railroad track adjacent to the IGCC plant site. The rail car system consists of a covered single car receiving hopper with a car shaker and two belt feeders. The belt feeders discharge onto a conveyor, which forms a 50-foot-high pile adjacent to a 60-day storage pile. A dust suppression system is used at the receiving hopper and the stockout conveyor. The limestone reclaim system has two reclaim hoppers located under the live storage pile. A slide gate and a belt feeder are located at the bottom of each hopper. The belt feeders discharge onto a conveyor which transports the limestone to two storage silos.

The limestone from the storage silos is fed into two pulverizers for Case 1 and one pulverizer for Case 3, where it is dried with the heat supplied by an oil-fired heater. The pulverizers are operated two shifts per day with the third shift used for maintenance. Limestone from the pulverizers is collected in a storage silo with a storage capacity sufficient for 24 hours of operation. Two pneumatic conveying systems (one spare) transport the limestone to the common coal/limestone bunker. From this point, the coal and limestone share a common transport system to the surge bin located in the gasifier structure.

4.2.1.3 Air Compression and Supply

This system consists of two subsystems: the boost air-compression system and the lockhopper air-compression system. The subsystems are described separately below.

Boost Air Compression (Area 250)

The boost air-compression system provides the air necessary for gasifier operation and for regeneration of the zinc-ferrite desulfurizer. (Details of this system are provided in Subsection 4.2.1.6.) Air for this system is obtained by extraction from the compressor of the gas turbine located in the power island.

Air from the gas turbine compressor discharge is first cooled in the air recuperator. This exchanger simultaneously cools the relatively high-temperature air obtained from the gas turbine and heats the compressor discharge air before delivery to the gasifier. After the initial air cooling is accomplished, a boiler feedwater (BFW) preheater and a trim cooler are employed to cool the air to a lower temperature for more efficient air compression.

Before entering the compressor, a knockout drum is provided at the compressor suction to remove condensed moisture.

Lockhopper Air Compression

This subsystem is included as part of Area 300, Gasification. A separate air compressor is provided for lockhopper pressurization, as shown in Figure 4-5. The compressor is sized for the average flow required by the lockhoppers. To accommodate the large periodic flows to the lockhoppers, a surge vessel is included.

4.2.1.4 Coal Gasification and Recycle Gas Compression

This area consists of five functional subsystems: coal/sorbent pressurization, gasification, ash removal, heat recovery, and recycle gas compression. The subsystems are described separately below.

Coal/Sorbent Pressurization

Figure 4-5 is a process flow diagram of the gasification step for Cases 1 and 3. A conventional lockhopper system is used to pressurize the coal and limestone mixture. The two feedstocks are mixed in a single surge bin in the coal and limestone preparation steps. This bin operates at atmospheric pressure and serves as a reserve of raw material for the pressurizing lockhoppers located beneath the bin.

The feed-pressurization lockhopper periodically receives coal from the feed surge bin. Two valves are used between the vessels: the upper valve to initiate or stop solids flow and the lower to provide a gas-tight seal. After a batch of coal is received, the feed pressurization lockhopper is pressurized with high-pressure air. When the pressure in the lockhopper is equalized with the feed hopper, a valve located beneath the pressurization hopper is opened to dump the material into the feed hopper. When the lockhopper is empty, the valve is closed and the lockhopper is vented to atmospheric pressure to repeat the cycle.

301J	302F	310L	310F	311F
PRESSURIZATION	PRESSURIZATION	COAL FEED	FEED SURGE DRUM	FEED
AIR COMPRESSOR	AIR RECEIVER	VENT FILTER		PRESSURIZATION
				ROPPER

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Figure 4-5 Cases 1 and 3 Gasification
Process Flow Diagram

The vent filter is installed on the pressurizing/venting line to essentially eliminate particulate emissions when the pressurization hopper is vented. The lockhopper is pressurized through this same filter to provide an automatic backflushing of the filter media.

The feed hopper is operated at a constant pressure sufficient to permit starwheel feed metering and pneumatic conveying of the coal/limestone mixture to the gasifier by an air stream. The relative proportions of coal and limestone sent to the surge bin will be established by setpoint adjustments based on an analysis of the gasifier product gas obtained from an online sulfur analyzer.

Gasification

The coal is gasified in a KRW pressurized-fluidized-bed gasifier. The gasifier is a refractory-lined carbon steel pressure vessel, divided into a number of functional zones, where the necessary processes of coal devolatilization, combustion, gasification, desulfurization, and ash and spent sorbent cooling occur.

Coal and limestone are pneumatically fed to the gasifier through a central feed tube. Additional air is also fed through this same feed tube and the streams merge to form a central jet where the coal is quickly devolatilized, with the remaining char and limestone entering the bed. Combustion of char and gas occurs within the jet to provide the heat necessary for endothermic devolatilization, gasification, calcination, and desulfurization chemical reactions.

Within the bed a number of competing reactions occur. The principal reactions of interest are the gasification reactions:

These reactions occur between the gas-phase steam and combustion products and the carbon in the char. Carbon monoxide and hydrogen form the major combustible constituents of the product gas. Methane and other hydrocarbons are produced in lesser quantities, primarily from the devolatilization process. The operating temperature of the gasifier is sufficiently high to crack any tars or oils that might be produced.

The fluidized bed also presents a favorable environment for desulfurization of the gas. Hydrogen sulfide, produced from the sulfur in the coal, is the primary sulfur species and is removed from the gas phase by the following reaction:

Calcium oxide is obtained by the calcination or decomposition of limestone according to:

This reaction occurs very quickly when the limestone reaches the bed due to the low CO_2 partial pressure in the gasifier. As the carbon in the char is consumed, the particles become enriched in ash. The ash particles then tend to adhere to each other and agglomerate and, along with the dense calcium sulfide/oxide particles, separate from the char bed because of their different fluidization characteristics. This separation occurs primarily in a region that surrounds the central feed tube at the bottom of the gasifier vessel.

Gas leaves the gasifier at the top of the vessel and enters a cyclone. The cyclone separates the majority of the fine particulates that escape the gasifier bed and returns them to the bed by a fluidized dipleg.

The startup heater is a natural-gas-fired heater that provides hot air necessary for the initial heatup and combustion in the gasifier. The heater is designed to provide 1000°F air during the startup phase. This is a standard industrial combustor, with all necessary auxiliaries included.

Ash Removal

Ash and spent sorbent are removed from the gasifier by the ash feeder. The material is then conveyed from the gasifier area to the bottom of the gas filter where it is mixed with the fines collected by a filter.

Heat Recovery

Heat recovery is accomplished in two exchangers, the product gas cooler and the recycle gas cooler. Figure 4-6 is a process flow diagram of the heat recovery step for Cases 1 and 3. The product gas from the gasification area is partially cooled by the product gas cooler. Gas exiting the cooler is additionally cooled by steam/water quench. The amount of quench is determined by the desired moisture content of the gas, and desired mixed steam temperature is controlled by an internal bypass of the gas cooler.

The second exchanger, the recycle gas cooler, is used to cool the gas that is recycled to the gasifier and the gas that is used for blowback of the gas filter. The outlet gas temperature of the recycle gas cooler is controlled by the flow of boiler feedwater to the cooler.

360C
PRODUCT GAS
COOLER

360F
STEAM DRUM

580F
RECYCLE GAS
RECEIVER

NOTES:
1. THE
CASE
2. THE
AVER
3. FOR
SUAL
TREA

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CASE 1 CASE 3
 HEAT EXCHANGER
 T-100
 P-15.4-0
 P-15.5-0
 P-15.3-0

**Figure 4-6 Cases 1 and 3 Heat Recovery
Process Flow Diagram**

Recycle-Gas Compression

A two-stage recycle-gas-compression system is employed. The high-pressure stage supplies high-pressure recycle gas to be used for filter blowback and lockhopper pressurization. A gas receiver is included to provide the surge capacity necessary to accommodate the periodic gas flows to the depressurization hopper and the filter. Gasifier fluidization and ash-transport gas is provided by the low-pressure, high-volume stage of the recycle-gas compressor.

4.2.1.5 Gas Conditioning

This area consists of two subsystems, particulate removal and the chloride guard beds. Figure 4-7 is the process flow diagram of the gas conditioning system for Cases 1 and 3.

The cooled product gas (1000°F) is passed through a non-recycle cyclone and a ceramic candle gas filter to remove all of the remaining particulates from the gas. The collected solids are transferred to a depressurization lockhopper where the pressure is reduced to atmospheric and from which the solids are sent through a water-cooled conveyor to the sulfation area. Lockhopper pressurization is accomplished with gas from the high-pressure recycle gas compressor. The chloride guard consists of two parallel fixed beds of calcined nahcolite sorbent. The beds have isolation valves to allow alternate operation for chloride capture. When a bed becomes loaded, as indicated by outlet gas analysis, it will be isolated; the sorbent will be replaced and the bed will be available for the next duty cycle.

4.2.1.6 External Desulfurization

Zinc ferrite was selected as the external bed sorbent for high-temperature coal gas desulfurization because of its effectiveness and capability for sulfur absorption combined with its regenerative characteristics. Figure 4-8 is the process flow diagram of the external desulfurization system for Cases 1 and 3. The external desulfurization system consists of two zinc ferrite reactors and an exit-gas cooler. In a similar manner to the chloride guard drums, the two reactors are operated in a parallel arrangement.

Final product gas desulfurization is achieved in a fixed bed of zinc ferrite sorbent. As developed, the sorbent is a mixture of binders and modifiers and equimolar amounts of zinc oxide (ZnO) and ferric oxide (Fe_2O_3) with a combined chemical formula of ZnFe_2O_4 . The sulfur capture mechanism is described by the following reactions:

In addition, the water-gas shift reaction and the COS hydrolysis reaction are catalyzed by the iron oxide in the sorbent.

Reactions 2, 3, 4, and 5 are all exothermic, and a rise in temperature occurs as the gas passes through the bed.

The level of H_2S in the reactor exit gas is determined by the equilibrium of hydrogen sulfide with zinc oxide (reaction 3) and is a function of temperature and steam concentration. As the hydrogen sulfide is reacted with the sorbent, COS hydrolysis (reaction 5) proceeds to extremely low levels of COS. At the design operating conditions, the reactor exit gas will contain less than 20 ppm total sulfur.

As the gas passes through the bed, the sulfur is adsorbed by fresh sorbent, creating a moving front between the desulfurized gas and the feed gas. As this front approaches the end of the bed, and the sorbent reaches design loading, the gas flow is switched to the other reactor.

The zinc ferrite is regenerated by oxidation of the sulfides by the following reactions:

The oxidation reactions are highly exothermic and proceed as fast as oxygen is supplied. The progression of the moving front through the bed and the local bed temperature are controlled by the air and diluent steam flow. The regeneration exit gas is recycled to the gasifier, and the sulfur dioxide in the gas is captured by the limestone in the gasifier bed.

Regeneration of the first vessel begins with steam pressurization to a level above that in the gasifier. The flow of steam and air to the reactor is controlled to maintain the local bed temperature (at the oxidation front) below 1500°F to avoid sintering and destroying the sorbent. The minimum temperature range is 1100°F since lower temperatures will promote the formation of zinc sulfate.

NOTE:

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WHERE APPROPRIATE, THE FOLLOWING TRAINING IS REQUIRED:
ADDITIONAL TRAINING IS REQUIRED.

**Figure 4-7 Cases 1 and 3 Gas Conditioning
Process Flow Diagram**

5100, DA
ZINC FERRITE
REACTORS

220B
START-UP HEATER

251J
BOOSTER
AIR COMPRESSOR

251C
AIR
RECUPERATOR

250A
HEAT
PREHEATER

NOTES
1. 10
2. 10
3. 10
4. 10

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THERE ARE A TOTAL OF FOUR TRAINS FOR
CASE 1 AND TWO TRAINS FOR CASE 2.

ARD ON M. IN PROTECT THE GAS
IRBINE IN ASSESS AASTROPHIC
ALLURE OF THE A. PRITE HED.

Figure 4-8 Cases 1 and 3 External Desulfurization Process Flow Diagram

Any zinc sulfate formed during the oxidative regeneration is converted to zinc oxide by a subsequent reductive regeneration with hydrogen and carbon monoxide (via recycled fuel gas) and is then ready for the next desulfurization cycle. The desulfurized product gas enters the exit-gas cooler where the temperature is controlled at the maximum fuel-supply valve temperature for supply to the gas turbine. High-pressure saturated steam is produced at varied rates to effect the temperature control.

4.2.1.7 Gasifier Solids Sulfation

The mixture of spent (sulfided) limestone, ash, and fines is prepared for disposal in the fluidized-bed sulfator. This system is used to oxidize the calcium sulfide to calcium sulfate, a compound similar to gypsum, which can be readily disposed of in a dry landfill. The calcium sulfide oxidation reaction is highly exothermic and, at temperatures below 1600°F, is accomplished with little formation of sulfur dioxide. Figure 4-9 is the process flow diagram for the sulfation step in Cases 1 and 3.

The sulfation combustor is a fluidized-bed system. Because of the highly exothermic reaction of calcium sulfide, the environment in the fluidized bed is favorable for the combustion of any residual carbon remaining in the char fines. The considerable quantity of heat generated in the combustor is used to produce high-pressure saturated steam, thereby improving overall plant efficiency. The heat is recovered by in-bed exchanger tubes that keep the bed temperature less than 1600°F without the need for high excess air. The gas from the combustor is cooled by additional exchanger surface before it is routed to the coal preparation section of the plant for coal drying.

Feed to the sulfation combustor from the surge bin is controlled by a screw feeder included with the fluidized-bed system. Discharge solids from the bed are cooled in a screw conveyor/cooler and discharged for eventual disposal.

4.2.1.8 Gas Turbine System

The advanced-design GE MS7001F combustion turbine serves as the heart of the power island design for the IGCC cases. The GE MS7001F is a new combustion turbine design which has been introduced by General Electric, with the first unit having been recently installed and placed into commercial operation at Virginia Power's Chesterfield station in mid-1990. The GE MS7001F is a high-firing-temperature design which employs a turbine inlet temperature of approximately 2300°F.

The configuration of the power island for Case 1 is shown in Figure 4-10. The power island configuration for Case 3 is shown in Figure 4-11.

Inlet air to the combustion turbine is taken into the compressor and compressed to about 200 psi and 730°F. Some compressor discharge air is bled from the primary air stream and sent to the gasification island where it is used in the production of the coal gas. The fraction of the total compressor discharge air flow that is bled to the gasification process is approximately 15 percent. The remaining compressor discharge air goes to the combustor of the GE MS7001F where it mixes with the coal gas which then ignites. The hot gases from the combustor discharge pass through the power turbine section of the combustion turbine producing power which is used to drive the compressor and an electric generator.

4.2.1.9 HRSG System

The heat recovery steam generator (HRSG) takes the exhaust flow from the combustion turbine and uses the heat to produce steam for both power production in the steam turbine and for process steam use in the gasification island. One HRSG is used for each combustion turbine. The HRSG configuration is heavily integrated with the gasification island in order to achieve maximum steam production for the overall plant.

Two pressure levels are used in the steam system, one for the production of high-pressure (HP) steam to the turbine throttle and the other for intermediate-pressure (IP) steam production to supplement the steam flow to the reheat section of the steam turbine. An additional superheater section is also used to superheat the blast steam that is used in the gasification process. Figures 4-10 and 4-11 show the basic configurations of the HRSGs for Cases 1 and 3.

The following heat transfer sections are utilized in the HRSG:

- HP steam superheater
- IP steam reheater
- Blast steam superheater
- HP evaporator
- IP evaporator
- HP economizer 2
- IP economizer
- HP economizer 1
- Integral deaerator
- Feedwater heating coil

601F
LASH RECEIVER

601LV
SCREW FEEDER

601J
SULFATION
AIR COMPRESSOR

601B
SULFATOR

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NOTES:

1. THERE ARE A TOTAL OF THREE TRAINS FOR CASE 1 AND TWO TRAINS FOR CASE 2 FOR THE SULFATOR UNITING.
2. NORMALLY NOT FLOW.

**Figure 4-9 Cases 1 and 3 Sulfation
Process Flow Diagram**

4-33

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Figure 4-11 Case 3 Power Island Process Flow Diagram

The integral deaerator and feedwater heating coil are located within the HRSG to maximize the heat recovered from the exhaust gas stream and therefore maximize the power production of the steam turbine by eliminating extractions for deaerating steam and feedwater heating.

4.2.1.10 Steam Turbine System

A reheat steam turbine is the main component of the steam turbine system. The HP steam from the HRSG enters the HP turbine section of the turbine at 1450 PSIG and 1000°F and exhausts to the HRSG reheater. The steam is reheated to 1000°F and enters the IP turbine section. After passing through the IP/LP section of the turbine the steam exhausts to a condenser. An extraction port in the HP turbine is used to provide steam at approximately 500 psi for both quench steam and blast steam use in the gasifier island. There are no other extractions except typical shaft leakage and sealing steam flows.

The condensate from the condenser is sent to the HRSG feedwater heating coil and HRSG integral deaerator via condensate pumps. Boiler feedpumps send the feedwater from the deaerator to the HRSG economizers and gasifier island. A circulating water system using a mechanical draft cooling tower and circulating water pumps is used for condenser cooling water. The condenser design pressure is 3.5 in. HgA at the 90°F ambient design point with a 90°F cooling water temperature.

4.2.1.11 Ash Handling and Disposal

The ash handling system transports the ash from the discharge of the screw conveyors in the sulfator area to an onsite landfill for disposal. Figure 4-12 is the process flow diagram for the ash handling system for Case 1. The configuration of the system for Case 3 is similar to that of Case 1. The conveying system is a dilute-phase vacuum type with surge bins at the discharge of each screw conveyor to allow for up to one hour of storage. The ash is conveyed to a 40-hour storage silo with mechanical lobe type blowers acting as vacuum pumps. The 40-hour ash silo allows the landfill to be operated only one shift per day and also allows for ash storage when the landfill cannot be operated due to inclement weather. The silo has a flat bottom with aeration to assist in discharge of ash. The discharge of the silo is equipped with a conditioner that adds the optimum amount of water prior to discharge to a dump truck. A dry loader allowing discharge to a tanker truck is also provided, although the normal mode of operation would be to use dump trucks. The conditioned ash is transported to an onsite landfill by dump truck.

Figure 4-12 Case 1 Ash Handling System Process Flow Diagram

4.2.1.12 General Facilities

The general facilities include all site preparation and improvements, roads and rail facilities, buildings, fuel oil storage and handling, and fire protection. The cooling water systems, raw water supply, potable and utility water, sewers, and effluent water treating were also included in the general facilities. Electrical and control systems were considered part of the general facilities and included a central computer system, plant and instrument air, electrical switchyard and distribution, lighting, and communications. Miscellaneous items such as painting and interconnection of piping between area battery limits were also included in the general facilities.

4.2.2 Plant Layouts

Equipment design parameters were developed based on the results of the thermal analysis and the specific structural and material requirements for the various pieces of equipment. The detailed equipment descriptions are provided in Appendix A. Based on the functions, physical features, and interconnections of the equipment, conceptual plant layouts were developed for each IGCC plant.

The IGCC plant is located in an area on the Plant Wansley site reasonably convenient to the existing coal pile, switchyard, and transmission distribution system. However, one constraint on location was that the IGCC plant not interfere with the areas set aside for future Plant Wansley Units 3 and 4, which are currently designated to be pulverized coal-fired units. This constraint could be eliminated in the future if the IGCC technology replaces pulverized coal units as the technology of choice for baseload power generation at Plant Wansley.

Figures 4-13 and 4-14 show the overall site plan for the IGCC plants of Cases 1 and 3. The overall arrangement can be divided into four distinct and separate areas:

- Power Island
- Gasifier Island
- Coal and Limestone Bulk Storage
- Electrical Switchyard

A conceptual plot plan and elevation and isometric views of the gasification train for Cases 1 and 3 are given in Figures 4-15 and 4-16.

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

Figure 4-13 Case 1 Overall Site Plan

Figure 4-14 Case 3 Overall Site Plan

The general approach on developing the site layout was to make the most efficient use of the existing facilities (particularly the coal unloading facilities) while totaling specific considerations for a variety of factors such as environmental constraints regarding HRSG stack height, personnel safety, and maintenance access to equipment.

The power island layout is arranged such that the HRSG stacks are located at a sufficient distance from the gasifier island to minimize the height of the stacks as determined by environmental constraints. The combustion turbines and steam turbine are housed in a powerhouse which has an overhead crane for equipment maintenance. The combustion turbine inlet air filters are located on the sides of the powerhouse in order to minimize the effects of fugitive dust and emissions from the gasifier island and to minimize the effects of the inlet air ducting on crane access. Due to the location of the hot gas piping and requirements for maintenance access at the east end of the building (near the generators), the electrical switchyard cannot be located at the east end of the power island but must be located at the south end of the power island. This requires slightly longer lengths of high voltage cabling but maximizes employee safety and equipment accessibility. The cooling tower is located at a point which would minimize the effect of the cooling tower plume on nearby structures (prevailing winds are from the west/northwest) and minimize the circulating water system piping requirements. The power island equipment arrangement was developed using actual vendor equipment drawings for the major equipment including the HRSG, steam turbine, combustion turbine, and supporting pumps.

The layout of the coal and limestone handling facilities was developed in a manner to make the optimum use of the existing coal unloading facilities and to minimize the lengths of conveyors needed for coal and limestone transport to the gasifier island.

4.3 400-MW AIR-BLOWN IGCC WITH BULK EXTERNAL DESULFURIZATION (Case 5)

4.3.1 Process Discussion

Case 5 is a nominal 400-MW IGCC plant that is similar to Case 1 except for utilizing an external desulfurization step for the bulk of the sulfur removal in the process. In this concept, no limestone is fed to the air-blown KRW gasifier. The regenerative gas from the external desulfurization step is fed to a sulfuric acid plant to produce a byproduct suitable for sale. The power island configuration for Case 5 is similar to that of Case 1 with two GE MS7001F gas turbines, two HRSGs, and one steam turbine.

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-15
Conceptual Plot Plan -
Cases 1 and 3

4-43

Figure 4-16 (Sheet 2 of 2) Gasification Train Layout,
Isometric View - Cases 1 and 3

4.3.1.1 Overall Process Configuration

The Case 5 design is shown by the block flow diagram in Figure 4-17. The corresponding stream flows for the process stream numbers in Figure 4-17 are listed in Table 4-3. The process configuration in this case is similar to that employed in Case 1 with the following exceptions:

- Limestone addition to the gasifier for in-bed sulfur removal has been eliminated, and all desulfurization is accomplished in the external desulfurizers (zinc ferrite).
- Six gasifier trains (instead of four) are required to fuel an identical power island primarily because of gasifier vessel diameter limitations coupled with lower coal throughput per gasifier when limestone is not present.
- Since no calcium sulfide is produced in this configuration, the external sulfator has been eliminated. Fines and ash are disposed of without treatment.
- Nitrogen from the air separation plant is heated with steam and used for coal drying. No sulfator gas is available for this purpose.
- Regeneration gas from the zinc ferrite reactors is routed to a sulfuric acid plant for capture of the contained sulfur as sulfuric acid. An oxygen plant is required for operation of the sulfuric acid plant.

4.3.1.2 Coal Receiving, Handling, and Preparation

As shown in Figure 4-18, the layout of the coal preparation system for Case 5 is generally the same as in Case 1 except that the limestone system is not present. In addition, some equipment capacities are different because of differences in coal feed rate.

Since limestone is not fed to the gasifier in Case 5, a sulfator is not needed, and thus a different method of coal drying is required. Therefore, a mixture of nitrogen and air is used for coal drying. A 98-percent pure nitrogen stream is extracted from the oxygen plant which supplies the sulfuric acid plant, and the nitrogen is heated with steam and mixed with air for supply to coal preparation. The installed cost of the oxygen plant is increased by about 5 percent when a relatively pure nitrogen byproduct is extracted, since nitrogen is not required in the sulfuric acid plant. Nitrogen will not be available for coal drying during oxygen plant outages, since the installed cost would increase significantly if liquid nitrogen were produced for storage. However, the availability of the oxygen plant is approximately 99 percent. During oxygen plant outages, the oil-fired heaters used at startup would be utilized for coal drying.

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

Figure 4-17 Case 5 Block Flow Diagram

Table 4-3 (Sheet 1 of 5) Case 5 Process Stream Flows
Note: Streams Not Shown Are Proprietary to M.W. Kellogg

STREAM NUMBER	1	4	4A	10	13	14
DESCRIPTION	PREPARED COAL FEEDSTOCK	TOTAL AIR FROM POWER ISLAND	AIR TO CT COMPRESSOR	STEAM TO GASIFIER	HP BLOWDOWN TO POWER ISLAND	HIGH TEMP. BFW TO STEAM DRUM
	WT%	VOL%	VOL%	VOL%	VOL%	WT%
GASES, LB/HR						
CO	28,010	0.00	0.00	0		
H ₂	2,016	0.00	0.00	0		
CO ₂	44,010	0.00	0.00	0		
CH ₄	16,043	0.00	0.00	0		
N ₂	28,013	78.08	78.08	3,805,813		
Ar	39,948	0.98	0.98	46,787		
O ₂	31,999	20.86	20.86	1,021,024		
NH ₃	17,031	0.00	0.00	0		
H ₂ S	34,080	0.00	0.00	0		
CO _S	60,075	0.00	0.00	0		
SO ₂	64,063	0.00	0.00	0		
H ₂ O	13,015	0.00	0.00	58,574		
HCl	38,461	0.00	0.00	0		
TOTAL GASES		100.00	100.00	58,574	100.00	100.00
GAS FLOW, LB MOL/HR						
MOLECULAR WEIGHT						
GAS VOLUME, ACFM						
GAS VOLUME, SCFM						
LIQUIDS, LB/HR						
WATER	18,015					
SOLIDS, LB/HR						
COAL	88.86	28,194	188,238	3,140		
LIMESTONE		28.96	28.96	18.02		
ASH	10.34	43,839	1,125,347	1,758		
TOTAL SOLIDS	100.00	241,829	1,063,861	18,859	100.00	837,291
TOTAL FLOW, LB/HR	315,975	1,108,169	4,873,824	58,574	10,867	837,291
TEMPERATURE, F	200	755	90	850	605	600
PRESSURE, PSIA	14.0	180.0	14.7	450 ±	1,000	1,050

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 4-3 (Sheet 2 of 5) Case 5 Process Stream Flows

STREAM NUMBER	16	24	30	32*	39	44	48
DESCRIPTION	HP STEAM FROM STEAM DRUM	SOLIDS TO DISPOSAL	STEAM TO ZnF ₂ REACTOR	PRODUCT GAS TO GAS TURBINE	PRODUCT GAS QUENCH STEAM	FINES LOG-HOPPER VENT GAS	MLP BLOWDOWN TO POWER ISLAND
	VOL%	WT%	VOL%	VOL%	VOL%	VOL%	VOL%
GASES, LB/HR							
CO	28,010			7.52 186,070		7.52 40	
H ₂	2,016			20.83 32,907		20.83 8	
CO ₂	44,016			13.57 487,828		13.57 114	
CH ₄	18,043			0.85 8,228		0.85 2	
N ₂	28,013			38.43 798,716		38.43 184	
Ar	38,948			0.44 13,871		0.44 3	
O ₂	31,888			0.00 0		0.00 0	
NH ₃	17,031			0.06 1,008		0.06 0	
H ₂ S	34,080			0.00 0		0.00 0	
CO ₂	60,075			0.00 0		0.00 0	
SO ₂	94,063			0.00 0		0.00 0	
H ₂ O	18,015		100.00 328,423	20.48 288,061	100.00 230,510	20.48 70	
HCL	38,461			0.00 0		0.00 0	
TOTAL GASES	100.00 1,068,885		100.00 328,423	100.00 1,777,737	100.00 230,510	100.00 432	
GAS FLOW, LB MOL/HR	69,321		18,231	78,340	12,785	18	
MOLECULAR WEIGHT	18.02		18.02	22.99	18.02	22.99	
GAS VOLUME, ACFM	7,065		10,207	70,481	4,877	233	
GAS VOLUME, SCFM	375,135		115,287	485,538	80,818	120	
LIQUIDS, LB/HR							100.00 1,421.00
WATER	18,015						
SOLIDS, LB/HR							
COAL		27.58 12,420					
LIMESTONE		72.41 32,584					
ASH		100.00 45,014					
TOTAL SOLIDS							
TOTAL FLOW, LB/HR	1,068,885	45,014	328,423	1,777,737	230,510	432	1,421
TEMPERATURE, F	605	150	950	1,000	500	524	240
PRESSURE, PSIA	1,800	14.0	450.0	280.0	450.0	14.0	1,850

* Final fuel gas composition based on 98 percent removal of HCl across chloride guard and 99.2 percent removal of H₂S across zinc ferrite bulk removal system.

Table 4-3 (Sheet 3 of 5) Case 5 Process Stream Flows

STREAM NUMBER	50	51	52	53	54	55
DESCRIPTION	PREHEATED BFW TO POWER ISLAND	COOLED REGEN. GAS TO ACID PLANT	TOTAL DEALER BFW FROM POWER ISLAND	EXHAUST GAS TO HPSSG	FLUE GAS TO MAIN STACK	BFW TO HPSSG
	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %
GASES, LB/HR						
CO	28,010	0.00	0			
H ₂	2,018	0.00	0			
CO ₂	44,910	0.00	0			
CH ₄	18,043	0.00	0			
N ₂	28,013	7.92	42,008			
Ar	39,948	0.09	735			
O ₂	31,899	0.00	0			
NH ₃	17,031	0.00	0			
H ₂ S	34,080	0.00	0			
CO ₈	80,075	0.00	0			
SO ₂	84,083	1.21	15,498			
H ₂ O	18,015	91.18	329,423			
HCl	38,481	0.00	0			
TOTAL GASES		100.00	388,752	100.00	100.00	
GAS FLOW, LB MOL/HR			19,908			
MOLECULAR WEIGHT			19			
GAS VOLUME, ACFM			221,470			
GAS VOLUME, SCF*			128,438			
LIQUIDS, LB/HR						
WATER	18,015		571,858			100.00
SOLIDS, LB/HR						
COAL						
LIMESTONE						
ASH						
TOTAL SOLIDS						
TOTAL FLOW, LB/HR	258,533	388,752	571,858	9,651,361	9,651,774	1,822,615
TEMPERATURE, F	436	470	240	1,100	280	121
PRESSURE, PSIA	1,600.0	15.0	1,650.0		15.0	1.7

Table 4-3 (Sheet 4 of 5) Case 5 Process Stream Flows

STREAM NUMBER	60	67A	67B	68	69	61
DESCRIPTION	BPW MAKEUP	HP BOILER BLOWDOWN	HP BOILER BLOWDOWN	COOLING TOWER MAKEUP	COOLING TOWER BLOWDOWN EVAP., DRIFT	STEAM FROM POWER ISLAND
	VOL. %	VOL. %	VOL. %	VOL. %	VOL. %	VOL. %
GASES, LB/HR						
CO	28,612					
H ₂	2,916					
CO ₂	44,816					
CH ₄	16,943					
N ₂	28,915					25,184
A ₂	28,948					
O ₂	21,888					
NH ₃	17,831					
H ₂ S	24,000					
CO _S	80,875					
SO ₂	94,963					
H ₂ O	18,915					
HCl	28,491					
TOTAL GASES						
GAS FLOW, LB MOL/HR						
MOLECULAR WEIGHT						
GAS VOLUME, ACFM						
GAS VOLUME, SCFM						
LIQUIDS, LB/HR	100.00	100.00	100.00	100.00	100.00	100.00
WATER	800,000	12,867	1,154	2,843,917	2,843,917	284,287
SOLIDS, LB/HR						
COAL						
LMESTONE						
ASH						
TOTAL SOLIDS						
TOTAL FLOW, LB/HR	800,000	12,867	1,154	2,843,917	2,843,917	284,287
TEMPERATURE, F	80	800	420	80	80	80
PRESSURE, PSIA	14.7	1,000.0	230.5	14.7	14.7	14.7

Table 4-3 (Sheet 5 of 5) Case 5 Process Stream Flows

STREAM NUMBER	62	63	64
DESCRIPTION	CONDENSATE RETURN	HEATED COAL DRYING GAS	COAL DRYING GAS TO LOCAL STACK
	VOL %	VOL %	VOL %
GASES, LB/HR			
CO	24,010		
H ₂	2,016		
CO ₂	44,010		
CH ₄	19,043		
N ₂	25,013		
Ar	20,948		
O ₂	31,990		
NH ₃	17,031		
H ₂ S	34,090		
CO ₂	60,075		
SO ₂	64,003		
H ₂ O	18,015		
HCl	26,401		
TOTAL GASES		100.00	100.00
			127,945
GAS FLOW, LB MOL/HR			
MOLECULAR WEIGHT			
GAS VOLUME, ACFM			
GAS VOLUME, ICFM			
LIQUIDS, LB/HR			
WATER			
			18,015
SOLIDS, LB/HR			
COAL			
LIMESTONE			
ASH			
TOTAL SOLIDS			
		100,238	127,945
TOTAL FLOW, LB/HR			
TEMPERATURE, F			
PRESSURE, PSIA			

COAL SYSTEM

Figure 4-18 Case 5 Coal Preparation Process Flow Diagram

The coal drying gas entering the pulverizers is a mixture of nitrogen, air, and recycle gas from the drying loop and has a maximum oxygen content of 12 percent by volume. A steam coil heater is used to heat the gas stream to 365°F. This approach is more economical than exclusively using an oil-fired heater for coal drying. An economic evaluation indicated that the payout period for the increased capital costs is approximately 1 year.

4.3.1.3 Air Compression and Supply

This area shares the same design as Cases 1 and 3. Only the equipment sizing is significantly different. (Refer to Section 4.2.1.3.)

4.3.1.4 Coal Gasification and Recycle Gas Compression

As can be seen from the process flow diagrams in Figures 4-19 and 4-20 for the gasification step and the recycle gas system, the process configuration is similar to that of Case 1. One difference is that the product gas from the gasifier passes through a cooler prior to entering a cyclone to remove particulates. This is due to the absence of limestone in Case 5, which results in a stickier ash in the product gas and requires that the gas be cooled prior to entering the cyclone.

4.3.1.5 Gas Conditioning

The gas conditioning step shown in Figure 4-21 is similar to that in Case 1 except that the ash and fines are sent to disposal rather than to the sulfator.

4.3.1.6 External Desulfurization

The process flow diagram for the fixed-bed external desulfurization step for Case 5 is shown in Figure 4-22. This step is similar in configuration to that of Case 1, but is used to remove the bulk of the sulfur from the product gas rather than as a sulfur polishing step. This necessitates greater sorbent capacity and a different mode of operation. It will be noted that for bulk sulfur removal, two zinc ferrite vessels are normally connected in series. During regeneration, a vessel would be taken offline. After regeneration, it would be reconnected as the last vessel in the gas path.

4.3.1.7 Sulfuric Acid Plant - Area 700

The sulfuric acid plant employed in this facility is a double-contact, double-absorption sulfuric acid process with facilities for burning molten sulfur to keep the plant on hot standby during lapses in regeneration gas flow. A process flow diagram is given in Figure 4-23. The plant is designed for an availability of 98 percent and for an acid conversion of 99.7 percent to produce a product acid of 98.5-percent strength. The plant has a maximum capacity of 400 short tons per day of monohydrate sulfuric acid.

Under normal conditions, the sulfur bearing regeneration gas from the zinc ferrite absorbers is cooled, mixed with oxygen, and cleaned in a wash tower and electrostatic precipitator, after which it passes to the drying tower where the contained moisture is absorbed in circulating, strong sulfuric acid. The dried gas is heated and then enters the first catalyst layer in a multibed vanadium pentoxide catalyst converter.

Conversion of sulfur dioxide to sulfur trioxide takes place as the gas passes through each of the first three layers. The reaction heat evolved in each layer is removed by heat transfer to the cooler incoming gas stream in a series of external heat exchangers.

The gas exiting the third layer is cooled and then enters the first absorption tower where most of the sulfur trioxide is removed by absorption in circulating sulfuric acid.

After the sulfur trioxide is absorbed, the remaining gas is reheated by heat exchange with hot reaction gas and flows to the next catalyst layer in the converter. At this point in the process, a greater driving force is available than in a conventional single-contact, single-absorption sulfuric acid plant because there is a high oxygen-to-sulfur-dioxide ratio and, at the same time, there is almost no sulfur trioxide to drive the reaction in the reverse direction. Therefore, a high conversion efficiency is obtained, thus reducing the amount of unconverted sulfur dioxide leaving in the exit stack gas.

Gas from the fourth catalyst layer is cooled and passes to a second absorber for recovery of the produced sulfur trioxide. A stripping tower is incorporated in the process to reduce the loss of dissolved sulfur dioxide to the final absorption system. This process step results in a reduction in the amount of sulfur dioxide discharged to the stack and simultaneously increases the yield of the plant.

When the flow of regeneration gas from the zinc ferrite ceases, or contains insufficient sulfur dioxide to maintain the acid plant in operation, molten sulfur will be burned in a furnace as an alternate source of sulfur dioxide. Since the sulfur dioxide produced in this manner will be clean, dry, and hot, it will be fed directly to the converter.

Oxygen for the sulfuric acid plant is supplied by an air separation plant described below. A plot plan of the acid plant is given in Figure 4-23 showing the battery limits (BL) of the sulfuric acid plant. Also shown on the plot plan are outside battery limit (OBL) areas of sulfur storage, acid storage, and a neutralization system.

301J
PRESSURIZATION
AIR COMPRESSOR

302F
PRESSURIZATION
AIR RECEIVER

310L
COAL FEED
VENT FILTER

310F
FEED SURGE DRUM

311F
FEED
PRESSURIZATION
HOPPER

COAL FROM
AREA 100

PRESSURIZATION
AIR

TO ATMOSPHERE

301J
302F
310L
310F
311F
312F
310V

301J
302F
310L
310F
311F
312F
310V

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

**Figure 4-19 Case 5 Gasification
Process Flow Diagram**

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

NOTES:

1. FLOW RATE FOR STREAM IS 2% TIME AVERAGED.

**Figure 4-20 Case 5 Recycle Gas System
Process Flow Diagram**

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-21 Case 5 Gas Conditioning Process Flow Diagram

NOTES

1. GUARD DRUM IS TO PROTECT THE GAS TURBINE IN CASE OF CATASTROPHIC FAILURE OF ZINC FERRITE BID.
2. THERE IS ONE COMMON REGENERATION GAS COOL DOWN TRAIN (5, 10 AND 5) FOR ALL SIX GASIFICATION TRAINS.

1.

The acid plant cost provided in Section 8.0 includes both the BL and OBL portions of the plant. Cooling water for the plant is required for the power island, and a twin rail spur is recommended for delivery of molten sulfur to the plant by insulated tank car and for outgoing shipment of product acid by tank car.

4.3.1.8 Oxygen Plant - (for Sulfuric Acid Plant)

The oxygen plant is a standard 125-ton-per-day package unit that would be obtained on a subcontract basis from a major supplier. The oxygen plant can be designed to recover nitrogen. The nitrogen from the oxygen plant is used for coal drying in the coal preparation step.

4.3.1.9 Gas Turbine System

The gas turbine configuration for Case 5 is similar to that developed for Case 1 (Section 4.2.1.8) in that two advanced-design GE MS7001F combustion turbines serve as the heart of the power island design.

The only significant differences between the GE MS7001F design for Cases 1 and 3 versus Case 5 are the composition and heating value of the fuel produced by the gasifier island. The power island process flow diagram for Case 5 is shown in Figure 4-24.

4.3.1.10 HRSG System

The heat recovery steam generator configuration for Case 5 is virtually identical to that developed for Case 1 with the major variation being differences in interface flows with the gasifier island. Refer to Section 4.2.1.9 for details.

4.3.1.11 Steam Turbine System

The steam turbine configuration for Case 5 is practically identical to that developed for Case 1 with the only variation being differences in interface flows with the gasifier island. Refer to Section 4.2.1.10 for details.

4.3.1.12 Ash and Fines Handling and Disposal

The design of the system for this case is the same as for Case 1 (Section 4.2.1.11), except that the interface with the gasifier island is at a different location and the number of interface points is different. Also, some equipment capacities are different because of differences in the amount of solid waste.

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-24 Case 5 Power Island Process Flow Diagram

4.3.1.13 General Facilities

The general facilities are generally the same as those listed for Case 1 in Section 4.2.1.12.

4.3.2 Plant Layout

Plant layout for Case 5 is very similar to that for Case 1. The only significant differences are found in the layout of the gasifier island for Case 5 which requires an oxygen plant and a sulfuric acid plant and has eliminated the sulfator and the limestone storage and handling facilities. The power island layout is identical to that of Case 1. Figure 4-25 shows the overall site plan for the Case 5 IGCC plant. A conceptual plot plan and elevation and isometric views of the gasification train for Case 5 are given in Figures 4-26 and 4-27. Details on plant equipment are given for this case in Appendix A.

4.3.3 Alternative Desulfurization/Sulfuric Acid Production Approach

As described, the bulk desulfurization concept selected for Case 5 is a fixed-bed zinc ferrite system coupled to a double contact, double absorption sulfuric acid plant. GE Environmental Services, Inc., (GEESI) is currently conducting pilot plant trials of a moving-bed zinc ferrite system which has been proposed for demonstration in the Innovative Clean Coal Technologies program. GEESI believes that the moving-bed system, coupled with an alternative sulfuric acid plant design, has substantial performance and cost advantages over the system used in the current study. Detailed evaluation of these alternatives was beyond the scope of the current study. However, further discussion of the GEESI concept is provided in Appendix B in order to give additional information on this approach. As development of the GEESI system proceeds, it will be appropriate to address the process performance and costs in detailed studies.

4.4 400-MW IGCC WITH TRANSPORT GASIFIER (Case 6)

4.4.1 Process Discussion

The IGCC plant for Case 6 is sized at nominally 400 MW and utilizes the transport gasifier concept currently under development by MWK. As described subsequently, the transport gasifier is a circulating-bed gasifier which uses pulverized coal and limestone. It is expected that the small particle size of the limestone for the transport gasifier will increase the sulfur capture within the gasifier sufficiently to eliminate the zinc ferrite polishing step that was required with the KRW gasifiers. Also, the coal throughput is higher for the transport gasifier, which reduces the number of gasifiers and gas processing trains from four as in Case 1 down to two in Case 6. Solid waste from each gasifier is fed to a pressurized sulfator for calcium

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-25 Case 5 Overall Site Plan

sulfide oxidation, with the pressurized flue gas from the sulfator sent to the power island for additional power generating capacity. The power island configuration for Case 6 is similar to that of Case 1 with two gas turbines, two HRSGs, and a single steam turbine.

4.4.1.1 Overall Arrangement

The block flow process diagram for Case 6 is shown in Figure 4-28 with corresponding process stream flows listed in Table 4-4. Coal and limestone is sized and dried prior to being fed to the transport gasifier. Limestone is dried with an oil-fired heater as in Case 1. The coal is dried using a slipstream of hot flue gas taken from the HRSG. Gasification air, as well as air for the sulfator, is provided by two air booster compressors (one train per gas turbine) which boost the pressure of the air extracted from the gas turbines.

The product gas from the transport gasifier is cooled in a heat recovery step to produce high pressure steam before entering the high temperature ceramic filters for removal of particulates from the product gas. The clean product gas from the gas filters is then sent directly to the combustion turbine.

The ash from the gasifier and fines from the gas filtration step are sent to a pressurized transport sulfator where calcium sulfide is oxidized to calcium sulfate. The pressurized flue gas from the sulfator goes to a gas filtration step before being sent to the combustion turbine. The solid waste from the sulfator is cooled in a heat recovery step to produce steam followed by a final cooling step prior to disposal.

The cleaned product gas from the gasifier and the pressurized flue gas from the sulfator are both utilized in the combustion turbine to produce power. The flue gas from each combustion turbine is sent to a separate HRSG to produce steam for the gas island and power generation in a single steam turbine as in Case 1.

4.4.1.2 Coal and Limestone Receiving, Handling, and Preparation

The configuration of the system for Case 6 is the same as for Case 1 except for the differences listed below:

- The grind size is smaller (pulverized product versus minus ¼-inch).
- The coal silos and pulverizers have been relocated to a site closer to the HRSGs to make the use of hot flue gas cost effective.
- Case 6 has five storage silos and five pulverizers (four operating and one spare pulverizer).
- Coal and limestone are conveyed into the gasifier separately.

BATTERY UNIT 120 C.N.

120 B

TIME STACK

4.1967
CONNA

BATTERY UNIT 120 C.N.

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-26
Conceptual Plot Plan - Case 5

Figure 4-27 (Sheet 1 of 2) Gasification Train Layout, Elevation View - Case 5

Figure 4-27 (Sheet 2 of 2) Gasification Train Layout, Isometric View - Case 5

**Assessment of Coal Gasification/Hot Gas
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Figure 4-28 Case 6 Block Flow Diagram

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 4-4 (Sheet 1 of 4) Case 6 Process Stream Flows
NOTE: M. W. Kellogg Proprietary Streams Are Not Shown

STREAM NUMBER	1	2	4	4A	5	10	13
DESCRIPTION	PREPARED COAL FEEDSTOCK	LIMESTONE FEEDSTOCK	TOTAL AIR FROM POWER ISLAND	AIR TO CT COMPRESSOR	VENT TO ATMOSPHERE	STEAM TO GLASSIER	TOTAL BLOWDOWN TO POWER ISLAND
	WT%	WT%	VOL%	VOL%	VOL%	VOL%	WT%
GASES, LB/HR							
CO	28,010		0	0.00	0		
H ₂	2,010		0	0.00	0		
CO ₂	44,010		0	0.00	0		
CH ₄	10,043		0	0.00	0		
N ₂	28,013		78.00	78.00	78.00		
Ar	30,948		13,221	0.98	24,960		
O ₂	31,800		231,163	20.95	544,701		
NH ₃	17,031		0	0.00	0		
H ₂ S	34,000		0	0.00	0		
CO ₂	90,075		0	0.00	0		
SO ₂	64,063		0	0.00	0		
H ₂ O	10,015		0	0.00	0		
HCl	30,461		0	0.00	0		
C ₂ H ₄	70,114		0	0.00	0		
HCN	27,028		0	0.00	0		
TOTAL GASES			100.00	100.00	100.00	100.00	
			808,460	2,000,005	1,590	80,234	
GAS FLOW, LB MOL/HR			34,474	80,770		6,342	
MOLECULAR WEIGHT			28.96	28.96		18.02	
GAS VOLUME, ACFM			39,954	600,365		3,027	
GAS VOLUME, SCFM			219,008	567,553		35,781	
LIQUIDS, LB/HR							
WATER							100.00
							5,427
SOLIDS, LB/HR							
COAL	80.80	253,775					
LIMESTONE	10.34	29,208					
ASH							
TOTAL SOLIDS	100.00	283,041					
			808,460	2,000,005	1,590	80,234	
TOTAL FLOW, LB/HR	283,041	27,987					5,427
TEMPERATURE, F	200	90	740	80	520	850	805
PRESSURE, PSIA	14.0	14.0	185.0	14.7	800	350.0	1,600.0

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

Table 4-4 (Sheet 2 of 4) Case 6 Process Stream Flows

STREAM NUMBER	17	28	28	32	33	35
DESCRIPTION	PRODUCT GAS TO GAS TURBINE	FLUE GAS TO POWER ISLAND	SOLIDS TO DISPOSAL	TOTAL DEASER BFW FROM POWER ISLAND	PREHEATED BFW TO POWER ISLAND	TOTAL HT BFW FROM POWER ISLAND
	VOL%	VOL%	WT%	VOL%	VOL%	VOL%
GASES, LB/HR						
CO	21.38	300,941	0			
H ₂	17.43	17,971	0			
CO ₂	7.50	185,927	12,753			
CH ₄	0.81	0,837	0			
N ₂	46.04	657,253	86,187			
Ar	0.57	11,460	1,729			
O ₂	0.00	0	8,745			
NH ₃	0.00	18	0			
H ₂ S	0.00	61	0			
COS	0.00	0	0			
SO ₂	0.00	0	2			
H ₂ O	5.26	47,630	625			
HCl	0.01	135	0			
HCN	0.03	368	0			
TOTAL GASES	89.80	1,206,029	120,051	100.00	100.00	100.00
GAS FLOW, LB MOL/HR						
MOLECULAR WEIGHT	49.654	4,088				
GAS VOLUME, ACFM	24.33	29.37				
GAS VOLUME, SCFM	46,867	5,424				
LIQUIDS, LB/HR	318,104	25,853				
WATER				100.00	100.00	100.00
18.015				324,308		479,670
SOLIDS, LB/HR						
COAL			1.14			
LIMESTONE			57.20			
ASH			41.86			
TOTAL SOLIDS			100.00			
			70.248	324,308	255,942	479,670
TOTAL FLOW, LB/HR	1,223,825	120,051				
TEMPERATURE, F	1,000	1,819	150	240	345	800
PRESSURE, PSIA	280.0	280.0	14.0	1,650	1,650	1,650

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 4-4 (Sheet 3 of 4) Case 6 Process Stream Flows

STREAM NUMBER	36	37	38	39	40
DESCRIPTION	TOTAL STEAM EXPORT TO POWER ISLAND	EXHAUST GAS TO HRSG	FLUE GAS TO MAIN STACK	BFW TO HRSG	BFW MAKEUP
	VOL %	VOL %	VOL %	VOL %	VOL %
GASES, LB/MR					
CO	25.010				
H ₂	2.916				
CO ₂	44.610				
CH ₄	16.043				
N ₂	23.913				
Ar	20.948				
O ₂	31.989				
NH ₃	17.031				
H ₂ S	34.080				
CO ₂	60.075				
SO ₂	64.063				
H ₂ O	18.916				
HCl	26.461				
HCN	27.026				
	100.00	542.671			
TOTAL GASES	100.00	542.671	100.00	5,324,258	
GAS FLOW, LB/MOLR					
MOLECULAR WEIGHT	30.123				
GAS VOLUME, ACFM	18.02				
GAS VOLUME, SCFM	3,682				
	180,452				
LIQUIDS, LB/MR					
WATER	18.016				
				100.00	1,181,480
110,900					
SOLIDS, LB/MR					
COAL					
LIMESTONE					
ASH					
TOTAL SOLIDS					
	542.671	6,423,940	5,324,258	1,181,480	110,900
TOTAL FLOW, LB/MR					
	605	1,120	281.00	121	80
TEMPERATURE, F					
	1,000.0	16.0	15.0	1.7	14.7
PRESSURE, PSIA					

Table 4-4 (Sheet 4 of 4) Case 6 Process Stream Flows

STREAM NUMBER	41A HP BOILER BLOWDOWN	41B IP BOILER BLOWDOWN	42 COOLING TOWER MAKEUP	43 COOLING TOWER BLOWDOWN EVAP., DRIFT	44 FLUE GAS TO LOCAL STACK	61 FLUE GAS TO LOCAL STACK
DESCRIPTION	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %
GASES, LB/HR						
CO	28.010					
H ₂	2.018					
CO ₂	44.059					
CH ₄	16.043					
N ₂	28.013					
Ar	39.948					
O ₂	31.868					
NH ₃	17.031					
H ₂ S	34.060					
CO _S	90.075					
SO ₂	64.063					
H ₂ O	18.015					
HCl	36.461					
HCN	27.028					
TOTAL GASES						
GAS FLOW, LB MOL/HR						
MOLECULAR WEIGHT						
GAS VOLUME, ACFM						
GAS VOLUME, SCFM						
LIQUIDS, LB/HR						
WATER	18.015					
SOLIDS, LB/HR						
COAL						
LIMESTONE						
ASH						
TOTAL SOLIDS						
TOTAL FLOW, LB/HR	10,460	1,317	3,630,231	3,630,231	130,540	106,258
TEMPERATURE, F	605	429	80	14.7	90	90
PRESSURE, PSIA	1,000	339.5	14.7	14.7	15.0	15.5

- Nitrogen is used to convey the coal to the gasification area with a closed-loop pneumatic conveying system; nitrogen is required for inerting due to the small coal particle size.
- Some equipment capacities are different because of the differences in coal feed rate.

Figure 4-29 is the process flow diagram for the coal and limestone preparation area.

In this case, the pulverizers were relocated to allow hot flue gas from the HRSGs to be used for coal drying, since sulfator flue gas is not available. A sulfator is present in Case 6, but the sulfator is adjacent to the gasifier and is pressurized to the same pressure as the gasifier. Since the flue gas from the sulfator is under pressure, it is sent to the combustion turbines. Also, nitrogen is not available for drying since an air separation plant is not needed for Case 6.

With air-blown gasification using fine coal particles, coal drying becomes a very important design consideration since nitrogen is not available for coal drying and inerting. Typically, coal handling and preparation are located far from the combined-cycle equipment to minimize the exposure of the equipment to coal dust. However, the economics are much more favorable using HRSG flue gas for coal drying as compared to oil-fired heaters. Heat losses from the drying gas would be too great if the flue gas were transported from the HRSG to the gasifier side of the plant and then back to the stacks at the HRSG. Therefore, the coal preparation area for Case 6 has been moved to a location nearer the HRSGs.

The layout of the limestone preparation system is the same as for Case 1 except that a dedicated pneumatic system is used to convey the limestone from the 24-hour storage silo to a surge bin in the gasification area.

4.4.1.3 Air Compression and Supply

This area is comprised of two subsystems: the boost air compression system and the lockhopper air compression system. The subsystems are described separately below and are shown schematically in Figure 4-30.

Boost Air Compression (Area 250)

The boost air compression system provides the air necessary for the operation of the gasifier and the sulfator. Air for this system is obtained by extraction from the compressor section of the gas turbine located in the combined-cycle power plant.

Figure 4-29 Case 6 Coal and Limestone Preparation Process Flow Diagram

Air from the gas turbine compressor discharge is first cooled in the air recuperator. This exchanger simultaneously cools the relatively high-temperature air obtained from the gas turbine and heats the compressor discharge air prior to delivery to the gasifier and the sulfator. After the initial air cooling is accomplished, a BFW preheater and a trim cooler are employed to cool the air to a lower temperature for more efficient air compression. A small stream of air from the booster compressor discharge is bypassed around the recuperator for compression and utilization as sulfator filter blowback gas. A knockout drum is provided at the compressor suction to remove condensed moisture.

Lockhopper Air Compression

A separate air compressor is provided for lockhopper pressurization. The compressor is sized for the average flow required by the lockhoppers. To accommodate the relatively large periodic flows to the lockhoppers, a surge vessel is provided. There are three lockhopper air compression trains.

4.4.1.4 Coal Gasification and Recycle Gas Compression

This area consists of three functional subsystems: coal/sorbent pressurization, gasification, and heat recovery. These subsystems are described below and illustrated in Figure 4-30, the gasification process flow diagram.

Coal/Sorbent Pressurization

A conventional lockhopper system is used to pressurize the coal and limestone feed to the gasifier. Unlike the KRW gasifier, coal and limestone are not mixed together but are pressurized and fed separately to the gasifier to ensure that the desired calcium to sulfur ratio is maintained in the short-contact-time transport gasifier. The pressurization trains for coal and limestone are functionally identical, although the coal train is much bigger than the limestone train. The following description of the coal pressurization train is also applicable to the limestone pressurization train.

Coal is delivered to the coal feed surge drum from the coal preparation area. The coal feed pressurization hopper periodically receives coal from the feed surge drum. Two valves are used between the vessels: the upper valve to initiate or stop solids flow and the lower to provide a gas tight seal. After a batch of coal is received, the coal feed pressurization hopper is pressurized with high-pressure air. When the pressure in the coal feed pressurization hopper is equalized with that of the coal feed hopper, a valve located between these two vessels is opened to dump coal into the feed hopper. After the coal feed pressurization hopper is emptied, the valve is closed and the hopper is vented to atmospheric pressure to repeat the cycle. A vent filter is installed on the pressurizing/venting line to essentially eliminate particulate emissions

<u>3201A, A</u>	<u>3201B</u>	<u>3201C</u>	<u>3201D</u>	<u>3201A, B</u>	<u>3211A, B</u>
IMSTONE FEED SAMPLER	IMSTONE FEED PREGRINDITION HOPPER	IMSTONE FEED HOPPER	IMSTONE FEEDER	TRANSPORT TRANSFER	PRIMARY CYCLONE
	<u>3201A, B</u>	<u>3201C</u>			
	TRANSFER STANDPIPE	TRANSFER STANDPIPE			
	CYCLONES

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when the coal feed pressurization hopper is vented. The hopper is pressurized through this same filter to provide an automatic backflushing of the filter media.

The coal feed hopper is operated at constant pressure sufficient to permit starwheel feed metering and pneumatic conveying of the coal to the gasifier by an air stream. The relative proportions of coal and limestone fed to the gasifier will be established by setpoint adjustments based on sulfur content of the gasifier product gas obtained from an online sulfur analyzer.

Gasification

The coal is gasified in a pressurized transport gasifier. The gasifier is a refractory-lined-riser-type reactor. The transport gasifier is made up of two sections, namely the mixing zone; where the recycled hot solids, coal and limestone, steam, and air are introduced; and the riser section. The mixing zone provides adequate residence time for gas and solid mixing. The fresh coal devolatilizes and the volatiles burn consuming all the supplied oxygen. Also, part of the sulfur in the coal is released here, and gasification and sulfur capture begin in this zone. The riser section provides additional residence time for coal gasification and sulfur capture.

Coal and limestone are fed continuously to the transport gasifier by their respective rotary feeders. Steam required for the gasification reactions is imported from the power island and fed to the gasifier. Air from the air recuperator is admitted at the bottom of the gasifier. All fresh feeds are mixed with recycle char from the gasifier standpipe in a mixing zone in the bottom of the gasifier where the devolatilization, gasification, and combustion reactions begin. The temperature in this zone is controlled to achieve desired carbon conversion.

A very high carbon conversion is predicted for the transport gasifier, with the remaining char being burned in the transport sulfactor for nearly 100-percent carbon conversion. Conversion in the gasifier is controlled by maintaining the proper solids circulation rate in the system, thus controlling the gasifier bed density. The limestone sorbent completely calcines in the gasifier and removes most of the H_2S produced by the gasification reactions.

The gas/solids mixture flows up through the gasifier at relatively high velocity and exits the reactor at the top. The heavily solids-laden fuel gas enters the primary cyclones where bulk solids removal occurs and then flows through secondary cyclones, which remove most of the remaining solids. Two primary and two secondary cyclones are provided for each gasifier. Collected solids from the cyclones flow by gravity to the gasifier standpipe. This vessel provides the pressure head needed to maintain adequate solids circulation. The vessel is aerated with a small amount of recycle gas,

and standpipe cyclones, two per standpipe, are provided to remove any entrained particles before the recycle gas exit stream is combined with the product fuel gas. Solids from the standpipe are returned to the gasifier. The unconverted carbon and partially sulfided sorbent are removed from the system by withdrawal from the primary cyclones and are sent directly to the transport sulfator. The gas leaving the cyclones is at 1455°F and contains small quantities of solids. It must be cooled to 1000°F and filtered prior to admission to the gas turbine.

Heat Recovery

Product gas leaving the gasifier secondary cyclone enters the fire-tube, downflow product gas cooler, where high-pressure steam is generated from BFW supplied by the HRSG steam drum. The water circulation from the steam drum to the product gas cooler is by natural convection. Heat is also recovered in the recycle gas cooler by preheating boiler feed water, which cools a small portion of fuel gas prior to compression for use as filter blowback gas.

4.4.1.5 Gas Conditioning

This area consists of two sections, namely, gas filtration and blowback gas compression, which are described below. Figure 4-31 is the gas conditioning process flow diagram.

Gas Filtration

The cooled gas stream from the product gas cooler flows to the gas filter, which removes all of the remaining particulates from the gas. These filters are of the ceramic-candle type and require periodic blowback. Fines collected in the gas filter are sent to the transport sulfator. The clean fuel gas, at about 1000°F and 280 psia, is sent directly to the gas turbine.

Blowback Gas Compression

A single-stage recycle compressor is employed here to compress a cooled slipstream of product gas for use as blowback gas. A gas receiver is included to provide the surge capacity necessary to accommodate the periodic flow of the blowback gas.

4.4.1.6 Gasifier Solids Sulfation

The process flow diagram for the sulfation step is illustrated in Figure 4-32. Solids fed to the transport sulfator flow from the dipleg of the gasifier primary cyclone through two refractory-lined and adequately aerated return legs. The transport sulfator is made up of a mixing zone and a riser section. The char and sulfided limestone from the gasifier are injected in the mixing zone where they are mixed

9330
RECEIVED
DIRECTOR

601B
TRANSPORT SULFATOR

601CA, B
RAYONLET BOILERS

602C
FLUID BED HEAT
EXCHANGER

603L
SULFATOR BARRIER
FILTER

604C
HIGH PRESSURE ASH
COOLER

610F
SULFATOR
STEAM DRUM

601LA, B
SULFATOR PRIMARY
CYCLONES

602LA, B
SULFATOR SECONDARY
CYCLONES

612L
ASH DEPRESSURIZATION
VENT FILTER

•

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with hot recycled solids and primary air. Combustion of char, sulfur release, sulfur capture, and oxidation of calcium sulfide to calcium sulfate start in the mixing zone. As the gas/solids mixture flows up through the riser at high velocity the sulfided sorbent in the feed from the gasifier is converted to CaSO_4 . Also, as residual char burns, sulfur is released as SO_2 . This is captured by the sorbent and converted to CaSO_4 . This reaction is expected to proceed fully because of the very fine particles used and the high degree of turbulence in the riser. Virtually complete SO_2 removal is obtained in the sulfator. The exothermic heat of the combustion and sulfation reactions results in a relatively small increase in gas/solids temperature leaving the reactor. The solids recirculation rate is set to control this temperature rise. Combustion of the char in the reactor is essentially complete.

The solids flow rate through the feed leg to the combustor is to be controlled by a plug valve. A mechanical valve is used in this leg to enable the gasifier and sulfator to be decoupled should this be required. The designed solids height in the feed leg is adequate to drive the char mixture into the mixing zone of the transport sulfator. Most of the air is fed to the mixing zone in the bottom of transport sulfator where it contacts char and sorbent fed from the gasifier system together with recycled solids from the fluid bed heat exchanger. Provisions are included to stage the combustion air feed in order to minimize NO_x formation.

The gas/solids mixture leaves the sulfator at 1650°F and flows through primary cyclones where bulk solids separation occurs. The flue gas then flows through secondary cyclones where most of the remaining solids are recovered from the gas. Two primary and two secondary cyclones are required for each sulfator.

Solids collected in the primary and secondary cyclones are fed to fluid-bed heat exchangers where the net heat of reaction is removed by cooling of the solids. Steam coils are located in these fluidized beds. A slipstream of solids is taken from these exchangers and sent to bayonet boilers for additional heat removal. In all of these exchangers BFW from the HRSG steam drum is imported to generate saturated steam at 1600 psig, which is exported to the power island for superheating and power generation. Water circulation between the sulfator steam drum and various exchangers that produce steam is maintained by natural convection. Fluidization air is required for all of these exchanger vessels and is supplied by a slip stream from the booster compressor. Air leaving these exchanger vessels flows through primary and secondary cyclones to remove entrained solids, which are returned to the bed. The cleaned air is then combined with the flue gas from the transport sulfator and fed to the sulfator barrier filter for final solids removal.

The sulfator barrier filter is similar to that used in the gasification system, being of the ceramic candle type. Cleaned flue gas exits the filter at about 1650°F and is sent directly to the gas turbine. Compressed air at 585 psig is used for filter blowback. Air from the blowback air compressor is stored in the blowback air receiver which provides surge capacity required for periodic flow of blowback gas.

Most of the solids leaving the fluidized-bed heat exchangers are recycled to the sulfator for further use. The net make of solids, which consists essentially of ash and spent sorbent, is sent to the high-pressure ash cooler where the hot solids are cooled to about 300°F. This vessel is fluidized with a small amount of air from the booster compressor, which then flows through an ash cooler cyclone for solids removal and is then combined with the flue gas feed to the gas filter. The ash cooling is accomplished via a coil in the fluid bed which preheats boiler feed water from the deaerator and supplies a portion of water to the sulfator steam drum.

Fines from the gas filters are collected in the ash depressurization hoppers together with ash/spent sorbent mixture from the high-pressure ash cooler. Two hoppers are provided per train. These vessels operate in parallel, with one filling while the other is depressurizing and discharging solids. Air from the blowback air receiver is used as pressurization gas. The ash depressurization vent filter removes any entrained dust during the depressurization cycle. After depressurization the solids are cooled by cooling water to 150°F in the ash screw conveyor and are then sent to disposal.

4.4.1.7 Gas Turbine System

The gas turbine configuration for Case 6 is similar to that developed for Case 1. (Refer to Section 4.2.1.8.) Two advanced-design GE MS7001F combustion turbines form the basis of the power island. Figure 4-33 shows the power island configuration for Case 6.

The major flowsheet differences between the GE MS7001F designs for Cases 1, 3, and 5 versus that for Case 6 are the heating value of the fuel produced by the gasifier island and the larger compressor discharge air extraction required for Case 6.

4.4.1.8 HRSG System

The heat recovery steam generator configuration for Case 6 is virtually identical to that developed for Case 1 with the only variations being differences in interface flows with the gasifier island and differences in steam or flue gas extractions for coal drying purposes. (Refer to Section 4.2.1.9.)

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-33 Case 6 Power Island Process Flow Diagram

4.4.1.9 Steam Turbine System

The steam turbine configuration for Case 6 is virtually identical to that developed for Case 1 with the major variation being differences in interface flows with the gasifier island. (Refer to Section 4.2.1.10.)

4.4.2 Plant Layout

The plant layout for Case 6 is illustrated in Figure 4-34. A conceptual plot plan and elevation and isometric views of the gasification train for Case 6 are given in Figures 4-36 and 4-37. Details on plant equipment for this case are provided in Appendix A.

4.5 IGCC PLANT OPERATIONS AND PERFORMANCE (CASES 1, 3, 5, AND 6)

4.5.1 Plant Equipment and Supplies

4.5.1.1 Spare Equipment and Material

Equipment sparing was an important concern in the design of each IGCC case in order to achieve the desired reliability targets. The sparing approach for the power island and balance-of-plant portions of the plants is considered to be conventional and is not discussed in detail here. However, equipment sparing in the gasification island portions of the IGCC plants is dependent on the technology which has been selected, and can have a significant impact on operability and costs. Therefore, gasification island spare equipment was addressed in detail in the study.

There are two categories of gasification island capital spares included in the design cases: inline spare equipment and warehouse spare parts. Inline spares are those pieces of equipment that are permanently connected to the plant and are used only when a failure occurs in the piece of equipment being spared. These items are listed in Table 4-5. Warehouse spare parts in this context are for major components and are in addition to those items included in normal spare part cost factors. Warehouse spares are itemized for the IGCC cases in Table 4-6.

4.5.1.2 Catalysts and Chemicals

Initial catalysts and chemicals are included in the capital cost estimate as a separate line item. These material quantities are shown in Table 4-7 for the various IGCC cases. Annual usages of these items are also provided as a basis for determination of operating costs. The recommendations for spare catalyst quantities are also shown in Table 4-7 as inventory capital. The quantities shown are sufficient to replace a single reactor bed for each indicated catalyst.

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 4-34 Case 6 Overall Site Plan

Table 4-5
Inline Spare Equipment

Items	Numbers of Units			
	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
Coal Feeder (310-V)				
Operational Units	4	2	6	3
Spare Units	2	1	2	1
Ash Feeder (330-V)				
Operational Units	4	2	6	3
Spare Units	2	1	2	1
Recycle Gas Compressor (381-J)				
Operational Units	4	2	6	3
Spare Units	2	1	2	1
Sulfator Vent Fan				
Operational Units	3	2	-	-
Spare Units	3	2	-	-

Table 4-6
Major Component Warehouse Spares

Items	Numbers of Units			
	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
Ceramic Candle Filters				
Operational Elements	3928	1964	3576	3285
Spare Elements	400	300	500	400
Hot Gas Valves (18 inch)				
Operational Valves	32	16	48	-
Spare Valves	2	1	3	-
Hot Solids Valves				
Operational Valves	20	10	30	24
Spare Valves	4	2	6	4
Sulfator Conveyor				
Operational Units	3	2	-	3
Spare Units (602 V)	1	1	-	1
Spare Units (603 V)	1	1	-	1

Figure 4-36 (Sheet 1 of 2) Gasification Train Layout,
Elevation View - Case 6

Figure 4-36 (Sheet 2 of 2) Gasification Train Layout, Isometric View - Case 6

Table 4-7

Catalysts and Chemicals

	<u>Initial Charge</u>	<u>Inventory (Spare)</u>	<u>Annual Usage (85% Capacity)</u>
CASE 1			
Nahcolite (tons)	740	--	1,482
Zinc Ferrite (tons)	1,600	200	346
CASE 3			
Nahcolite (tons)	370	--	741
Zinc Ferrite (tons)	800	200	173
CASE 5			
Nahcolite (tons)	640	--	1,482
Zinc Ferrite (tons)	2,042	170	2,422
Sulfuric Acid (tons)	615	--	--
Molten Sulfur (tons)	53	--	1,025
V ₂ O ₅ Catalyst (liters)	65,600	--	1,684
Slaked Lime (tons)	41	--	3,485

CASE 6

There is no major catalyst or chemical usage for Case 6.

4.5.1.3 Spare Capacity

As part of the design process, reliability analysis was used to determine the equipment sparing requirements in order to achieve availability targets. This process was applicable primarily to the gasification island. The gasifier island equipment components are designed with excess capacity. Should it be necessary to idle or shut down one gasifier train, the capacity of the remaining trains can be increased to make up for a portion of the inoperative gasifier train output. A gasifier train consists of all equipment from the pressurization of feedstocks to the production of clean gas at the exit of the external desulfurizer. The complete train will be isolated and taken out of service if a component failure requires depressurization for repair. This excess capacity and the capacity less one operating train are shown for the IGCC cases in Table 4-8 along with the calculated equivalent availability of the gasification as determined from the reliability analysis described subsequently in Section 4.5.1.4.

Table 4-8
Spare Capacities

<u>Items</u>	<u>Case Number</u>			
	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
No. Gasifier Trains	4	2	6	3
Excess Capacity, %	19	19	2	26
Capacity, Less One Train, %	89	59	85	84
Equivalent Availability of Gasifier Island	95	93	94	96

4.5.1.4 Equipment Reliability

A reliability-availability-maintainability (RAM) analysis was conducted for each of the IGCC plant designs (Cases 1, 3, 5, and 6) to determine equipment reliability. The results of the RAM analyses were compared to an equivalent availability (EA) goal of 85 percent for the large (400 MW) plants and a 65-percent minimum for the small (200 MW) plant, Case 3.

The EPRI computer code, UNIRAM, was used as the analysis tool. UNIRAM was developed for EPRI by ARINC Research Corporation and has been in use since the early 1980s. The basic objective of UNIRAM is to relate failures or combinations of failures at the component level to the resultant curtailment, if any, of the power plant. UNIRAM calculates predicted estimates of equivalent availability (EA) and other performance measures.

The RAM model and analysis have been used not only to provide estimates of expected plant performance but also as an interactive tool in refining the plant designs to assist in optimizing the designs by balancing cost with performance.

Plant RAM Simulation Models

Models for the four different IGCC plant designs are depicted with simplified availability block diagrams (ABD) in Figures 4-37 through 4-40. Some blocks on the ABDs represent more than one component.

Figure 4-37 (Sheet 1 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 1 (Gasification and Power Island)

Figure 4-38 (Sheet 1 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 3 (Gasification and Power Island)

Figure 4-38 Sheet 2 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 3 (Coal, Limestone, and Ash Handling)

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Figure 4-39 (Sheet 2 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 5 (Coal and Ash Handling)

Figure 4-40 (Sheet 1 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 6 (Gasification and Power Island)

Figure 4-40 (Sheet 2 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 6 (Coal, Limestone and Ash Handling)

An ABD is a combination of series and parallel trains, composed of blocks representing major plant equipment or subsystems. If a component in series fails, the entire train in that series becomes unavailable. When a component in parallel fails, however, the other components in parallel with the failed component can carry all (100 percent) or a portion (less than 100 percent) of the plant rated capacity. The percentages printed above the blocks show the percentage of plant capacity that the block can carry.

It should be noted that, although the ABD can not always accurately portray the exact physical layout or flow of the plant, the ABD logic of series and parallel relationships does accurately model the plant availability.

In general, the gasification island portion of the model is based on design drawings provided by and discussions with MWK; the power island, coal, limestone, and ash handling systems, and balance-of-plant auxiliaries were modeled with information based on designs by SCS.

Data

The UNIRAM modeling program requires input data for each component in the form of mean time between failures (MTBF), inverse of the failure rate, and mean downtime (MDT).

Special components, such as material storage bins, provide surge time. Some components upstream (or downstream) of the surge components are affected in that a failure of one of these upstream (or downstream) components has a lesser impact on plant EA because the surge time provides time for repair of the upstream (or downstream) components before the plant must be derated or shut down.

Other inputs to the UNIRAM model are expected annual planned outage time and reserve shutdown hours, if any. Surge time, planned outage time and reserve shutdown time are discussed in the following assumptions section. Note that some component input data have been adjusted by UNIRAM to account for any surge time allowances.

The sources of the data were primarily published EPRI reports. Complete MTBF and MDT data for all components were provided for each gasification design by MWK. The MTBF and MTR data for the gasification island were derived from several EPRI and industrial data base sources, and the values presented are typical for actual components of similar design and service.

Several points can be made with respect to the data for the gasification island:
(1) Where solids valves and pressure valves are used, the solids valve has a shorter MTBF due to seat erosion; (2) The 431L filter has a shorter MTBF than

similar filters because of very fine particulate present in this system; (3) MTBF data given for the 410D and 510D reactors reflect the expected sorbent life for which a scheduled change-out would be planned; (4) the 410D valves have a shorter MTBF than similar valves because of the presence of chloride in the gas; (5) the 510-D regeneration valves likewise have shorter MTBF because of SO₂ in the gas stream.

Most of the components in the power island and balance-of-plant portion of the plant designs used EPRI data from the following two reports:

- Component Failure and Repair Data: Gasification-Combined-Cycle Power Generation Units (EPRI AP-2205)
- Component Failure and Repair Data for Coal-Fired Power Plants (EPRI AP-2071).

Data for a few components were developed using SCS engineering estimates and/or manufacturer information.

Assumptions

Some of the major assumptions used in the UNIRAM analysis are highlighted here. Each plant is assumed to be baseloaded, with no reserve shutdown hours. On average, the IGCC plant is assumed to require 412 hours per year for planned overhauls, which reduces plant EA by 4.7 percent. The gross output of the plant varies with the design case under study, and is presented elsewhere in this report. The plant has been designed without backup fuel capability.

With regard to materials handling, it is assumed that the coal handling magnetic separator can be bypassed. The coal handling silos are sized to provide 14 hours of surge time, while the coal handling and limestone handling holding bins provide 2 hours surge time each. Except for the pneumatic limestone conveyers, the limestone handling facility was not included in the model. Based on design information, the limestone handling system is assumed to have sufficient redundancies and surge capacities to consider it available 100 percent of the time. The milled limestone silo has 24-hour surge capacity. In addition, milled limestone can be trucked in on an emergency basis. Twenty-four hours is considered to be sufficient lead time for this eventuality. The ash handling silo capacity permits 20 hours of surge time for components downstream of the silo. The ash collection surge bin provides one hour surge time.

In the gasification area, the gasification feed surge drums have capacity for 1½ hours of surge time. The zinc ferrite reactor sorbent changeout is assumed to occur during planned plant overhauls. There would, therefore, be no unplanned unavailability associated with this component.

It is assumed that the sulfator barrier filters (applicable to Case 6 only) would be replaced annually during the planned maintenance overhaul. Also, if a filter element were to fail in service, a blowback operation would be performed to mitigate the impact on the remainder of the plant. Therefore, the availability impact on the plant caused by the sulfator barrier filters was assumed to be insignificant for Case 6.

The oxygen plant utilized in Case 5 is designed to have 24-hours storage (surge time). The sulfuric acid plant for Case 5 has very high reliability. The redundancies are sufficient to assume no negative impact on availability.

In the power island area, the HRSGs include deaerators and feedwater heaters for each case. It is assumed that the boiler feedwater preheater in the booster air compressor system can be bypassed for each of the IGCC plants.

Results

The plant design configurations, as represented by the ABDs, were modeled with the UNIRAM computer software program. The major output of the analysis is the equivalent availability, which can be defined as the percentage of the plant's maximum energy production which was available during a time period. The expected equivalent availabilities for the four IGCC plants are as follows:

Equivalent Availability

	<u>Target</u>	<u>Actual</u>
Case 1	85%	84.40%
Case 3	>65%	82.97%
Case 5	85%	83.96%
Case 6	85%	85.27%

Compared to the EA targets set out at the beginning of the study, three of the four IGCC cases (1, 3, and 6) have met the goal. The other case is within approximately 2 percentage points of the goal which is considered acceptable for the present analysis.

As stated previously, the annual planned overhaul of 412 hours reduces the EA by 4.7 percentage points. This reduction is included in the resultant EA figures given above. With the plant scheduled to be in service the balance of the year, the expected annual generation can be estimated. For example, in Case 1, a 400 MW plant, the expected generation is calculated as:

$$(400 \text{ MW}) (8760 \text{ HOURS/YEAR}) (84.40\%) = 2,957,376 \text{ MWHR/YEAR}$$

Additional RAM information is provided for each of the IGCC cases in Table 4-9.

Critical components

For the IGCC cases under discussion, the three component types causing the most forced lost generation were generally the combustion turbine generators (CTG), heat recovery steam generators (HRSG) and the steam turbine generator (STG). Together, these components accounted for 49 percent to 60 percent of the equivalent forced outage hours (EFOH), depending on the case.

Although the gasification island had no particular components contributing a large percentage to the total forced outage problems, the gasification island, in total, contributed significantly because it has many components. The gasification island accounted for 36 percent to 48 percent of the EFOH. The sulfator system (part of the gasification island) accounted for 40 percent to 52 percent of the gasification island's forced outage problems. The three sulfator components causing the most lost generation were the LASH receiver, the sulfator, and the sulfator's heat exchanger coil. The exception was Case 5 which has no sulfator. In Case 5, the product gas cooler in the gasification heat recovery system was the second largest contributor to plant EFOR, second only to the CTG. The product gas cooler has a mean down time (MDT) of 193 hours in Case 5 compared to a MDT of only 60 hours in Cases 1, 3 and 6.

A complete list of all components ranked by criticality is included in Appendix C for all cases. A summary listing is provided in Table 4-10.

Bypass Option

The option of installing HRSG bypass stacks was investigated. Each CTG has a dedicated HRSG which recovers CTG exhaust heat for further electricity production by expanding the steam through a steam turbine. Without bypass stacks, when an HRSG fails, its corresponding CTG must be shut down. With a bypass stack, the CTG can continue to operate and produce electricity during an HRSG forced outage. The CTG exhaust will be sent up the stack. If the STG fails, both HRSGs can be bypassed, and the CTGs operated.

To evaluate the effect, the Case 1 model was modified to include HRSG bypasses. The EA improved two percentage points from 84.40 percent to 86.40 percent.

Table 4-9

Plant Wansley IGCC Projected Performance Indexes

	<u>EA(%)</u>	<u>A(%)</u>	<u>EFOR(%)</u>	<u>FOR(%)</u>	<u>Days Per Year At Maximum Output</u>
Case 1	84.40	93.06	11.43	2.35	236
Case 3	82.97	88.12	12.93	7.53	279
Case 5	83.96	92.98	11.89	2.44	215
Case 6	85.27	92.28	10.52	3.17	253

Where: EA = Equivalent availability
A = Availability
EFOR = Equivalent forced outage rate
FOR = Forced outage rate

Table 4-10

IGCC Cases - Critical Components

	<u>Percentage of total EFOR</u>			
Component	Case 1	Case 3	Case 5	Case 6
CTG	27	25	26	30
HRSG	15	14	14	17
STG	12	10	11	13
Gasification Island/ Sulfator System	42/22	48/20	—	36/14
Gasification Island/ Product Gas Cooler	—	—	44/15	—

Availability measures the percentage of time during which the plant is capable of operating (at any level of generation). Since there is no reserve shutdown time expected, the annual service hours can be estimated using the availability. In Case 1 the expected annual service hours are:

$$(8760 \text{ hours/year}) (93.06\%) = 8152 \text{ service hours per year}$$

The EFOR represents the percentage of the plant maximum generation that was desired during a certain time period but was not available because of forced events (both full forced outages and forced deratings).

The FOR is the percentage of time that the plant's service was desired within a certain time period but was unavailable due to full forced outages only.

HRSG bypasses do, however, have some negative impacts which should be considered before deciding to install bypasses. The initial cost, the negative heat rate impact, the maintenance cost, and maintenance work conditions are all factors to consider. Efficiency losses may even occur during normal HRSG operation because the bypass stacks may leak. For these reasons, bypass of the HRSG was not included in the cases presented in this study.

Sensitivity to Changes in Planned Overhaul Maintenance Schedules

The basic assumption presented in the analysis assumes an aggressive, optimistic planned overhaul schedule. The 412 hours per year or 17 days per year (4.7-percent planned outage factor) is consistent with the assumption used in the earlier EPRI Plant Wansley IGCC study. The "highest availability plan" was developed by Fluor Engineers and Constructors. It is a 10-year plan which assumes shutting down the entire plant once per year to perform the longest duration maintenance activity and all shorter duration activities at the same time. Sufficient personnel, materials, and parts are assumed to be available.

This optimistic plan is presented along with four other more conservative overhaul plans in the EPRI Report AP-5276, Availability Analysis of an Integrated Gasification - Combined Cycle, June 1987. These five plans achieve planned outage factors (POF) ranging from 4.7 percent to 8.8 percent. As a sensitivity check, the EAF for the IGCC cases in the current study were recalculated using the most conservative planned overhaul schedule (8.8 percent POF) developed by ARINC Research Corporation. This conservative plan averages 32 days per year overhaul downtime, which approximates the projections used for some large coal-fired steam plants in The Southern Company. In every case, the EAF falls about 3.6 percentage points. The results of this sensitivity check are presented in Table 4-11.

Table 4-11

**Effect of Planned Overhaul Schedules on
Equivalent Availability**

	Optimistic Plan (17 days per year)	More Conservative Plan (32 days per year)
Case 1	84.40%	80.77%
Case 3	82.97%	79.41%
Case 5	83.96%	80.35%
Case 6	85.27%	81.61%

Since every case has a different design configuration, overhaul maintenance plans should be tailored to each individual plant design. This was not considered. A blanket assumption for overhaul downtime was used for all cases in the study.

Other factors affecting the choice and impact of overhaul plans are the availability of backup fuel, equipment bypass capability, and whether the overhaul plan is staggered or constitutes a complete yearly plant shutdown.

4.5.1.5 Gasifier Controls/Turndown

Important aspects of IGCC plant design are the control strategy and turndown capability of the plant. Detailed instrumentation and controls strategies were not developed for the current study. However, the basic control concept is outlined as follows.

The gasifier is controlled by a load demand signal received from the power island. Coal/limestone feed rate and process air flow to the gasifier are adjusted to meet the load demand. The gasifier temperature is modulated by the adjustment of steam flow rate to maintain a near constant temperature. The gasifier bed level is maintained within a tight band by adjustment of the ash withdrawal rate. For Case 5 where the gasifier operation is without limestone, the ash withdrawal rate is set by ash annulus temperature. System pressure is maintained by a control valve in the gas delivery line to the power island. Recycle gas to the gasifier grid and annulus is flow controlled.

The outlet temperature of the product gas cooler is controlled by an internal gas flow tube side bypass valve. The gas outlet temperature of the recycle gas cooler is maintained by controlling flowrate of boiler feed water from the deaerator.

The steam drum control is typical three-element control on the drum level, boiler feed water supply rate, and steam production rate.

4.5.2 Process Performance

4.5.2.1 Heat and Material Balances

Heat and material balances were conducted separately for the gasification island and power island of the plant. The material balances for the gas island conducted by MWK were performed using the ASPEN process simulation computer model which takes into account the proper energy flows in deriving the material balance. The power island used a SYNTHA thermal modeling computer program to determine the heat and material balances based on the interface points with the gas island.

The overall plant material balances for Cases 1, 3, 5, and 6 are summarized in Table 4-12. The material descriptions cover all four cases and thus some items do not apply to all cases.

4.5.2.2 Gasifier Island Performance and Interfaces

Gasifier island performance and interface parameters are summarized in Table 4-13, which shows the quantities of plant feeds to the process and the gasifier island products produced. Performance of the various cases are compared by cold gas efficiency, carbon conversion, and sulfur removal in this table. Interface parameters between the gasification and power islands are summarized in Table 4-14. These parameters are reflected in the material balances presented in earlier sections.

4.5.2.3 Power Island Performance

Four major areas were focused on for developing of the thermal cycle: (1) combustion turbine configuration, (2) HRSG configuration, (3) steam turbine island, and (4) interface points with the gasifier island. A thermal model of each of the cases was developed using the SYNTHA thermal modeling computer program. The thermal model was based on the equipment configurations and interface flows previously described. The following major design points were utilized in the development of the thermal model:

Table 4-12

Overall Material Balances - IGCC Cases

	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
<u>Material In, LB/HR</u>				
Limestone (Dried)	87,733	43,866	---	35,936
Coal (As Received)	336,573	168,287	335,837	285,212
BFW Makeup	460,924	230,462	800,950	102,883
Combustion Air to Gas Turbine	4,909,830	2,454,923	4,876,532	5,033,400
Cooling Tower Makeup	3,485,164	1,742,582	2,643,917	3,530,231
Air to Sulfator	195,650	97,825	---	120,902
Filter Blowback Air	---	---	---	3,328
Air Compressor Feed to Gasifier	1,073,957	536,979	1,106,169	820,313
Nitrogen to Coal Preparation	---	---	35,184	---
Coal Drying dilution Air	213,665	109,224	71,236	---
Total Material In	10,768,496	5,384,148	9,869,825	9,932,205
<u>Material Out, LB/HR</u>				
HRSG Stack Gas to Main Stack	6,740,748	3,370,385	6,654,268	6,322,163
Coal Drying Gas to Local Stack	424,339	212,060	126,973	---
Solids to Disposal	102,576	51,288	45,014	66,455
ZnFe Regeneration Gas to Acid Plant	---	---	386,752	---
Vent Gas to Atmosphere	---	---	---	1,513
Vent Fuel Gas to Flare	1,261	631	432	---
Boiler Blowdown	14,448	7,224	14,091	11,411
Cooling Tower Blowdown, Evap., Drift	3,485,164	1,742,582	2,643,917	3,530,231
Total Material Out	10,768,536	5,384,170	9,871,447	9,931,773
<u>Closure</u>	100.00%	100.00%	100.02%	100.00%

Table 4-13
Gas Island Design Basis Performance Comparison

	Case 1	Case 3	Case 5	Case 6
FEEDS				
Coal Feed (T/Day, After Drying)	3,752	1,896	3,783	3,213
Limestone Feed (T/Day)	1,053	526	—	431
Air Supply (T/Day)	12,688	6,444	13,274	11,334
Steam (MMlb/Day)				
450 lb	5,357	2,678	7,386	1,092*
850 lb	—	—	—	—
Solids Waste (T/Day)	1,231	615	540	795
Oxygen (T/Day)	—	—	121	—
PRODUCTS				
Fuel Gas (LHV, MMBtu/Day)	63,336	31,868	62,184	59,757
Steam (MMlb/Day)				
1600 lb SAT.	12,447	5,224	12,824	6,160
850 lb SAT.	—	—	—	—
Sulfuric Acid (T/Day)	—	—	280	—
PERFORMANCE				
Cold Gas Efficiency- LHV (%)	69.7	69.7	68.1	77.6
Carbon Conversion (%)				
Gasifier Only	96.5	96.5	95.2	98.2
Gas Island	99.9	99.9	95.2	99.9
Sulfur Removal (%)				
Gasifier Only	86.15	86.5	—	—
Gas Island	99.4	99.4	99.4	99.2

* 350 psig steam supply pressure is adequate for Case 6.

Table 4-14
MWK/SCS Interface Parameters

GASIFIER/POWER ISLAND INTERFACE		Case 1	Case 3	Case 5	Case 6
Coal (Sized)	Flow, lb/hr	316,978	187,888	316,284	287,767
	Temp, deg °F	200	200	200	200
	Pressure	Amb	Amb	Amb	Amb
Limestone (Sized)	Flow, lb/hr	87,733	43,866	—	36,936
	Flow, lb/hr	—	—	—	—
Process Air	Flow, lb/hr	1,073,987	536,978	1,106,189	844,543
	Temp, deg °F	760	760	768	740
	Pressure, psia	200	200	180	188
Product Gas	LHV, Btu/SCF	87.4	87.4	87.2	137.9
	Flow, MMSCu/hr (LHV)	2,639	1,320	2,691	2,490
BFW From Power Island	Flow, lb/hr	812,373/488,088	406,136/243,044	937,291/671,888	453,789/308,953
	Temp °F	600/240	600/240	600/240	600/240
	Pressure, psia	1,860	1,860	1,860	1,860
BFW To Power Island	Flow, lb/hr	10,373/250,719	5,186/126,390	10,897/258,533	5,134/242,121
	Temp °F	606/428	606/428	606/438	606/348
	Pressure, psia	1,800	1,800	1,800	1,800
Steam From Power Island	Flow, lb/hr	56,889/388,707	28,348/184,854	384,977/230,510	81,037
	Temp °F	850/600	850/600	850/600	850
	Pressure, psia	450/450	450/450	450/450	350
Steam To Power Island	Flow, lb/hr	1,037,288	518,635	1,065,685	513,367
	Temp °F	606	606	606	606
	Pressure, psia	1,800	1,800	1,800	1,800
Solids to Disposal, lb/hr		102,576	51,288	45,014	66,455
Sulfator Flue Gas to Power Island	Flow, lb/hr	188,784	94,377	—	113,568
	Temp °F	1,400	1,400	—	1,818
Vent Gas to Flare, SCFH		347	174	120	—

For case 5, a blowdown stream of 1421 lb/hr at 400 psia and 445°F is also exported to the power island.
For case 6, 633 gpm water stream that contains 2.3 wt% H₂SO₄ is also exported to the power island.

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

- Reheat steam turbine design based on General Electric 1450 psig, 1000°F/1000°F design with two flow IP/LP section and 23-inch last stage blades - cold reheat pressure 350 psi
- HRSG - dual pressure with 1600 psi HP drum and 350 psi IP drum, integral deaerator and feedwater heating coil - overall 1.5-percent heat loss per section, 1-percent HP and IP drum blowdown
- Condenser pressure - 3.5-inch HgA
- Deaerator pressure - 25 psi to provide 240°F feedwater temperature
- Boiler feedpump discharge pressure - 1650 psi to allow for 13-percent pressure drop through HRSG
- LP turbine exhaust losses based on 23-inch LSB with 32.9 sq. ft. annulus area.

Many model iterations were run to optimize the power output of the steam turbine. The two major points which were used to establish optimum steam flows were the HP evaporator pinch point and the HRSG exhaust temperature. Based on the composition of the exhaust gas, 280°F was set as the minimum exhaust gas temperature in order to minimize acidic deposition and avoid stack corrosion. The HP evaporator pinch point was varied with lower pinch points naturally providing higher overall power production. A value of 35°F was established as the design value for the HP evaporator pinch point based on a tradeoff between improved performance and higher capital cost.

Gas Turbine Performance

The gas turbine performance was provided by General Electric based on the applicable site conditions for Plant Wansley. The GE MS7001F design would be virtually identical to the design for a natural gas firing GE MS7001F with the exceptions of special considerations for the bleeding of compressor discharge air to the gasifier island and adjustments for high-temperature, low-Btu fuel gas flow to the combustor. Both of these areas require further research and testing before the GE MS7001F could be considered commercially operational for this application.

One important factor employed by General Electric in developing the performance data for the GE MS7001F was the use of the flat rating concept. As combustion turbines are constant volume machines, their output is highly dependent on the density of air which is greatly affected by changes in ambient temperature. Lower ambient temperatures result in higher output due to higher air densities and therefore more mass flow through the gas turbine. Higher ambient temperatures result in lower output due to decreases in air density.

Table 4-15 shows design information produced by General Electric regarding the calculated performance of the MS7001F using the coal gas compositions developed for Cases 1 and 3. Table 4-16 shows design information produced by General Electric regarding the calculated performance of the GE MS7001F using the coal gas for case 5. Table 4-17 shows design information produced by General Electric regarding the calculated performance of the GE MS7001F using the coal gas for case 6. Table 4-18 summarizes the performance data for the combustion turbines for all four IGCC cases at the 90°F design point. The gas turbine performance data is based on operation under the flat rating design scheme previously described.

Steam Turbine Performance

Using the SYNTHA thermal modeling program, a computer model of the steam cycle (HRSG and steam turbine) was developed. Using the gas turbine exhaust conditions developed by General Electric, corresponding steam turbine performance data were developed based on the HRSG(s) receiving the exhaust flow from the combustion turbine(s). The thermal model was developed using the gasifier interface flows previously described. Table 4-19 shows the summary of the performance data developed for the steam turbine.

Station Service Calculations

Based on the thermal cycles developed for Cases 1, 3, 5, and 6, major equipment was sized and station service power requirements determined. Table 4-20 provides the breakdown of station service values for the complete IGCC plants including the power island, gasifier island, and coal and limestone handling areas.

Net Power Island Performance

Table 4-21 shows the calculated net power output for Cases 1, 3, 5, and 6. The net power values are merely the gross power outputs of the power island less the station service requirements of the power island, gasifier island, and coal/limestone handling facilities. The fuel consumption and consequently the heat rates shown in Table 4-21 are based on the design flow rates of the coal to the gasifier island on a higher heating value (HHV) basis (12,469 Btu/lb as prepared - after drying).

4.5.2.4 Environmental Performance

The environmental performance for Cases 1, 3, 5, and 6 is discussed in this section in terms of air emissions, and solid waste production for each IGCC plant. The solid waste includes spent limestone, ash, nahcolite from the chloride guard beds and spent zinc ferrite sorbent. Environmental compliance and impacts are treated in Section 9 of the study.

Table 4-15

GE MS7001F Performance — Flt Rating
Cases 1 and 3

Ambient Temperature (°F)	20	59	90
Relative Humidity (%)	60	60	60
GT COMPRESSOR AIR TO GASIFIER ISLAND			
Ratio to fuel gas	0.5869	0.5869	0.5869
Mass flow (lb/sec)	149.1	149.1	149.1
Temperature (°F)	635	695	760
Pressure (psia)	185	190	200
FUEL GAS FLOW TO GT			
Heating Value (LHV, Btu/scf)	87.4	87.4	87.4
Mass flow (lb/sec)	254.1	254.1	254.1
Temperature (°F)	1000	1000	1000
Pressure (psia)	285	290	295
GENERATOR OUTPUT (kW)	160,410	160,410	155,820
TURBINE EXHAUST FLOW			
Mass Flow (lb/sec)	900.8	918.9	935.5
Temperature (°F)	1093	1093	1098

RELEVANT SITE CONDITIONS

Site elevation	725 feet
Barometric pressure (sea level)	14.7 psia
Inlet air pressure loss	3.5 in. H ₂ O
Exhaust pressure loss	15 in. H ₂ O
Steam injection for NO _x control	none
Supplementary fuel	none
Generator power factor	0.9
GT auxiliary load (not included above)	775 kW

Table 4-16

**GE MS7001F Performance — Flat Rating
Case 5**

Ambient Temperatures (°F)	20	59	90
Relative Humidity (%)	60	60	60
CT COMPRESSOR AIR TO GASIFIER ISLAND			
Ratio to fuel gas	0.6222	0.6222	0.6222
Mass flow (lb/sec)	153.5	153.5	153.5
Temperature (°F)	630	690	755
Pressure (psia)	185	185	190
FUEL GAS FLOW TO CT			
Heating Value LHV, Btu/scf	87.9	87.9	87.9
Mass flow lb/sec	246.8	246.8	246.8
Temperature (°F)	1000	1000	1000
Pressure (psia)	280	285	290
GENERATOR OUTPUT (kW)	157,190	157,160	152,810
TURBINE EXHAUST FLOW			
Mass Flow (lb/sec)	890.2	907.2	923.8
Temperature (°F)	1097	1095	1100
RELEVANT SITE CONDITIONS			
Site elevation	725 feet		
Barometric pressure (sea level)	14.7 psia		
Inlet air pressure loss	3.5 in. H ₂ O		
Exhaust pressure loss	15 in. H ₂ O		
Steam injection for NO _x control	none		
Supplementary fuel	none		
Generator power factor	0.9		

Table 4-17

GE MS7001F Performance — Flat Rating
Case 6

Ambient Temperature (°F)	20	59	90
Relative Humidity (%)	60	60	60
GT COMPRESSOR AIR TO GASIFIER ISLAND			
Ratio to fuel gas	0.816	0.816	0.816
Mass flow (lb/sec)	131.2	131.2	131.2
Temperature (°F)	610	670	735
Pressure (psia)	170	175	180
FUEL GAS FLOW TO GT			
Heat Value (LHV, Btu/scf)	138	138	138
Mass flow (lb/sec)	160.8	160.8	160.8
Temperature (°F)	1000	1000	1000
Pressure (psia)	260	265	270
GENERATOR OUTPUT (kW)	133,730	134,035	131,585
TURBINE EXHAUST FLOW			
Mass Flow (lb/sec)	840.2	859.9	875.8
Temperature (°F)	1143	1138	1137
RELEVANT SITE CONDITIONS			
Site elevation	725 feet		
Barometric pressure (sea level)	14.7 psia		
Inlet air pressure loss	3.5 in. H ₂ O		
Exhaust pressure loss	15 in. H ₂ O		
Steam injection for NO _x control	none		
Supplementary fuel	none		
Generator power factor	0.9		
GT auxiliary load (not included above)	775 kW		

Table 4-18

**Combustion Turbine Performance Summary
90°F Design Point**

	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
Power Output per CT (kW)	155,820	155,820	152,810	131,585
Total CT Power Output (kW)	311,640	155,820	305,620	263,170
CDP Conditions				
Temperature (°F)	760	760	755	735
Pressure (psia)	200	200	190	180
Exhaust Temperature (°F)	1098	1098	1100	1137
Exhaust Flow (lb/sec)	935.5	935.5	892.2	875.8

Table 4-19

Steam Turbine Performance Data - IGCC Cases

	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
Throttle Flow (lb/hr)	1,344,245	672,123	1,293,662	1,048,665
Reheat Flow (lb/hr)	1,025,827	512,914	801,443	936,954
Makeup Flow (lb/hr)	460,924	230,462	800,950	102,833
HRSG exhaust Temp (°F)	280.1	281.3	280.1	280.1
Steam Turbine Gross Power Output (kW)	182,190	94,008	153,390	178,410
Total CT Power Output (kW)	311,640	155,820	305,620	263,170
Total Gross Power Output (kW)	493,830	249,828	459,010	441,580

Table 4-20

**Detailed Plant Station Service Values - IGCC Cases
(All Values in Kilowatts)**

	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
Gasifier Island				
Booster Air Compressor	17,341	8,670	19,873	17,476
KRW Gasification	1,004	502	1,369	729
Recycle Gas Compression	1,714	857	1,820	254
Sulfuric Acid Plant	-	-	1,195	-
Oxygen Plant	-	-	1,500	-
Sulfation	1,281	641	-	53
Miscellaneous	320	160	350	250
Total Gasifier Island	<u>21,660</u>	<u>10,830</u>	<u>26,107</u>	<u>18,762</u>
Power Island				
HRSG Feedpumps	3,420	1,710	3,683	2,742
Condensate Pumps	144	72	155	115
Circ. Water Pumps	2,027	1,014	1,581	2,083
Cooling Tower Fans	1,600	800	1,217	1,645
Vacuum Pumps	400	200	400	400
Service Water Pumps	227	114	227	227
Make-up Pumps	19	9	33	5
Service Air System	149	75	149	149
Gas Turbine Aux.	1,550	775	1,550	1,550
Steam Turbine Aux.	1,500	750	1,500	1,500
Water Treatment	320	160	544	77
Powerhouse	15	10	15	15
Service Building/ Control Room	18	12	18	18
Total Power Island	<u>11,389</u>	<u>5,701</u>	<u>11,072</u>	<u>10,526</u>
Coal & Limestone Handling	<u>2,372</u>	<u>1,621</u>	<u>1,664</u>	<u>2,765</u>
Total	35,421	18,152	38,843	32,053

Table 4-21

**Overall Performance Summary for IGCC
Cases 1, 3, 5, and 6 (90°F)**

	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
Gross Power Output (kW)	493,830	249,828	459,010	441,580
Gasifier Island Station Service (kW)	21,660	10,803	26,107	18,762
Power Island Station Service (kW)	11,389	5,701	11,072	10,526
Coal/Limestone Handling Station Service (kW)	2,372	1,621	1,664	2,765
Net Plant Power Output (kW)	458,409	231,703	420,167	409,527
Fuel Consumption (MMBtu/hr - HHV)	3,940	1,970	3,931	3,339
Net Plant Heat Rate (Btu/kWh - HHV)	8,595	8,503	9,357	8,153

Air Emissions

The air emission rates for major pollutants are listed in Table 4-22, which also includes the NSPS standards outlined in Section 3.2.3 for comparison.

For all of the cases sulfur removal is approximately 99 percent. Table 4-23 lists the distribution of sulfur in the product streams for Cases 1, 3, 5, and 6. For Cases 1, 3, and 6, most of the sulfur is removed in the solid waste which is disposed of in a landfill. For Case 5, only about 8 percent of the sulfur is removed in the solid waste stream. The remainder of the sulfur in Case 5 is processed to produce sulfuric acid.

Solid Waste

The solid wastes from the IGCC plants include spent limestone and ash from the sulfator for Cases 1, 3, and 6; ash and fines for Case 5, nahcolite from the chloride guard bed in Cases 1, 3, and 5; and spent zinc ferrite sorbent in Cases 1, 3, and 5. The quantities of solid wastes generated are shown in Table 4-24.

The solid material that is discharged from the gasifier island is extracted from the sulfator (except for Case 5). The sulfator discharge contains a small portion of calcium sulfide which is of concern because it can release hydrogen sulfide gas if it is exposed to an acid solution. The material does, however, meet the requirements for nonhazardous waste, and the solid wastes pass the EP toxicity test and are suitable for dry landfill.

Table 4-22

Air Emission Rates for Major Pollutants — IGCC Cases

	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>	<u>NSPS</u>
<u>NO_x</u>					
lb/MMBtu	0.167	0.168	0.144*	0.202	---
ppmv at 15% O ₂ (dry)	65.6	65.9	57.3*	76.0	75.0
<u>SO₂</u>					
lb/MMBtu					
HRSG Stack	0.026	0.026	0.025	0.033	---
Coal Prep Stack	0.025	0.024	---	---	---
Total	0.051	0.050	0.025	0.033	1.2

Particulates:

Particulate removal is by cyclones and ceramic filters. The removal efficiency is effectively 100 percent. Modeling was conducted at the maximum concentration allowed in the combustion turbine, although the actual concentration is less.

- * SCR used to reduce emission of NO_x below target; emission rate without SCR is 0.721 lb/MMBtu (287 ppmv)

Table 4-23

Overall Sulfur Balance for IGCC Cases (1, 3, 5, and 6)

	<u>Case 1</u>	<u>Case 3</u>	<u>Case 5</u>	<u>Case 6</u>
<u>Sulfur In. t/d</u>				
Sulfur in feed coal	101.7	50.9	101.5	86.1
<u>Sulfur Out. t/d</u>				
Solids to landfill	100.5	50.3	7.9	85.4
Sulfuric acid plant	---	---	93.0	---
Sulfator flue gas	0.6	0.3	0.0	0.0
HRSG flue gas	<u>0.6</u>	<u>0.3</u>	<u>0.6</u>	<u>0.7</u>
Total Out	101.7	50.9	101.5	86.1

Table 4-24

Total Solid Waste Quantities Generated — IGCC Cases

<u>CASE</u>	<u>Tons/day</u>
1	1,231
3	615
5	540
6	797

5.0 PLANT DESCRIPTIONS - ISTIG CASES 2 AND 4

5.1 SUMMARY OF ISTIG STUDY CASES

5.1.1 Background and General Discussion

Cases 2 and 4 of the study evaluate a power cycle referred to as the Intercooled Steam-Injected Gas Turbine (ISTIG) design. The major motivations behind the development of these cycles are: (1) to take advantage of the higher efficiencies and reliability of gas turbines developed for use in modern jet aircraft; i.e., the aeroderivative turbines; and (2) to eliminate the steam bottoming cycle from the gas turbine power plant in order to reduce capital cost, simplify the design, and produce a highly modular plant.

ISTIG Evolution

The ISTIG design has evolved from the GE LM-5000 STIG design which was developed directly from the simple-cycle GE LM-5000. The simple-cycle version of the LM-5000 is a 33-MW machine composed of a two-shaft gas generator section and a rear drive power turbine section. The gas generator section is a dual rotor unit consisting of an air inlet section, a five-stage low-pressure compressor, a 14-stage variable-geometry high-pressure compressor, an annular axial swirler combustor section, a two-stage air-cooled high-pressure turbine, and a single-stage low-pressure turbine. The low-pressure compressor and low-pressure turbine are connected by one shaft with the LP turbine providing the power to drive the LP compressor, and the high-pressure compressor and high-pressure turbine are connected by the second shaft with the HP turbine providing the power to drive the HP compressor. The power turbine is a three-stage configuration with a rear drive designed for connection to an electric generator through a flexible coupling and is driven by the exhaust gases from the gas generator section.

The STIG cycle utilizes a two-pressure unfired heat recovery steam generator (HRSG) to produce steam for injection into the LM-5000 at various points. High-pressure steam is injected into the fuel nozzles at the combustors and into the compressor discharge bleed air port (CDP). Low-pressure steam is injected just ahead of the low-pressure turbine rotor. Net power output is approximately 47 MW utilizing 80,000 lb/hr of HP (700 psi) steam and 72,000 lb/hr of LP (210 psi) steam with the HRSG in the unfired mode. A full STIG LM-5000 power plant has been installed at Simpson Paper Company in Ripon, California (Reference 4).

Although directly derived from the LM-5000 and LM-5000 STIG, the design basis of the LM-8000 ISTIG unit is the new CF6-80C2 high bypass turbofan aircraft engine. The ISTIG design involves removal of the large front-mounted bypass fan

with further efficiency improvements realized by utilizing an intercooler between the low-pressure and high-pressure gas generator compressor sections. The effect of the intercooler is to increase the flow capability and efficiency of the high-pressure compressor (HPC) section due to the reduced temperature of the air entering the HPC inlet. This also results in a lower temperature for the HPC discharge air which is used for hot section cooling and therefore allows for higher firing temperatures while still maintaining acceptable bulk metal temperatures in both the combustors and the turbine blading. This higher firing temperature results in increased power output and cycle efficiency.

The ISTIG cycle utilizes three pressure levels of steam injection with high-pressure steam injected into the combustor fuel nozzles, low-pressure steam injected just ahead of the low-pressure turbine rotor, and even lower-pressure steam injected into the last stage of the power turbine. The total required steam flow is about 210,000 lb/hr.

Table 5-1 provides a basic performance comparison between the LM-5000, the LM-5000 STIG, and the LM-8000 ISTIG based on operation with natural gas firing.

Effects of Low-Btu Coal Gas on ISTIG Design

Although the overall LM-8000 design is strictly of a conceptual nature, previous analyses by GE have shown that the LM-8000 ISTIG configuration is readily adaptable (at least in theory) to operation on low-Btu coal gas. The 1986 report "System Analysis of Simplified IGCC Plants" by J. C. Corman of GE included an analysis of an ISTIG unit integrated with an air-blown Lurgi fixed-bed gasifier (Reference 5). The heating value of the gas used in that study was 1747 Btu/lb (LHV), while the heating value of the gas in the KRW-based system of the present study is approximately 1440 Btu/lb (LHV). Discussions with design personnel at GE have indicated that the LM-8000 combustion system could be expected to operate with the lower Btu fuel, but many tests would be required before the system could be declared commercially viable.

Based on the currently proposed design of the LM-8000, the major effect of the low-Btu gas, as compared to operation on natural gas, is an increase in the incoming mass flow to the combustor and therefore an increase in mass flow to the turbine section. This effect increases net power production of the ISTIG unit, but also limits the steam injection flows that can be admitted into the ISTIG below the levels that would be anticipated with a natural-gas-fired unit because the unit has reached its mass flow limit.

Table 5-1
Comparison of LM-5000 to LM-8000 (Natural Gas-Fired)

	LM-5000 <u>Simple Cycle</u>	LM-5000 <u>STIG</u>	LM-8000 <u>ISTIG</u>
Power (MW)	36	50	114
HPC Air Flow (lb/sec)	272	272	353
HPC Inlet Temperature (°F)	220	220	73
HPC Outlet Temperature (°F)	971	971	726
HPC Outlet Pressure (psi)	394	394	500
HP Turbine In Temperature (°F)	2,110	2,100	2,500
Pressure Ratio	26.8	26.8	34.0
Turb Exhaust Temperature (°F)	792	780	887
Turb Exhaust Flow (lb/sec)	272	322	419
Steam Injection Flow (lb/sec)	0	40.0	60.5

5.1.2 Development Status

As indicated in the discussion above, the GE LM-8000 ISTIG is a study concept which has not been configured in hardware. The concept was put forward by GE in recent years with the goal of attracting sufficient industry support to proceed with commercialization. Some 10 to 15 orders were required by GE in order to complete development of the concept. However, the required level of industry commitment was apparently not forthcoming. GE has recently announced that it no longer supports the LM-8000 design and has instead proposed an LM-X000 package which combines elements of several GE aeroderivative products. This change occurred during the preparation of the current study and therefore could not be accommodated herein.

It is the understanding of SCS that commercialization of an ISTIG concept based on a GE aeroderivative machine (whether the LM-8000 ISTIG or its most recent variant) would require a substantial development effort, consisting primarily of the component addition and packaging needed to convert the base aircraft engine to a complete stationary power plant. In addition to overall system development, specific areas of development include a new LP compressor, a new LP turbine and a new power turbine; an intercooler to provide interstage cooling of the compressor air; and modifications to the gas combustor and the CF60-80C2 core engine. The involvement of one or more third party equipment suppliers would be needed to fully develop the LM-8000 for power production.

5.1.3 Design Approach

The design approach for the gasification islands for Cases 2 and 4 was similar to that for the IGCC cases described in Section 4. However, because the LM-8000 ISTIG is not a fully developed concept, limited power island design and modeling support was available from GE for the current study. Prior studies of the LM-8000 ISTIG have included extensive performance modeling, particularly for natural gas firing, and some of these modeling results were provided to SCS by GE. Using this database, SCS developed a SYNTHA thermal model of the ISTIG power island which was used to extrapolate to study conditions and match interface parameters with the gasification island. Complete energy and material balances and performance predictions were developed for the ISTIG cases using this approach. This information was provided to GE for review and informal concurrence.

5.2 AIR-BLOWN ISTIG WITH HOT GAS CLEANUP

5.2.1 Process Discussion

The process designs for the ISTIG Cases 2 and 4 are discussed below. From a process standpoint, the two cases are essentially identical, with Case 2 consisting of a single gasification and power island module and Case 4 consisting of four such modules. The modularity of the ISTIG plants was considered an important feature and was thus maintained for both the gasification and power island areas.

5.2.1.1 Overall Arrangement

The overall process arrangement for Case 2 is illustrated in the block flow diagram in Figure 5-1. The accompanying material balance is provided in Table 5-2. Figure 5-2 and Table 5-3 are the corresponding flow diagram and material balance for Case 4. As will be noted, the major differences between the ISTIG cases and the IGCC cases described in Section 4 occur in the power island area. The ISTIG power island consists of an aeroderivative gas turbine with intercooler, an HRSG (one per gas turbine), and associated auxiliaries. The net steam generated in the HRSG is fed directly to the combustion turbine, obviating the need for a steam turbine. The following sections discuss the ISTIG plant designs in greater detail by plant system.

5.2.1.2 Coal Receiving, Handling, and Preparation

Process flow diagrams for the coal receiving, handling, and preparation areas of Cases 2 and 4 are shown in Figures 5-3 and 5-4. The layout of the system for Case 2 is the same as for Case 1 except for differences related to the size of the plant. Differences in the capacities of certain equipment items are identified in the process flow diagram. With regard to equipment trains, Case 2 has one operating pulverizer with one backup, and the coal/limestone mixture is fed to the gasifier structure using one operating pneumatic system with one backup. The entire system for Case 4 is virtually identical to that for Case 1, since these cases are similar in size and gasification island design.

5.2.1.3 Limestone Receiving, Handling, and Preparation

Limestone systems for Cases 2 and 4 are included in the flowsheets shown in Figures 5-3 and 5-4. As will be noted, the layout of the system for Case 2 is the same as that for Case 1 except that only one limestone pulverizer is used in Case 2. Also, some equipment capacities are different because of differences in coal feed rate. As in the case of coal preparation, the limestone preparation system for Case 4 is identical to that for Case 1.

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Figure 5-1 Case 2 Block Flow Diagram

Table 5-2 (Sheet 1 of 4) Case 2 Process Stream Flows
Note: Major Streams not shown are proprietary to M.W. Kellogg

STREAM NUMBER	1	2	4	4A	10	14
DESCRIPTION	PREPARED COAL FEEDSTOCK	LIMESTONE FEEDSTOCK	TOTAL AIR FROM POWER ISLAND	AIR TO CT COMPRESSOR	STEAM TO GASIFIER	EXIT GAS COOLER STEAM TO POWER ISLAND
	WT%	WT%	VOL%	VOL%	VOL%	WT%
GASES, LB/HR						
CO	28,010		0.00	0.00		
H ₂	2,018		0.00	0.00		
CO ₂	44,010		0.00	0.00		
CH ₄	16,043		0.00	0.00		
N ₂	28,013		78.73	78.08		
Ar	38,948		0.00	0.00		
O ₂	31,989		0.00	0.00		
NH ₃	17,031		0.00	0.00		
H ₂ S	34,080		0.00	0.00		
CO ₂	60,075		0.00	0.00		
SO ₂	84,083		0.00	0.00		
H ₂ O	18,015		0.00	0.00		
HCl	38,481		0.00	0.00		
TOTAL GASES			100.00	100.00	100.00	100.00
GAS FLOW, LB MOL/HR			248,873	889,277	2,417	47,848
MOLECULAR WEIGHT			8.587	31.032	134	2,844
GAS VOLUME, ACFM			28.98	28	18.02	18.02
GAS VOLUME, SCFM			3,851	297,948	37	548
LIQUIDS, LB/HR			54,297	186,303	848	16,725
WATER	18,015					
SOLIDS, LB/HR						
COAL (ASH-FREE)	88.86	88,625				
LIMESTONE	10.34	7,681				
ASH	100.00	74,304				
TOTAL SOLIDS			248,873	889,277	2,417	47,848
TOTAL FLOW, LB/HR						
TEMPERATURE, F	200	80	745	80	840	825
PRESSURE, PSIA	14.0	14.0	480.0	14.7	850.0	850.0

Table 5-2 (Sheet 2 of 4) Case 2 Process Stream Flows

STREAM NUMBER	16	24	30	32*	44	47
DESCRIPTION	TOTAL HT. BPW FROM POWER ISLAND	SOLIDS TO DISPOSAL	STEAM TO ZnF ₂ REACTORS	PRODUCT GAS TO GAS TURBINE	FINES LOCK-HOPPER VENT GAS	SULFATOR STEAM TO POWER ISLAND
	VOL%	WT%	VOL%	VOL%	VOL%	VOL%
GASES, LB/HR						
CO	23,010			7.78	40,358	
H ₂	2,916			20.46	7,641	29
CO ₂	44,010			14.46	118,120	5
CH ₄	18,043			0.93	2,760	84
N ₂	23,913			36.46	189,385	2
Ar	39,948			0.44	3,290	134
O ₂	31,968			0.00	0	2
NH ₃	17,031			0.02	48	0
H ₂ S	34,060			0.00	13	0
CO _S	60,975			0.00	0	0
SO ₂	64,063			0.00	0	0
H ₂ O	18,915		100.00	18.36	64,063	31,419
HCl	38,461			0.00	1	100.00
TOTAL GASES			100.00	100.00	429,311	100.00
GAS FLOW, LB MOUN/HR						
MOLECULAR WEIGHT						
GAS VOLUME, ACFM						
GAS VOLUME, SCFM						
LIQUIDS, LB/HR						
WATER	18,915					
SOLIDS, LB/HR						
CARBON		1.1				
LIMESTONE		67.2				
ASH		31.8				
TOTAL SOLIDS		100.0				
TOTAL FLOW, LB/HR	61,772	24,115	10,918	429,311	303	31,419
TEMPERATURE, F	525	150	840	1,000	579	525
PRESSURE, PSIA	850.0	14.0	850.0	850.0	14.0	850.0

* Final fuel gas composition based on 97 percent removal of HCl across chloride guard and 95 percent removal of H₂S across zinc ferrite polishing.

Table 5-2 (Sheet 3 of 4) Case 2 Process Stream Flows

STREAM NUMBER	48	51	52	53	54	55	56
DESCRIPTION	SULFATOR FLUE GAS	TOTAL BLOWDOWN TO POWER ISLAND	TOTAL DEAFER BFW FROM POWER ISLAND	EXHAUST GAS TO HRSG	INJECTION STEAM	FLUE GAS TO MAIN STACK	BFW TO HRSG
	VOL%	VOL%	VOL%	VOL%	VOL%	VOL%	VOL%
GASES, LB/HR							
CO	28.010	0	0	0	0	0	0
H ₂	2.016	0	0	0	0	0	0
CO ₂	44.010	8.83	5,189	0	0	0	0
CH ₄	18.043	0	0	0	0	0	0
N ₂	28.013	53.60	20,885	0	0	0	0
Ar	39.948	1.02	540	0	0	0	0
O ₂	31.999	3.73	1,577	0	0	0	0
NH ₃	17.031	0.00	0	0	0	0	0
H ₂ S	34.080	0.30	0	0	0	0	0
CO ₈	60.075	0.00	0	0	0	0	0
SO ₂	64.063	0.02	19	0	0	0	0
H ₂ O	18.015	2.80	688	0	0	0	0
HCL	36.461	0.00	0	0	0	0	0
TOTAL GASES	100.00	28.878	0	100.00	100.00	100.00	100.00
GAS FLOW, LB MOL/HR	1,321	29.44	0	5,549	18.02	1,539	0
MOLECULAR WEIGHT	20.479	8.351	0	25.081	0	0	0
GAS VOLUME, ACFM	0	0	0	0	0	0	0
GAS VOLUME, SCFM	0	0	0	0	0	0	0
LIQUIDS, LB/HR	0	0	0	0	0	0	0
WATER	18.015	314	108,680	1,425,584	100,000	1,425,402	204,400
SOLIDS, LB/HR	0	0	0	0	0	0	0
COAL	1	0	0	0	0	0	0
LIMESTONE	32	67.32	0	0	0	0	0
ASH	15	31.70	0	0	0	0	0
TOTAL SOLIDS	48	100.00	0	0	0	0	0
TOTAL FLOW, LB/HR	38,824	314	108,680	1,425,584	100,000	1,425,402	204,400
TEMPERATURE, F	1,400	525	240	900	850	803	241
PRESSURE, PSIA	14.0	850.0	900.0	18.0	850.0	18.0	900.0

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 5-2 (Sheet 4 of 4) Case 2 Process Stream Flows

STREAM NUMBER	67	68	69	70	71	72
DESCRIPTION	BFW MAKEUP	BOILER BLOWDOWN	COOLING TOWER MAKEUP	COOLING TOWER BLOWDOWN EVAP., DRIFT	DILUTION AIR	FLUE GAS TO LOCAL STACK
	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %
MOLECULAR WEIGHT						
GASES, LB/HR						
CO	28.010					
H ₂	2.016					
CO ₂	44.010					
CH ₄	16.043					
N ₂	28.013					
Ar	39.948					
O ₂	31.999					
NH ₃	17.031					
H ₂ S	34.080					
CO _S	60.075					
SO ₂	64.063					
H ₂ O	18.016					
HCl	36.461					
TOTAL GASES						
GAS FLOW, LB Molar/HR						
MOLECULAR WEIGHT						
GAS VOLUME, ACFM						
GAS VOLUME, SCFM						
LIQUIDS, LB/HR	100.00	100.00	100.00	100.00	100.00	100.00
WATER	204,400	1,000	18,000	18,000	48,223	44,941
SOLIDS, LB/HR						
CARBON						
LIMESTONE						
ASH						
TOTAL SOLIDS						
TOTAL FLOW, LB/HR	204,400	1,000	18,000	18,000	48,223	44,941
TEMPERATURE, F	75	523	80	80	80	80
PRESSURE, PSIA	10.0	850.0	14.7	14.7	14.7	14.7

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 5-2 Case 4 Block Flow Diagram

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 5-3 (Sheet 1 of 4) Case 4 Process Stream Flows
Note: Major Streams not shown are proprietary to M.W. Kellogg

STREAM NUMBER	1	2	4	4A	10	14
DESCRIPTION	PREPARED COAL FEEDSTOCK	LIMESTONE FEEDSTOCK	TOTAL AIR FROM POWER ISLAND	AIR TO GT COMPRESSOR	STEAM TO GASIFIER	EXIT GAS COOLER STEAM TO POWER ISLAND
	WT%	WT%	VOL%	VOL%	VOL%	WT%
GASES, LB/HR						
CO	28,010		0.00	0.00		
H ₂	2,016		0.00	0.00		
CO ₂	44,010		0.00	0.00		
CH ₄	18,043		0.00	0.00		
N ₂	28,013		78.00	78.00		
Ar	38,948		0.98	0.98		
O ₂	31,988		230,228	20.85		
NH ₃	17,031		0.00	0.00		
H ₂ S	34,080		0.00	0.00		
CO ₂	60,075		0.00	0.00		
SO ₂	94,063		0.00	0.00		
H ₂ O	18,915		0.00	0.00		
HCl	38,461		0.00	0.00		
TOTAL GASES			100.00	100.00	100.00	100.00
GAS FLOW, LB MOL/HR			984,880	3,597,108	9,068	180,583
MOLECULAR WEIGHT			34,347	124,210	838	10,576
GAS VOLUME, ACFM			28,98	28,98	18.02	18.02
GAS VOLUME, SCFM			15,406	830,593	147	2,182
LIQUIDS, LB/HR			217,188	785,211	3,383	88,900
WATER						
SOLIDS, LB/HR						
COAL (ASH-FREE)	88.88					
LIMESTONE	10.34	100.00				
ASH						
TOTAL SOLIDS	100.00	100.00				
TOTAL FLOW, LB/HR	287,218	82,524	984,880	3,597,108	9,068	180,583
TEMPERATURE, F	200	90	745	90	840	825
PRESSURE, PSIA	14.0	14.0	480.0	14.7	850.0	850.0

Table 5-3 (Sheet 2 of 4) Case 4 Process Stream Flows

STREAM NUMBER	16	24	20	32*	44	47
DESCRIPTION	TOTAL HIGH TEMP. BFW FROM POWER ISLAND	SOLIDS TO DISPOSAL	STEAM TO ZnFs REACTORS	PRODUCT GAS TO GAS TURBINE	FINES LOCKHOPPER VENT GAS	SULFATOR STEAM TO POWER ISLAND
	VOL%	WT%	VOL%	VOL%	VOL%	VOL%
GASES, LB/HR						
CO	28,010			7.78	161,423	
H ₂	2,016			20.46	30,863	
CO ₂	44,010			14.49	472,479	
CH ₄	18,043			0.83	11,975	
N ₂	28,013			36.89	757,540	
Ar	39,948			0.44	13,120	
O ₂	31,909			0.00	0	
NH ₃	17,031			0.02	191	
H ₂ S	34,060			0.00	51	
COS	60,076			0.00	0	
SO ₂	94,063			0.00	0	
H ₂ O	18,015		100.00	19.39	258,763	100.00
HCl	36,461			0.00	5	125,677
TOTAL GASES			100.00	100.01	1,705,230	100.00
GAS FLOW, LB/MO/HR						
MOLECULAR WEIGHT						
GAS VOLUME, ACFM						
GAS VOLUME, SCFM						
LIQUIDS, LB/HR						
WATER	247,068					
SOLIDS, LB/HR						
CARBON		1.1		2,423	74,066	83
LIMESTONE		67.2		18.02	23,01	18.02
ASH		31.8		662	29,736	1,444
TOTAL SOLIDS		100.0		15,327	498,822	44,117
TOTAL FLOW, LB/HR	247,068	98,461	43,863		1,705,230	125,677
TEMPERATURE, F	625	150	840		1,000	625
PRESSURE, PSIA	850.0	14.0	850.0		850.0	850.0

*Final fuel gas composition based on 98 percent removal of HCl across chloride guard and 95 percent removal of H₂S across zinc ferrite polishing.

Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems

Table 5-3 (Sheet 3 of 4) Case 4 Process Stream Flows

STREAM NUMBER	48	51	52	53	54	55	56
DESCRIPTION	SULFATOR FLUE GAS	TOTAL BLOWDOWN TO POWER ISLAND	TOTAL DEAFER, BFW FROM POWER ISLAND	EXHAUST GAS TO HPSSG	INJECTION STEAM	FLUE GAS TO MAIN STACK	BFW TO HPSSG
	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %	VOL %
GASES, LB/HR							
CO	28,010						
H ₂	0.00						
CO ₂	2,016						
CH ₄	44,910						
N ₂	9.93						
Ar	20,755						
O ₂	18,043						
NH ₃	0.00						
H ₂ S	83.50						
CO ₂	123,541						
SO ₂	2,159						
H ₂ O	2,065						
HCl	3,309						
TOTAL GASES	17,031						
GAS FLOW, LB MOL/HR	0.00						
MOLECULAR WEIGHT	0.00						
GAS VOLUME, ACFM	0.00						
GAS VOLUME, SCFM	0.00						
LIQUIDS, LB/HR	0.00						
WATER	0.00						
SOLIDS, LB/HR	0.00						
CARBON	0.00						
LIMESTONE	0.00						
ASH	0.00						
TOTAL SOLIDS	0.00						
TOTAL FLOW, LB/HR	0.00						
TEMPERATURE, F	0.00						
PRESSURE, PSIA	0.00						

Table 5-3 (Sheet 4 of 4) Case 4 Process Stream Flows

STREAM NUMBER		57		58		59		60		61	
DESCRIPTION	MOLECULAR WEIGHT	BFW MAKEUP	BOILER BLOWDOWN	COOLING TOWER MAKEUP	COOLING TOWER BLOWDOWN EVAP., DRIFT	DILUTION AIR	FLUE GAS TO LOCAL STACK				
		VOL %	VOL %	VOL %	VOL %	VOL %	VOL %				
GASES, LB/HR											
CO	28.010										
H2	2.016										
CO2	44.010										
CH4	16.043										
N2	28.012										
Ar	39.948										
O2	31.998										
NH3	17.031										
H2S	34.080										
CO8	60.076										
SO2	64.063										
H2O	18.016										
HCl	36.461										
TOTAL GASES											
GAS FLOW, LB MOL/HR											
MOLECULAR WEIGHT											
GAS VOLUME, ACFM											
GAS VOLUME, SCFM											
LIQUIDS, LB/HR											
WATER	18.016	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
TOTAL GASES											
SOLIDS, LB/HR											
COAL											
LIMESTONE											
ASH											
TOTAL SOLIDS											
TOTAL FLOW, LB/HR											
TOTAL FLOW, LB/HR											
TEMPERATURE, F											
PRESSURE, PSIA

Figure 5-3 Case 2 Coal and Limestone Preparation Process Flow Diagram

Figure 5-4 Case 4 Coal and Limestone Preparation Process Flow Diagram

5.2.1.4 Gasification Island

The Case 2 and Case 4 gasifier island, depicted schematically in the process flow diagrams of Figures 5-5 through 5-8, is similar in most of the process areas to those previously described for Case 1. The areas of difference are discussed below.

The Case 2 design consists of a single gasifier train that is matched to a single ISTIG power island. A single set of components in a series flow path make up the gasifier island train. Since the ISTIG power island does not include a steam turbine, the product gas cooler is eliminated from the gasifier island. A blend of boiler feed water and steam is used to quench the gasifier product gas. The quench is controlled to reduce the gas temperature and to humidify the product gas as required for zinc ferrite chemistry. Case 2 includes one sulfator train. Case 4 is identical to Case 2 except for two variations. The major difference is that the gasifier island is modularly expanded from one to four gasifier trains to fuel the four ISTIGs of the power island. The second difference is that the number of sulfators is increased from one to two trains.

5.2.1.5 Power Island

Figure 5-9 shows the configuration of the power island areas for Cases 2 and 4. The figure includes heat and material balance information and performance data. Following is a brief discussion of the two major power island components, the LM-8000 gas turbine and the HRSG.

Gas Turbine System

Figure 5-10 shows the internal configuration of the ISTIG gas turbine. For the most part, the core engine consisting of the high-pressure compressor, combustor, and high-pressure turbine for the ISTIG unit is the same as that for the CF6-80C2 aircraft engine, while the low-pressure compressor and turbine and the power turbine will be of unique designs due to the differences between the power generation cycle and the aircraft engine application.

Case 2 of this study is based on the use of one ISTIG unit with the associated gasifier and all accessories required to make a standalone power generation facility. This provides a nominally rated 100-MW unit. Case 4 is designed to provide approximately 400 MW, so that four separate ISTIG units are used in the power island design.

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 5-5 Cases 2 and 4 Gasification Process Flow Diagram

3801
RECYCLE GAS
RECEIVER

3610
RECYCLE GAS
COOLER

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 5-6
Heat Recovery
and Compression
Process Flow Diagram

Figure 5-6
Heat Recovery
and Compression
Process Flow Diagram

Figure 5-6 Cases 2 and 4 Heat Recovery and Compression Process Flow Diagram

<u>130F</u>	<u>4311</u>
ENTER	ENTER
RESEARCHER	SUBJECT
UNKNOWN	UNKNOWN

$$A \cap B = \emptyset \quad \text{if } A \in \mathcal{A}_1 \text{ and } B \in \mathcal{A}_2 \text{ or } A \in \mathcal{A}_2 \text{ and } B \in \mathcal{A}_1$$

510D, DA
ZINC FERRITE
REACTORS

220B
START-UP HEATER

251J
BOOSTER
AIR COMPRESSOR

251C
AIR
RECUPERATOR

252CA
BFW
PREHEATER

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 5-8 Cases 2 and 4 External Desulfurization Process Flow Diagram

ISTIG SCHEMATIC BASED ON
AIR-BLOWN FIXED BED GASIFIER

POWER	147,700 kW
EFF.	42.1%
HEAT INPUT	2,354,000 Btu/h
HEAT TEMP	900 °F
AIR FLOW	45,000 lb/h
EXH FLOW	57,000 lb/h
LP STEAM	4,000 lb/h
HP STEAM	2,000 lb/h

Case 2

Case 2: 1 GE LM-8000 ISTIG unit

Case 4: 4 GE LM-8000 ISTIG units

Flow shown represent performance of one unit

Case 4

Figure 5-9 Cases 2 and 4 Power Island Process Flow Diagram

Figure 5-10 Internal Configuration of the LM-8000 Gas Turbine for ISTIG Cases 2 and 4

HRSG System

The heat recovery steam generator (HRSG) takes the exhaust flow from the LM-8000 ISTIG and uses the heat to produce steam for injection into the ISTIG unit and for process use in the gasifier island. For modularity of design, one HRSG is used per gas turbine. The thermal configuration of the HRSG is heavily integrated with the gasifier island in order to achieve maximum steam production for the overall plant. Figure 5-9 shows how the HRSG is integrated into the power island.

To meet the steam injection requirements of the ISTIG unit, a two-pressure-level HRSG configuration is used with an HP section producing both gasifier blast steam and HP injection steam and an IP section producing IP injection steam for the ISTIG unit.

5.2.1.6 Balance of Plant and General Facilities

The design approach for the balance of the plant and general facilities for Cases 2 and 4 was the same as for the IGCC cases. Of course, the lack of steam turbine systems for these cases reduced the scope of the balance-of-plant systems and eliminated or curtailed certain general facility items.

5.2.2 Plant Layout and Equipment

The plant layouts for the ISTIG designs are located in the same general area of Plant Wansley as the combined-cycle Cases 1, 3, 5, and 6. Equipment arrangements and equipment design parameters were developed based on the results of the thermal analysis of the ISTIG cycle and based on specific structural and material requirements for the various pieces of equipment. The equipment arrangements and designs were developed based on the modular concept for the ISTIG unit. Each ISTIG unit, representing a nominal 100 MW, stands as a separate unit which can be installed incrementally as demand for system power increases.

Figure 5-11 shows the general arrangement for the ISTIG gas turbine and support equipment. A conceptual plot plan and elevation and isometric views of the gasification train for Cases 2 and 4 are given in Figures 5-12 and 5-13. The major equipment required includes the ISTIG gas turbine, inlet air filters, compressor intercooler, cooling system for the intercooler, HRSG, water treatment system, and storage tanks.

Figure 5-11 Cases 2 and 4 Overall Site Plan

[illegible]

5-31

Figure 5-13 (Sheet 1 of 2) Gasification Train Layout, Elevation View
Cases 2 and 4

Figure 5-13 (Sheet 2 of 2) Gasification Train Layout, Isometric View,
Cases 2 and 4

5.3 ISTIG PLANT OPERATIONS AND PERFORMANCE (CASES 2 AND 4)

5.3.1 Operational Considerations

5.3.1.1 Spare Equipment and Material

As discussed previously for the IGCC cases (Section 4.5.1.1), equipment and material sparing for the gasification island was given special attention in the current study in order to ensure that plant operability, reliability, and capital costs are fully addressed. There are two major categories of spare equipment included in the ISTIG design cases: inline spare equipment and warehouse spare parts. Inline spares are those pieces of equipment that are permanently connected to the plant and are used only when a failure occurs in the piece of equipment being spared. The inline spare parts requirements for Cases 2 and 4 are given in Table 5-4. Warehouse spare parts as referenced here are for major components and are in addition to those items included in normal spare parts cost factors. These spares are listed in Table 5-5.

5.3.1.2 Catalysts and Chemical Requirements

Initial catalysts and chemicals are included in the capital cost estimate as a separate line item. Catalyst and chemical requirements are substantial for the gasification island, and thus the quantities have been estimated based on material balance data. Table 5-6 shows these quantities for the ISTIG cases. Annual usages of these items (shown at 85-percent capacity factor) are also provided for determination of operating cost.

5.3.1.3 Spare Capacity

The gasifier island equipment components are designed with excess capacity. Should it be necessary to idle or shut down one gasifier train, the capacity of the remaining trains can be increased to make up for a portion of the inoperative gasifier train output. A gasifier train consists of all equipment from the pressurization of feedstocks to the production of clean gas at the exit of the external desulfurizer. The complete train will be isolated and taken out of service if a component failure requires depressurization for repair. This excess capacity and the capacity less one operating train are shown for ISTIG Cases 2 and 4 in Table 5-7 along with the calculated equivalent availability, as determined from the reliability analysis, described in the subsequent section.

Table 5-4
Inline Spare Equipment

<u>Items</u>	Numbers of Units	
	<u>Case</u> <u>2</u>	<u>Case</u> <u>4</u>
Coal Feeder (310-V)		
- Operational Units	1	4
- Spare Units	1	2
Ash Feeder (330-V)		
- Operational Units	1	4
- Spare Units	1	2
Recycle Gas Compressor (381-J)		
- Operational Units	1	4
- Spare Units	1	2
Sulfator Vent Fan		
- Operational Units	1	2
- Spare Units	1	2

Table 5-5
Major Component Warehouse Spares

<u>Items</u>	Numbers of Units	
	<u>Case</u> <u>2</u>	<u>Case</u> <u>4</u>
Ceramic Candle Filters		
- No. Operational Elements	425	1,700
- No. Spare Elements	100	300
Hot Gas Valves (18 inch)		
- No. Operational Valves	8	32
- No. Spare Valves	1	2
Hot Solids Valves		
- No. Operational Valves	5	20
- No. Spare Valves	2	4
Sulfator Conveyor		
- No. of Operational Units	1	2
- Spare Units 602 V	1	1
- Spare Units 603 V	1	1

Table 5-6

Catalysts and Chemicals

	<u>Initial Charge</u>	<u>Inventory Capital (Spare)</u>	<u>Annual Usage (85% C.F.)</u>
<u>Case 2</u>			
Nahcolite (tons)	88	--	355
Zinc Ferrite (tons)	284	142	62
<u>Case 4</u>			
Nahcolite (tons)	350	--	1,420
Zinc Ferrite (tons)	1,136	142	249

Table 5-7

Spare Capacities

	<u>Case 2</u>	<u>Case 4</u>
No. of Gasifier Trains	1	4
Excess Capacity, %	19	19
Capacity, Less One Train, %	0	89
Equivalent Availability of Gasifier Island, %	93	95

5.3.1.4 Equipment and Plant Reliability

RAM analyses were conducted for the two ISTIG cases to determine plant availability and adjust the plant design to achieve the availability target. An equivalent availability of 85 percent was targeted for the large gasification plants, including Case 4, while a minimum availability of 65 percent was specified for small plants such as Case 2.

The approach to the RAM analysis was described in detail in Section 4.5.1.4 of the report and, therefore, will not be reiterated here. The availability block diagrams (ABD) for Cases 2 and 4 are shown in Figures 5-14 and 5-15. Equipment reliability data in the form of mean time between failure (MTBF) and mean downtime (MDT) were developed for each component in the ABD, and these data were input to the UNIRAM code along with annual planned outage time and other data for the plants. Data sources and study assumptions were the same as those described in Section 4.5. The UNIRAM program inputs are presented in Appendix D.

The major results of the UNIRAM analyses are the following calculated (actual) equivalent availabilities, here compared to the target values:

	<u>Target EA (%)</u>	<u>Actual EA (%)</u>
Case 2	> 65	84.17
Case 4	85	86.10

The 100-MW ISTIG of Case 2 easily met its conservative availability goal, which shows the high reliability which could potentially be achieved by this type of plant, even at the scale of the smallest module. The equivalent availability of the 400-MW ISTIG of Case 4 slightly exceeded the goal of 85 percent. Because of the highly modular nature of the ISTIG concept, little availability improvement is seen when multiple modules are installed.

Additional reliability data and performance indexes are shown for the ISTIG plants in Table 5-8. Definitions of the reliability performance nomenclature are discussed in Section 4.5.

The two component types causing the most forced lost generation for the ISTIG plants were generally the combustion turbine generators (CTG) and the heat recovery steam generators (HRSG). Together, these components accounted for 42 percent to 50 percent of the equivalent forced outage hours (EFOH), depending on the case as shown in Table 5-9. The gasification island accounted

Figure 5-14 (Sheet 1 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 2

Figure 5-14 (Sheet 2 of 2) Integrated Gasification Combined Cycle Power Plant
Availability Block Diagram - Case 2

Figure 5-15 (Sheet 1 of 2) Integrated Gasification Combined Cycle Power Plant Availability Block Diagram - Case 4

Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems

Figure 5-15 (Sheet 2 of 2) Integrated Gasification Combined Cycle Power Plant Availability Block Diagram - Case 4

Table 5-8

**Plant Wansley IGCC Projected Reliability
Performance Indexes**

	<u>EA(%)</u>	<u>A(%)</u>	<u>EFOR(%)</u>	<u>FOR(%)</u>	<u>Days Per Year At Maximum Output</u>
Case 2	84.17	84.18	11.67	11.67	307
Case 4	86.10	94.67	9.65	0.66	229

Where:

EA = Equivalent availability

A = Availability

EFOR = Equivalent forced outage rate

FOR = Forced outage rate

Table 5-9

Critical Components for ISTIG Plants

	Percentage of total EFOH*	
<u>Component</u>	<u>Case 2</u>	<u>Case 4</u>
Combustion Turbine/Generator	27	32
HRSG	15	18
Gasification Island/ Sulfator System	55/23	46/27

* Equivalent forced outage hours

for 46 percent to 55 percent of the EFOH. The sulfator system (part of the gasification island) accounted for 42 percent to 59 percent of the forced outage problems for the gasification island. A complete list of all components ranked by criticality is included in Appendix C for the ISTIG cases.

5.3.1.5 Utility Requirements

Table 5-10 provides the utility requirements of the gasifier island. Electrical power requirements are tabulated by plant area with the booster air compressor requiring the most power. Cooling water requirements are also tabulated by plant area. As with the IGCC cases, the utility requirement decreases with lower ambient plant operating temperature, due primarily to power demand for the booster air compression area.

5.3.2 Plant Performance

5.3.2.1 Heat and Material Balances

As was described for the IGCC cases in Section 4.5.2, heat and material balances for the ISTIG plants were conducted separately for the gasification island and the power island portions of the plant. The overall plant material balances for Cases 2 and 4 are summarized in Table 5-11.

5.3.2.2 Gasifier Island Performance and Interfaces

The gasifier island performance and interface parameters are summarized in Table 5-12, which shows the quantities of feeds to the process and the gasifier island products produced. The performance of the two ISTIG cases is the same in terms of cold gas efficiency, carbon conversion, and sulfur removal due to the single-train configuration of integrating a single gasifier with a single combustion turbine. Interface parameters between the gasification and power island are summarized in Table 5-13.

5.3.2.3 Power Island Performance

Three major areas were focused on for development of the thermal cycle: (1) combustion turbine configuration, (2) HRSG configuration, and (3) interface points to the gasifier island. A thermal model of the ISTIG system was developed using the SYNTHA thermal modeling computer program. Due to the prototype

Table 5-10

Utility Summary for ISTIG Cases

Electrical Power Requirements, kW:

<u>Area Number</u>	<u>Description</u>	<u>Case 2</u>	<u>Case 4</u>
250	Booster air compression	2,055	8,220
300	KRW gasification	243	972
380	Recycle gas compression	192	769
600	Sulfation	243	970
	Miscellaneous uses	80	320
		<hr/>	<hr/>
	TOTAL	2,813	11,251

Cooling
Water, GPM

250	Booster air compression	997	3,989
300	KRW Gasification	72	290
380	Recycle gas compression	6	24
400	Gas conditioning	45	178
600	Sulfation	492	1,968
		<hr/>	<hr/>
	TOTAL	1,612	6,449

Table 5-11
Overall Material Balances

	<u>Case 2</u>	<u>Case 4</u>
<u>Material In. lb/hr</u>		
Limestone (Dried)	20,631	82,524
Coal (As Received)	79,148	316,593
Combustion air to gas turbine	899,277	3,597,108
Cooling tower makeup	18,000	72,000
Air to sulfator	40,889	163,556
Air compressor feed to gasifier	248,673	994,690
Coal drying gas dilution air	46,223	184,891
BFW makeup	204,400	817,600
TOTAL MATERIAL IN	1,557,241	6,228,962
<u>Material Out. lb/hr</u>		
HRSG flue gas to main stack	1,425,584	5,702,336
Coal drying gas to local stack	89,277	357,109
Solids to disposal	24,115	96,461
Boiler blowdown	1,066	4,264
Cooling tower blowdown	18,000	72,000
TOTAL MATERIAL OUT	1,558,042	6,232,170
Closure	100.05%	100.05%

Table 5-12

**Gas Island Design Basis
Performance Comparison**

	<u>Case 2</u>	<u>Case 4</u>
<u>Feeds:</u>		
Coal feed (t/day)	892	3,567
Limestone feed (t/day)	248	990
Air supply (t/day)	2,984	11,936
Steam (MMlb/day)		
450 lb	----	----
850 lb	160	640
Solids waste (t/day)	289	1,158
Oxygen (t/day)	----	----
<u>Products:</u>		
Fuel gas (LHV, MMBtu/day)	15,096	60,384
Steam (MMlb/day)		
1600 lb sat.	----	----
850 lb sat.	949	3,795
Sulfuric acid (t/day)	----	----
<u>Performance:</u>		
Cold gas efficiency - LHV (%)	70.6	70.6
Carbon conversion (%)		
Gasifier only	97.1	97.1
Gas island	99.9	99.9
Sulfur removal (%)		
Gasifier only	86.5	86.5
Gas island	99.4	99.4

Table 5-13

MWK/SCS Interface Parameters

	<u>Case 2</u>	<u>Case 4</u>
<u>Gasifier/Power Island Interface</u>		
Coal (sized)		
Flow, lb/hr	74,304	297,218
Temp, °F	200	200
Pressure	amb	amb
Limestone (size)		
Flow, lb/hr	20,631	82,524
Process air		
Flow, lb/hr	248,673	994,690
Temp, °F	745	745
Pressure, psia	480	480
Product Gas		
LHV, Btu/SCF	89.5	89.5
Flow, MMbtu/hr (LHV)	629	2,516
BFW From power island		
Flow, lb/hr	108,690/61,772	434,762/247,088
Temp, °F	240/525	240/525
Pressure, psia	850	850
BFW to power island		
Flow, lb/hr	314	1,257
Temp, °f	525	525
Pressure, psia	850	850
Steam from power island		
Flow, lb/hr	13,333	53,329
Temp, °F	840	840
Pressure, psia	850	850
Steam to power island		
Flow, lb/hr	79,065	316,260
Temp, °F	525	525
Pressure, psia	850	850
Solids to disposal, lb/hr	24,115	96,461
Sulfator flue gas to power island		
Flow, lb/hr	38,925	155,699
Temp, °F	1,400	1,400
Vent gas to flare, scfm	83	332

nature of the ISTIG design, actual performance data were largely unavailable. The following design points were utilized in the development of the performance data:

- HRSG - dual pressure with 900-psi HP drum and 250-psi IP drum, integral deaerator and feedwater heating coil - overall 1.5-percent heat loss per section, 1-percent HP and IP drum blowdown
- Deaerator pressure - 25 psi to provide 240°F feedwater temperature
- Boiler feedpump discharge pressure - 1050 psi to allow for 13-percent pressure drop through HRSG

As in the thermal analysis of the combined-cycle cases, 280°F was set as the minimum exhaust gas temperature in order to minimize acidic deposition and avoid stack corrosion. The stack temperature ultimately used in the design (663°F) was well above this target because of heat recovery limitations. The HP evaporator design pinch point was set at 35°F.

ISTIG Performance

The ISTIG gas turbine performance was developed through thermal modeling of the ISTIG internal systems using the SYNTHA thermal modeling program and data provided by GE for the natural-gas-fired version of the ISTIG. The performance results are summarized in Table 5-14. The results were compared to those of the previous studies to assure that reasonable results were achieved. Due to the prototype nature of the LM-8000 ISTIG, the results should be considered as preliminary estimates of the ISTIG potential.

Figure 5-9 displays the resulting power island heat balances for Cases 2 and 4. Power output shown in this figure is gross power production without consideration of plant station service requirements.

Station Service Requirements

Based on the thermal cycles developed for Cases 2 and 4, major equipment was sized and power requirements determined to calculate the overall plant station service. Table 5-15 provides a detailed breakdown of the station service for the complete ISTIG plant including the power island, gasifier island, and coal and limestone handling areas.

Net Power Plant Performance

Table 5-16 shows the calculated net power output for Cases 2 and 4. The net power values are simply the gross power outputs of the power island less the

Table 5-14

Gas Turbine Performance Summary - Cases 2 and 4
90°F Design Point

	<u>Case 2</u>	<u>Case 4</u>
Ambient Temperature	90°F	90°F
Gross Power Output (kW)	104,744	418,976
Fuel Consumption (MMBtu/hr)	628.9	2,515.7
Compressor Bleed Air Temperature (°F)	745	745
Compressor Bleed Air Pressure (psi)	480	480
Exhaust Flow (1,000 lb/hr)	1,425	5,700
Exhaust Temp. (°F)	909	909

Table 5-15

Plant Station Service Requirements - Cases 2 and 4

	<u>Case 2</u>	<u>Case 4</u>
POWER ISLAND		
HRSG Feedpumps	220	880
Condensate Pumps	20	80
Intercooler Fans	384	1,561
Makeup Pumps	19	76
Gas Turbine Auxiliaries	750	3,000
Water Treatment	160	640
Miscellaneous	18	48
	<hr/>	<hr/>
Total Power Island	1,571	6,285
GASIFIER ISLAND		
Booster Air Compressor	2,055	8,220
KRW Gasification	243	972
Recycle Gas Compression	192	769
Sulfation	243	970
Miscellaneous	80	320
	<hr/>	<hr/>
Total Gasifier Island	2,813	11,251
COAL & LIMESTONE HANDLING	791	2,372
	<hr/>	<hr/>
Total	5,175	19,908

Table 5-16

Overall Performance for ISTIG Design
Cases 2 and 4 (90°F)

	<u>Case 2</u>	<u>Case 4</u>
Gross Power Output (kW)	104,744	418,976
Power Island Station Service (kW)	1,571	6,285
Coal/Limestone Handling Station Service (kW)	791	2,372
Gasifier Island Station Service (kW)	2,813	11,251
Net Plant Power Output (kW)	99,569	399,068
Fuel Consumption (HHV MMBtu/hr)	926.5	3,706
Net Heat Rate (Btu/kWh - HHV)	9,305	9,287

station service requirements of the power island, gasifier island, and coal/limestone handling facilities. The fuel consumption and the net plant heat rates shown in Table 5-16 are based on the design flow rates of the coal to the gasifier island on a higher heating value (HHV) basis.

5.3.2.4 Environmental Performance

The environmental performance for Cases 2 and 4 is discussed in this section.

The air emission rates for major pollutants are listed in the Table 5-17 and are compared to the NSPS standards outlined in Section 3.2.3.

As previously described, sulfur removal occurs with in-bed desulfurization with limestone in the gasifier for bulk sulfur removal followed by a zinc ferrite polishing step. The sulfur removal for Cases 2 and 4 is approximately 99 percent. Table 5-18 lists the distribution of sulfur in the product streams for Cases 2 and 4.

The solid waste from the ISTIG plants is similar to the IGCC cases which include spent limestone and ash from the sulfator, nahcolite from the chloride guard bed and spent zinc ferrite sorbent. The quantities of solid wastes generated are shown in Table 5-19.

Table 5-17

Air Emission Rates for Major Pollutants

	<u>Case 2</u>	<u>Case 4</u>	<u>NSPS</u>
<u>NO_x</u>			
lb/MMBtu	0.163	0.163	---
ppmv at 15% O ₂ (dry)	76.9	76.6	75
<u>SO₂</u>			
lb/MMBtu			
HRSG	0.026	0.026	----
Coal Prep Stack	0.021	0.021	----
Total	0.047	0.047	1.2

Particulates

Particulate removal is by cyclones and ceramic filters. The removal efficiency is effectively 100 percent. Modeling was conducted at the maximum concentration allowed in the combustion turbine, although the actual concentration is less.

Table 5-18

Overall Sulfur Balance for Cases 2 and 4

	<u>Case 2</u>	<u>Case 4</u>
<u>Sulfur In</u>		
Sulfur in feed coal, t/d	23.9	95.7
<u>Sulfur Out</u>		
Solids to landfill, t/d	23.6	94.6
Sulfator flue gas, t/d	0.1	0.5
HRSG flue gas, t/d	0.2	0.6
Total Out, t/d	23.9	95.7

Table 5-19

Solid Waste Quantities for ISTIG Plants

<u>Case</u>	<u>Tons/day</u>
2	289
4	1,158

6.0 PLANT DESCRIPTIONS — MILD GASIFICATION BASED ON IGCC WITH TRANSPORT REACTORS — CASE 7

6.1 DESIGN BASIS

This case was defined as a 400-MW IGCC power plant with a transport pyrolyzer for mild gasification to produce char feed for the transport gasifier and a liquid product for peaking turbine fuel. The 400-MW power plant utilizes a transport gasifier and a transport combustor in a hybrid mode to produce fuel gas and hot flue gas for the topping gas turbine. In Case 6, this reactor system is directly fed with coal. In Case 7, the coal is fed to a pyrolyzer and the resulting char is fed to the gasifier/combustor system. The concept is to continuously produce and store coal liquids in sufficient quantity to satisfy peaking turbine fuel requirements. In this way, a power plant would not have to rely on the availability of natural gas or fuel oil for peaking turbine fuel. The cost of producing the liquid fuel by this method is expected to be low because it involves only incremental capital addition.

The yield and quality of coal liquids is dependent not only on the coal properties but also on the conditions of the pyrolysis. There has been a great deal of research in this area, but unfortunately most of the information is not available at the anticipated operating conditions of the transport pyrolyzer. The greatest quantity of coal liquids will be produced at relatively low temperature and low pressure and short contact time. Also, coal with a relatively small particle size (50 to 100 micrometer average) is preferred because volatile material appears to crack in the pores of large coal particles. The quality of the liquid produced changes dramatically as reactor conditions change. The initial liquids produced are typically highly phenolic and cresylic and tend to be unstable in storage. The presence of limestone may catalytically crack the phenolic and cresylic components of the volatile material to form lighter products. Thus, the transport reactor appears to be a natural selection for the pyrolysis step under these conditions.

6.2 PROCESS DESCRIPTION

A simplified schematic of the conceptual process is shown in Figure 6-1. Pulverized coal (50 micrometers average particle size) is pneumatically delivered to a lockhopper system where it is pressurized and fed to the transport pyrolyzer. The coal is introduced into a circulating stream of hot solids where the coal is heated and devolatilized. Heat is supplied to the circulating solids by introduction of another hot solids stream for the transport combustor (sulfator). The volatiles released are condensed and stabilized producing a coal liquid product for storage and a fuel gas stream.

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Figure 6-1 Schematic of the Conceptual Process

The use of recycled hot solids from the char combustor is the preferred mode of supplying the reaction heat to the pyrolyzer as opposed to the use of combustion in the reactor. There are several problems associated with the combustion route. The use of air or oxygen adds inert combustion products into the system, complicating the separation of the coal liquids and resulting in lower-Btu fuel gas being produced. In the power production area of the plant, air would be the natural oxidant to use, and the nitrogen introduced would further lower the calorific content of the fuel gas, making it difficult to use as a gas turbine fuel.

The devolatilized char and circulated solids (containing ash and spent limestone) flow to the transport gasifier. The char mixes into a circulating stream of hot solids, fresh limestone, and process air, partially gasifying the char and producing a fuel gas rich in hydrogen and carbon oxides. The limestone calcines and absorbs hydrogen sulfide from the fuel gas as it is produced. The resulting desulfurized fuel stream is filtered to remove particulates and sent to the gas turbine as its primary fuel.

The partially gasified char produced in the gasifier flows to the transport combustor (sulfator) where it mixes into a circulating stream of hot solids (composed of ash and partially spent limestone) and process air, completing the char gasification. The limestone absorbs the sulfur dioxide released in the char combustion, producing a desulfurized flue gas. Steam is produced from the excess heat of combustion. The flue gas is filtered and sent to the gas turbine for recovery of power. In the presence of excess air, the calcium sulfide from the gasifier is oxidized to calcium sulfate. Hot ash and spent limestone are circulated to the pyrolyzer to supply reaction heat, and the net product is cooled and sent to disposal.

In the power island, the fuel gas produced in the char gasifier flows to the gas turbine combustor chamber (GE MS7001F) and is combusted in the presence of flue gas from the char combustor and air from the gas turbine air compressor. About one-third of the air compressor flow is used as process air and is boosted to the pressure required by the gasifier and combustor. The net energy produced by the gas turbine is used to generate electric power. The power island also includes a heat recovery steam generator (HRSG) which recovers the heat in the gas turbine exhaust before it is discharged to the stack. Steam produced in the HRSG and the char combustor, is used in the turbogenerator to produce electric power.

6.3 TECHNICAL FEASIBILITY ASSESSMENT

An assessment of the technical feasibility of an IGCC system utilizing mild gasification was conducted prior to initiating detailed design engineering to determine (1) if the peaking capacity additions for Georgia Power Company matched the quantity of liquid produced from the mild gasification step, and (2) if the mild

gasification liquids could be utilized directly in a combustion turbine. A decision point at the conclusion of the technical feasibility assessment determined whether to recommend proceeding with detailed design engineering.

A rough material balance was made around the gasification island of the IGCC system – which included the mild gasification reactor, gasifier, and sulfator – in order to estimate the quantity of liquids produced in the mild gasification step. The yields and compositions for the products from the mild gasification step were taken from readily available literature sources. The estimated product yields were adapted from the bench-scale work conducted at the Energy and Environmental Research Center (EERC) in Grand Forks, North Dakota, under a separate DOE contract to develop a mild gasification processing scheme (Reference 6). The mild gasification product yields estimated by EERC for an Illinois No. 6 bituminous coal are shown in Table 6-1. No composition analyses for the mild gasification liquids and gases had been performed at the time the referenced EERC report was issued. Therefore, the typical properties for the liquids produced from the FMC COED process for an Illinois No. 6 coal were used, as listed in Table 6-2 (Reference 7). The composition of the gases produced during mild gasification (N_2 - and H_2S -free basis) were estimated based on the typical gas composition from the Garrett flash pyrolysis process developed by Occidental Petroleum Company, as shown in Table 6-3 (Reference 8). The material balance information from Case 6 was used for coal and product gas compositions, steam/carbon and oxygen/carbon feed ratios, sulfur capture, and the total heat content requirement for the fuel gas for a nominal 400-MW plant.

The results of the material balance were used to estimate the number of peaking combustion turbines required. Approximately 800 tons/day of raw pyrolysis liquids are produced for peaking power generation under the Case 7 processing scenario. Based on the broad assumption that these liquids yield a similar heat rate for a GE7E combustion turbine as No. 2 fuel oil, and assuming 330 days/yr of operation of the mild gasification plant, the estimated annual power generation for peaking is 365 MW. This would require five GE7E combustion turbines rated at 73 MW each. The expansion plan for The Southern Company estimates the addition of peaking capacity for Georgia in increments of 252 MW in the year 2005, 126 MW in the year 2006, and 252 MW in the year 2009. Therefore, it is reasonable to assume that the peaking power generated from pyrolysis liquids produced in an IGCC mild gasification plant could be utilized by Georgia Power Company.

GE was contacted to obtain an assessment as to whether the pyrolysis liquids produced from the mild gasification step were suitable for use directly as a peaking fuel in a combustion turbine. The typical properties reported in Table 6-2 from the COED process for an Illinois No. 6 bituminous coal were sent

Table 6-1

**Estimated Yields from EERC'S Bench-Scale Work for
Illinois No. 6 Bituminous Coal**

	<u>Yield, Wt%</u>
Char	66.0
Liquids	21.0
Water	6.0
Gas	7.0

Table 6-2

**Typical Properties of Liquids from COED Process for
Illinois No. 6 Bituminous Coal**

	<u>Analysis, Wt%</u>
Carbon	79.6
Hydrogen	7.1
Nitrogen	1.1
Sulfur	2.8
Oxygen	8.5
Ash	0.9
H/C molar ratio	1.07
Spec. Gravity	1.11
Higher Heating Value, Btu/lb	15,435

Table 6-3

**Typical Gas Composition from Garrett Flash Pyrolysis
(N₂- and H₂S-Free Basis)**

	<u>Mol%</u>
H ₂	26.8
CO	30.0
CO ₂	8.5
CH ₄	22.4
C ₂ H ₄	12.3
Higher Heating Value (25°C)	625 Btu/scf

to GE for review and comment. GE responded that, based on prior GE experience with similar fuels, the liquids cannot be burned properly in a combustion turbine as a standalone fuel due to three major problems:

1. At a hydrogen level of nominally 7 percent, the liquids will substantially smoke during combustion, causing an opacity problem with the flue gas.
2. Coking of the firing nozzles will also occur due to the low hydrogen content.
3. The emissivity will be higher, causing more radiant heat to be spread to the walls of the combustion turbine. This would require using a different combustion liner to compensate for the additional heat load.

GE recommended that the hydrogen content of the liquids could be upgraded to improve operability in the combustion turbine either by blending with a distillate in a homogenizing step prior to the turbine or by hydrotreating the liquids. In either case, the minimum net hydrogen level of the liquids to be fired in the turbine should be approximately 11 percent. GE was also concerned with the types of contaminants that may be present in the ash of the mild gasification liquids, as well as organic compounds that make up the liquids, which may produce carcinogens as a result of combustion in the turbine exhaust. This type of information, however, is not readily available in the literature.

Blending the mild gasification liquids with a No. 2 fuel oil that would be available at the Wansley site was not considered feasible due to the large quantities required to achieve a turbine feed with 11-percent hydrogen content. Hydrotreating of the liquids to upgrade the hydrogen content would require extensive process engineering to develop a workable process flow concept that was beyond the scope of this study. A joint decision between DOE and SCS was made to cease further work on Case 7 for this program.

6.4 RECOMMENDATIONS FOR FURTHER STUDY

Based on the results of the feasibility assessment, further study is required in order to develop a workable processing scenario for an IGCC plant that includes a mild gasification step. The following are recommendations for future investigations:

1. Investigation in greater depth of the limits to firing of mild gasification liquids in a combustion turbine, including actual combustion tests if warranted and feasible.
2. Development of an upgrading step for the mild gasification liquids to increase the hydrogen content so that they can be used as a peaking fuel in a combustion turbine.

3. **Better characterization of the mild gasification liquids from advanced processing schemes to determine the amount of upgrading and identify possible contaminants for the combustion turbine.**
4. **Identification of alternative methods of providing process heating for the mild gasification step to avoid using fuel gas that could be used in the combustion turbine.**
5. **Investigation of methods of handling and storing raw mild gasification liquids.**

7.0 CAPITAL AND OPERATING COST ESTIMATES

7.1 CAPITAL COST ESTIMATES

Sections 4 and 5 described plant designs for four IGCC power plants (Cases 1, 3, 5, and 6) and two gasification-based ISTIG power plants (Cases 2 and 4). Based on these plant designs, conceptual capital cost estimates were developed using the standard methods, data sources, and costing philosophies employed by SCS and MWK. Section 7.0 presents the capital cost results for these plant designs.

7.1.1 Estimating Assumptions and Procedures

Using the information reported in the previous sections, such as design criteria, flow diagrams, equipment lists and specifications, material and energy balances, and equipment layouts, SCS and MWK prepared process area plant costs from which total capital requirement estimates were built. MWK developed the basis for the gasification and sulfuric acid plant costs, while SCS was responsible for the power island, materials handling, solids handling and disposal, and the general facilities. SCS was then responsible for preparing the overall capital cost estimates for site-specific plants located at Georgia Power Company's Plant Wansley. All capital costs are expressed in mid-1990 dollars.

The total capital requirement is projected by estimating the following components:

Process plant cost	A
General facilities, engineering, contingencies	B
Total plant cost	$C = A + B$
AFUDC	D
Total plant investment	$E = C + D$
Royalties, initial inventory, startup, spare parts, working capital, and land (= owner's costs)	F
Total capital requirement	$G = E + F$

It should be noted that DOE supplied a cost-of-electricity model and format for calculation of costs developed in the study. The usual EPRI Technology Assessment Guide (TAG) format for presenting capital costs was adapted in minor ways to conform to the DOE cost model format, as will be pointed out subsequently.

The major cost items specified above are described in more detail in the following sections.

7.1.1.1 Process Plant Cost

The process plant cost is the compilation of the individual process section costs. Both SCS and MWK employed engineers and professional cost estimators to develop these costs using the following sources: a.) firm vendor quotes, b.) budgetary vendor quotes, c.) recent procurement of similar items, d.) historical, in-house experience and data bases. The components of the process plant cost are direct field material costs, direct field labor and subcontract costs, indirect field material and labor costs, and costs of freight and sales tax.

Direct field material costs are the costs of the permanent physical plant facilities and include the following:

- **Equipment** - includes all machinery, such as tanks, vessels, rotating machinery, boilers, and exchangers, used in the completed facility.
- **Material** - includes the cost of all physical materials, such as cement, aggregate, sand, steel, building materials, pipe and fittings, valves, wire and conduit, instruments, insulation, and paint used in constructing the plant.

Direct field labor costs are determined by estimating field labor manhours based on SCS and MWK standard practice and multiplying manhours by wage rates. In some cases, direct labor costs are supplied by the vendors who provided quotes for specific major equipment pieces or subcontracts. The direct field labor wage rate is based on appropriate crafts applicable to the project type covered in this study and reflects a closed-shop project. Union rates appropriate for Plant Wansley are used as a basis in determining the rate, which does not include indirect field costs such as benefits and burdens. It should be noted that a site location can influence field labor manhour estimates and wage rates; thus, cost estimates for the Wansley site may be substantially different from those for other locations.

Subcontract costs are the costs charged by a subcontractor for providing equipment or materials and installation. Included in these costs are the subcontractor's overheads and profits. In some instances, costs are represented in a lump-sum fashion and listed as subcontract items during the cost estimation.

Indirect field cost estimates are based on construction experience with similar projects in The Southern Company and cover the following areas:

- The labor and material costs for temporary construction facilities, such as temporary buildings, roads, utilities, and railroads.

- The labor and material costs for construction services such as medical and first aid, janitorial work, final job cleanup, handling and servicing of construction equipment, and consumables (e.g., welding supplies).
- Contractor overhead and profit, which includes field staff and expenses.
- Field staff and craft burdens and benefits. Burdens include legally stipulated costs such as FICA, workmen's compensation, public liability and property damage, and state and federal unemployment insurance. Nonlegislated benefit costs include health, welfare, vacation, sick leave, holiday, travel expenses and mileage, pension fund, and apprentice fund.
- Construction equipment (purchased and rented) and small tools.

Freight charges are included in the direct field material costs. Vendor-supplied freight costs are used when possible; otherwise, a 3-percent charge, based on total direct field material, is applied.

A sales tax rate of 4 percent is applied to the total direct equipment and material costs and to the material portion of the indirect costs and subcontract costs (where applicable). The sales taxes are included in the indirect field costs.

7.1.1.2 Total Plant Cost

The total plant cost (TPC) is the sum of the following:

- Process plant cost
- General plant facilities
- Engineering fees
- Process contingency
- Project contingency

The major items constituting general plant facilities include plant and instrument air, buildings (such as the powerhouse and control room, warehouses, fire protection house, and service and cooling water houses), cooling water system, fire protection system, fuel oil storage and handling, gasification island support including the main flare system and fuel gas piping, intrasite communication systems, potable and utility water systems, road and rail facilities, sewer and effluent water treatment, site preparation and improvements, electrical switchyard and distribution, service water system, miscellaneous pipe racks, and instrumentation and controls. Estimating the costs of general plant facilities fell completely within the scope of work for SCS. Although general plant facilities are

often estimated as a percentage of process plant costs, usually 5 to 20 percent, SCS projected these costs exactly in the same manner as above for the process plant cost. The general facilities were actually treated as a process plant section for estimating purposes but were reported outside the process plant cost category to match the DOE cost-of-electricity format. In the DOE model, the default value for general facilities is 12.5 percent of the process plant cost. The range of costs for general facilities estimated by SCS varies from 9.5 percent to 17.3 percent of the process plant cost over the six study cases.

The engineering fees cover the home office and outside contractor costs for engineering, design, and procurement for SCS and outside contractors such as MWK and GE. Also included are office and personnel expenses (e.g., computer costs, reproduction costs, and travel), licenser costs for basic process engineering, office burdens and benefits, wrap-up and all-risk insurance, environmental permitting costs, and project management costs. These fees typically range from 7 percent to 15 percent of the process plant cost, and in the DOE model the default value is 11 percent of the process plant cost. In this study, the engineering fees ranged from about 7.4 percent to 14.4 percent of the process plant cost for the various study cases.

Two separate contingencies (i.e., a process contingency and a project contingency) are applied to the capital cost estimate. The process contingency is an effort to quantify the uncertainty in the design, performance, and cost of the commercial-scale equipment. Different process contingency allowances are applied to each major subsection of the plant. Process contingencies for the gasification island areas are given in Table 7-1. These factors were developed by MWK, with review and input from SCS and DOE. One of the primary bases for these values is a relative comparison with the reference case, which is the KRW-based oxygen-blown plant with cold gas cleanup case developed in the EPRI/SCS site-specific study. The rationales for this comparison and the development of process contingency factors are discussed below for key process areas.

- **Ash Depressurization** - The process contingency for ash removal and LASH transport is increased for all cases over the reference case because the method of pneumatically transporting LASH from the gasifier to the filter was not demonstrated at the pilot plant, while ash depressurization through a lockhopper, as in the reference case, was demonstrated. This transport method is very common; but, there is some uncertainty about pressure balance, continuity of LASH removal, and related control issues. A 10-percent process contingency is assigned to this area for Cases 1 through 5 (15 percent for Case 6) versus a reference value of 5 percent.

Table 7-1

Process Contingency Factors for Gasification Island Areas

	Air-Blown	Current Study		SCS/EPRI Study	
	Cases	Hot Gas Cleanup	Case	Oxygen-Blown	
	1 & 3	Cases	Case	Cold Gas Cleanup	
	5	2 & 4	5	6	
	%	%	%	%	
OXIDANT SUPPLY					
Oxygen Plant	N/A	N/A	0	N/A	0
Boost Air Compression	0	0	0	0	N/A
GASIFICATION/ SOLIDS REMOVAL					
Coal Pressurization/ Feeding	5	5	5	5	5
Ash Depressurization/ LASH Transport	10	10	10	15	5
Gasification/ Fines Recycle	15	20	20	20	20
Heat Recovery	5	N/A	10	10	15
Fines Removal/ Depressurization	15	15	15	15	10
Gas Cooling	0	0	0	0	0
Recycle Gas System	10	15	10	15	5
Resaturation/ Startup Heater	0	0	0	0	0
Sulfation	15	15	N/A	15	N/A
Balance of Gasification Island	0	0	0	0	0
GAS CLEANUP					
Selexol®, Claus, SCOT	N/A	N/A	N/A	N/A	0
Sour Water Stripping	N/A	N/A	N/A	N/A	10
Chloride Removal	10	10	10	N/A	N/A
ZnFe Sulfur Removal	10	10	15	N/A	N/A
Acid Plant	N/A	N/A	0	N/A	N/A

N/A = Not Applicable

- **Gasification** - The reference process contingency of 20 percent is substantial because of scale-up uncertainties for the KRW gasifier. The primary uncertainties are the axial temperature profile and the required bed volumes. This contingency is in addition to a 30-percent margin on bed volume already included in the gasifier size. Regarding the temperature profile, the operating temperature will be limited by the peak temperature at the central jet, and the gasification rate (or coal capacity) will be limited by the operating temperature and carbon volume. Pilot plant data show the gasifier with in-bed desulfurization to have a much flatter temperature profile than without limestone desulfurization, thus a higher design operating temperature is allowed for cases with limestone. In addition, gasifier operation is more stable with the dense limestone. These in-bed benefits are somewhat offset by scale-up uncertainties about the limestone addition, such as actual bed density (i.e., remaining carbon volume) and required feed rate. The high-pressure gasifier (Cases 2 and 4) requires a higher process contingency than other cases because operation at these pressures has not been fully demonstrated.

Taking into account the arguments above, Cases 1 and 3 have been assigned a 15-percent process contingency due to the benefits of limestone addition. Cases 2 and 4 have been assigned a 20-percent process contingency, higher because of the pressure scale-up uncertainty. Case 5 has a 20-percent contingency because of the lack of limestone addition. Gasification in this case is similar to the EPRI reference case. Case 6 is given a 20-percent process contingency, which is greater than the values for Cases 1 and 3, and the same value as Case 5 and the reference case. This is a result of the more conceptual status of the transport type gasifier. This relationship for Case 6 is about the same for the other areas below.

- **Heat Recovery** - Heat recovery as included in Cases 1, 3, and 5, including five tube heat exchangers, has recently been demonstrated to operate satisfactorily at commercial scale with a DOW gasifier. Additionally, for Cases 1 and 3 with in-bed sulfur removal, pilot plant operation has shown that sticky ash deposition does not occur as is the case without in-bed desulfurization. Thus a 5-percent process contingency is assigned for Cases 1 and 3 and a 10-percent process contingency is assigned for Cases 5 and 6. The reference process contingency is 15 percent, which was assigned prior to experience with the DOW gasifier. Arguably, this number should have been reduced to 10 percent, but this change was not accommodated in the current study. Cases 2 and 4 do not include a primary gas cooler.

- **Fines Removal** - Fines removal and depressurization at a temperature of 1000 to 1200°F is an area where little pilot plant data exist. Work planned in various DOE clean coal projects is expected to demonstrate the reliability of vendor claims for these components over the next 5 years. Thus these process contingencies are higher than the 10-percent reference value. Cases 1 through 6 are assigned a 15-percent process contingency.
- **Recycle Gas System** - Recycle gas compression carries a comparatively higher process contingency when zinc ferrite is used for sulfur removal. The dewpoint of the gas and the compressor gas temperature are increased because of the additional moisture required to maintain the integrity of the zinc ferrite. In addition, the pressure for filter blowback is substantially higher. There are very few suppliers of compressors which operate at these higher temperature and high-head conditions. Consequently, a two- to threefold increase in process contingency over the reference process contingency of 5-percent is provided. Cases 1, 3, and 5 are assigned a 10-percent process contingency while Cases 2, 4, and 6 are assigned a 15-percent process contingency. The contingencies for Cases 2 and 4 are higher because of the higher operating pressures.
- **Sulfation** - Sulfation equipment is conventional in design, but high conversion of calcium sulfide to calcium sulfate has not been demonstrated at pilot scale, thus a significant process contingency of 15 percent is provided.
- **Hot Gas Cleanup** - Chloride removal is performed in a conventional fixed-bed reactor; however, the calcined nahcolite sorbent has been demonstrated only at a laboratory scale. Thus a 10-percent process contingency is assigned, where applicable.

While zinc ferrite has been demonstrated at pilot scale, scale-up uncertainties of sorbent life and utilization remain. In Case 5, the regeneration gas must be depressurized, cooled, and sent to an acid plant, and because of the higher sulfur concentrations in the gas stream, the cycle time is much shorter than for the cases in which zinc ferrite is used in a polishing mode. For Cases 1, 2, 3, and 4, a 10-percent process contingency is assigned, while for Case 5, a 15-percent process contingency is assigned.

The process contingency percentages for the remaining process areas for this study and the previous reference case are listed below:

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<u>Process Area</u>	<u>Current Study Cases 1-6</u>	<u>EPRI/SCS Oxygen- Blown</u>
Wastewater treatment	N/A	30
Coal receiving and handling	0	0
Limestone receiving and handling	0	0
Combustion turbine generators	5	5
Heat recovery steam generators	5	5
Steam turbine system	0	0
Ash and fines handling and disposal	0	0
General facilities	0	0

The project contingency covers the cost of additional or improved equipment resulting from a more detailed design of an actual project. The project contingency for all cases in this study is 20 percent of the sum of process plant cost, general facilities, engineering fees, and the process contingency. These are the same definition and value as used in the reference case.

7.1.1.3 Total Plant Investment

The total plant investment (TPI) is the sum of the TPC and the allowance for funds used during construction (AFUDC). AFUDC is the adjustment to the capital cost (based on TPC) to reflect the cost of using money over the period of construction. The AFUDC is calculated for Georgia Power Company using a rate of 10.83 percent compounded semi-annually. The AFUDC is also a function of the required construction schedule and the expenditure schedule during this period. The assumed construction period and expenditure schedule for AFUDC calculation are shown below:

	<u>Cases 1, 4, 5, & 6</u>	<u>Cases 2 & 3</u>
Construction period, years	4.5	4.0

Expenditure schedule

Year 1	5%	10%
Year 2	15%	20%
Year 3	30%	40%
Year 4	35%	30%
Year 5	15%	0%

The estimated AFUDC is about 19.6 percent of the TPC for Cases 1, 4, 5, and 6 and about 19.1 percent of the TPC for Cases 2 and 3.

7.1.1.4 Total Capital Requirement

The total capital requirement (TCR) is the sum of the TPI and the other remaining capital costs necessary to complete the entire project. These remaining capital items, which are often referred to as owner's costs, include the following:

- Prepaid royalties
- Initial catalyst and chemical inventory
- Startup costs
- Spare parts
- Working capital
- Land

Vendor information on prepaid royalties is considered proprietary. Therefore, a typical royalty assumption of 0.5 percent of the process plant cost is made for all cases in the current study.

The initial catalyst and chemical inventory covers the initial cost of any catalysts or chemicals contained in the process equipment. The majority of these costs in this study are for the gasification island. The major items and quantities for the gasification island are listed below:

<u>Initial charge</u>	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Case 4</u>	<u>Case 5</u>	<u>Case 6</u>
Nahcolite (tons)	740	88	370	350	640	-
Zinc ferrite (tons)	1600	284	800	1136	2042	-
Sulfuric acid (tons)	-	-	-	-	615	-
Molten sulfur (tons)	-	-	-	-	53	-
V ₂ O ₅ catalyst (liters)	-	-	-	-	65,600	-
Slaked lime (tons)	-	-	-	-	41	-

Startup costs are intended to cover preproject administrative costs and operator training, equipment checkout, changes in plant equipment, extra maintenance, and inefficient use of fuel and other materials during plant startup. These costs are estimated as follows:

- Two percent of the TPI to cover expected changes and modifications to equipment that will be needed to bring the plant up to full capacity.
- One month of fixed operating costs (operating and maintenance labor, administrative and support labor, and maintenance materials).
- One month of variable operating costs (consumables) at full capacity, excluding fuel. These variable operating costs include consumables plus waste disposal charges.
- Twenty-five percent of full-capacity fuel costs for 1 month. This charge covers inefficient operation that occurs during the startup period.

Spare parts inventory is estimated as the sum of the following:

- One-half percent of the TPC, except for the coal, limestone, and ash handling systems.
- Seven percent of the coal, limestone, and ash handling plant costs.
- Costs for major component warehouse spares required in the gasification island, which are in addition to the normal spare part cost factors (as recommended by MWK). These major spare components are itemized in Table 7-2.

The value of inventories of fuels, consumables, and byproducts is capitalized and included in the working capital account. The working capital cost includes a 60-day supply of fuel and consumables and a 30-day byproduct inventory.

Land costs are assumed to be zero since Georgia Power Company is the owner of the property at Plant Wansley. The available land is sufficient for all the IGCC plant areas and 40-years storage of ash and fines.

7.1.2 Capital Cost Results

The capital cost results are presented in detail as the output from the DOE cost-of-electricity model in Appendix E for all of Cases 1 through 6. The capital requirements are summarized in both millions of dollars and \$/kW for Cases 1 through 6 in Table 7-3. Because of the net output megawatt differences among the cases, most of the comparisons which follow are made with reference to the \$/kW values. A more detailed breakdown of process plant costs by plant section is provided in Table 7-4. A breakdown of owner's costs is given in Table 7-5.

Table 7-2

Major Component Warehouse Spares

Items	Numbers of Units					
	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Case 4</u>	<u>Case 5</u>	<u>Case 6</u>
Ceramic candle filters						
No. operational elements	3928	425	1964	1700	3576	3285
No. spare elements	400	100	300	300	500	400
Hot gas valves (18 inch)						
No. operational valves	32	8	16	32	48	--
No. spare valves	2	1	1	2	3	--
Hot solids valves						
No. operational valves	20	5	10	20	30	24
No. spare valves	4	2	2	4	6	4
Sulfator conveyor						
No. of operational units	3	1	2	2	--	3
Spare units 602V	1	1	1	1	--	1
Spare units 603V	1	1	1	1	--	1

Table 7-3

Capital Cost Summary
Mid-1990 Dollars

<u>CAPITAL COST</u>	Case 1 \$x10 ⁶ \$/kW	Case 2 \$x10 ⁶ \$/kW	Case 3 \$x10 ⁶ \$/kW	Case 4 \$x10 ⁶ \$/kW	Case 5 \$x10 ⁶ \$/kW	Case 6 \$x10 ⁶ \$/kW
Net power output, MW	458.4	99.6	231.7	399.1	420.2	409.6
Plant construction cost	378.1	825	120.1	1206	218.7	944
Process contingency	19.8	43	6.1	61	10.4	45
Project contingency	79.6	174	25.3	254	45.8	197
Total plant cost	477.5	1042	151.5	1521	274.9	1186
AFUDC	93.7	204	28.9	290	52.4	226
Total plant investment	571.2	1246	180.4	1811	327.3	1413
Owner's costs	44.0	96	13.5	136	24.9	107
Total capital requirement	615.2	1342	193.9	1947	352.2	1520
					582.9	1461
					677.3	1612
					524.0	1279

Note: Plant construction cost = Process plant cost plus general facilities and engineering fees

Table 7-4
Breakdown of Process Plant Costs by Plant Section

Mid-1990 Dollars																			
		CASE 1			CASE 2			CASE 3			CASE 4			CASE 5			CASE 6		
		458.4			99.6			231.7			399.1			420.2			409.6		
Area No.	Plant Section Description	K\$	\$/KW		K\$	\$/KW		K\$	\$/KW		K\$	\$/KW		K\$	\$/KW		K\$	\$/KW	
100	Coal receiving/handling	15,901	34.7		8,907	89.4		11,405	49.2		15,947	40.0		16,245	38.7		18553	45.3	
150	Limestone receiving/handling	9,072	19.8		6,607	66.3		7,532	32.5		9,141	22.9		0	0.0		7628	18.6	
250	Booster compression	8,822	19.2		2,866	29.0		4,450	19.2		11,456	28.7		9,272	22.1		7933	19.4	
300	Gasification	55,163	120.3		14,178	142.3		26,643	123.6		49781	124.7		63,074	150.1		40622	99.2	
380	Recycle gas compression	13,939	30.4		4,091	41.1		7,054	30.4		14965	37.5		18,044	42.9		5855	14.3	
400	Gas conditioning	35,212	76.8		9,553	95.9		17,819	76.9		37144	93.1		38,645	92.0		12740	31.1	
500	External desulfurization	20,171	44.0		7,524	75.5		10,213	44.1		29336	73.5		32,697	78.3		0	0.0	
600	Sulfation	13,095	28.6		2,768	27.8		7,062	30.5		9133	22.9		0	0.0		21383	52.2	
700	Sulfuric acid plant	0	0.0		0	0.0		0	0.0		0	0.0		25,739	61.3		0	0.0	
900	Gas turbine system	70,790	154.4		23,702	238.0		36,123	155.9		91005	226.0		70,965	168.9		70737	172.7	
1000	HRSG system	24,929	54.4		6,106	61.3		13,054	56.3		22077	55.3		32,092	76.4		24914	60.8	
1100	Steam turbine system	46,268	100.9		1,945	19.5		25,643	110.7		6995	17.5		41,604	99.0		41926	102.4	
1200	Ash & fines handling/disposal	5,788	12.6		2,921	29.3		3,942	17.0		5658	14.2		4,491	10.7		4780	11.7	
Total Process Plant Cost		319,136	696.1		91,188	915.5		172,940	746.4		302,638	758.3		353,068	840.2		257,071	627.6	
General plant facilities		33,484	73.0		15,743	158.1		26,307	113.5		29,379	73.6		33,365	79.4		35,591	86.9	
Engineering fees		25,470	55.6		13,176	122.3		19,486	84.1		24,464	61.3		26,103	62.1		36,121	88.2	

Table 7-5
Breakdown of Owner's Costs
Mid-1990 Dollars

	Case 1		Case 2		Case 3		Case 4		Case 5		Case 6	
	K\$	\$/KW	K\$	\$/KW	K\$	\$/KW	K\$	\$/KW	K\$	\$/KW	K\$	\$/KW
Prepaid royalties	1,890	4.1	601	6.0	1,094	4.7	1,782	4.5	2,063	4.9	1,844	4.0
Initial cat. & chem. inventory	10,642	23.2	1,883	18.9	5,322	23.0	7,504	18.8	13,512	32.2	59	0.1
Startup costs	14,727	32.1	4,684	47.0	8,382	36.2	13,960	35.0	16,700	39.7	12,400	30.3
Spare parts	6,893	15.0	4,032	40.5	5,142	22.2	7,064	17.7	6,219	14.8	5,378	13.1
Working Capital	9,883	21.6	2,329	23.4	4,957	21.4	9,213	23.1	11,813	28.1	7,688	18.8
Land	0	0.0	0	0.0	0	0.0	0	0.0	0	0.0	0	0.0
Total Owner's Costs	44,035	96.1	13,529	135.8	24,907	107.5	39,543	99.1	50,307	119.7	27,169	66.3

Case 6 possesses the lowest total capital requirement (TCR) of all the cases at 1279 \$/kW. The TCR for Case 1 is the next lowest at 1342 \$/kW. This cost advantage for Case 6 of about 5 percent is primarily related to the rapid gasification and sulfur capture reaction rates predicted for the transport reactor, leading to significant reductions of gasification island equipment and material costs. (The predicted performance of the transport gasifier will be verified in future experimental programs.)

The highest TCR is for Case 2 at 1947 \$/kW, primarily a result of the smallest megawatt rating among the cases studied. For Case 4, which represents a fourfold ISTIG plant size increase over Case 2, the TCR decreases about 25 percent to 1461 \$/kW. However, the Case 4 TCR remains about 14 percent greater than for Case 6 and about 9 percent greater than for Case 1. The primary reason for this is the lower efficiency of the ISTIG plant configurations in this study. This lower efficiency is due to the inability to fully utilize the steam generating capability of the cycle due to mass flow limits assumed for the LM-8000 combustion turbines.

The TCR results from Table 7-3 are also plotted in Figure 7-1 as \$/kW versus net plant capacity in megawatts. This figure shows the economies of scale for ISTIG Cases 2 and 4, and for IGCC Cases 1 and 3. While the Case 4 (399 MW) TCR is 25 percent less than that for Case 2 (100 MW), the Case 1 (458 MW) TCR is about 12 percent less than that for Case 3 (231.7 MW). The scattered TCR results near the 400-MW capacity in Figure 7-1 for Cases 1, 4, 5, and 6 indicate the potential range in costs for the various gasification plant designs considered in this study.

Figure 7-2 is a bar chart displaying the TCR results broken down into the following four components:

- Plant construction costs = process plant cost + general facilities + engineering fees
- Contingencies = process contingency + project contingency
- AFUDC
- Owner costs = royalties + initial inventory + startup costs + spare parts working capital + land

As noted in Figure 7-2, the plant construction cost is the major component of the TCR, representing about 60 percent of the TCR in each case. Also, variations in the plant construction costs account for the majority of the differences in the TCR results among the six cases. The contingency and AFUDC items are the next

Figure 7-1 Capital Cost vs. Plant Size

Figure 7-2 Capital Cost of all Cases by Cost Element

most costly components. These are about equal for the various cases, each representing approximately 15 to 17 percent of TCR owner's costs.

A breakdown of the process plant costs by major plant area groupings is given in Figure 7-3. Based on the grouping of area process section costs in this figure, the single largest component in the process plant costs for all cases is the power island (composed of the gas turbine system, the HRSG system, and the steam turbine system). If the gasification and gas conditioning components are combined, the joint costs are slightly greater than the power island costs for most cases except Case 6. Case 6 has much smaller gasification island requirements because of the high gasification and sulfur capture efficiency predicted for the transport reactor. Most of the differences in the total process plant costs among the six cases results from variations in the materials handling, gasification, and gas conditioning costs, since power island equipment is virtually identical for similar size plants.

7.2 OPERATING AND MAINTENANCE COST ESTIMATES

7.2.1 Estimating Assumptions and Procedures

The O&M costs are estimated for a year of normal operation at a 65-percent capacity factor and are expressed in mid-1990 dollars. The 65-percent capacity factor was used because of the DOE requirement that costs of electricity be developed on this basis. The major unit costs for annual O&M items are shown in Table 7-6. O&M costs include fuel, variable O&M, and fixed O&M, which are described in the following paragraphs.

The fuel type for all cases is Illinois No. 6 coal. The DOE cost model (Appendix E) calculates the annual fuel cost for each plant from the specified capacity factor, coal rate, and unit price. The delivered cost of coal used in this study for Plant Wansley is \$35/ton, based on the as-received coal.

The variable O&M is composed of consumable costs, ash/sorbent disposal costs, and byproduct credits (applicable only for Case 5). Consumable costs cover the consumption of limestone, nahcolite, zinc ferrite, and other miscellaneous items. The annual costs for limestone, nahcolite, and zinc ferrite are calculated based on material balance data for each case. The miscellaneous items include the following:

- Fuel oil for startups
- Natural gas for operation of the emergency flare
- Cooling water treatment chemicals

Figure 7-3 Process Plant Cost Component Comparison

Table 7-6

Major Unit Costs for Annual O&M Items

<u>Cost Item</u>	<u>\$/Unit</u>
Fuel, Illinois No. 6 coal	35.00/ton
Limestone	11.20/ton
Nahcolite	261.25/ton
Zinc ferrite	6,270.00/ton
Solids disposal,	
Cases 1, 2, 3, 4, and 6	2.94/ton
Case 5 (No limestone)	4.40/ton
O&M labor, wage rate (average including burden)	21.80/hr
Byproduct credit, sulfuric acid	52.25/ton

- Service water corrosion inhibitor
- Demineralization
- Nitrogen and hydrogen for inerting, purging, and cooling
- For Case 5, V_2O_5 catalyst, slaked lime, selective catalytic reduction operating costs, and molten sulfur

Ash/sorbent disposal costs are another variable O&M cost. A disposal unit cost of \$2.94/ton is used for all cases except Case 5, where a rate of \$4.40/ton is assumed. The greater rate for Case 5 reflects the difference in waste product characteristics, notably the higher density and greater percentage of carbon. These unit costs are representative of current ash disposal costs being incurred by plants within The Southern Company.

In Case 5, sulfuric acid is produced as a salable product. A byproduct credit of \$52.25/ton of sulfuric acid is included in the variable O&M costs, reducing the net O&M annual cost. This price assumes that the market for sulfuric acid is not saturated. It should be noted that the IGCC plant of Case 5 would produce a substantial share of the current market for sulfuric acid.

The fixed O&M components consist of operating labor, supervision and clerical labor, maintenance material and labor, insurance and taxes, and general and administrative (G&A) expenses. The basis for estimating these fixed costs is summarized in Table 7-7 and further described below.

The operating labor for all cases is based on estimates of the number of operating jobs required per shift as shown in Table 7-7. A shift multiplier of 4.2 is assumed. The labor rate of \$21.80 per hour includes a 40-percent payroll burden and reflects an average O&M labor rate appropriate for Plant Wansley under the assumptions of the current study.

Supervision and clerical labor is estimated as 20 percent of the operating and maintenance labor. In many studies, supervision and clerical is taken as 30 percent of the O&M labor. A lower value is used here, however, because this study is based on an existing plant site with some of the necessary administrative and support labor already in place.

For studies in which maintenance costs are not known from experience, the EPRI TAG recommends that total maintenance be estimated as a percentage of process plant costs. A maintenance material-to-labor ratio of 60/40 is then applied to these costs. For the current study, however, maintenance labor costs were estimated directly from plant staffing plans. The annual maintenance labor cost is equal to the number of personnel times the burdened wage rate (\$21.80/hr) times 1920 hours per year. Maintenance material costs were estimated based on a weighted average of percentages applied to individual process plant area costs as shown in Table 7-8.

These percentages were developed based on an analysis of The Southern Company O&M costs as well as a number of published studies of gasification-based power plants. The total annual maintenance cost is the sum of the maintenance labor and material estimates.

Table 7-7

Fixed O&M Cost Basis

<u>Item</u>	<u>Basis</u>					
Maintenance material	Weighted average from projected percentage of area process capital					
Supervision and clerical	20% of operating and maintenance labor					
Insurance and taxes	6.6% of total plant cost					
G&A expenses	10% of operating and maintenance labor					
<u>Plant Staffing</u>	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Case 4</u>	<u>Case 5</u>	<u>Case 6</u>
Operating labor, persons/shift	23	15	17	21	24	21
Maintenance labor, total persons	30	15	20	30	30	30

Table 7-8

**Maintenance Material Cost as Percentage of
Process Plant Cost by Area**

<u>Area No.</u>	<u>Area description</u>	<u>Maintenance Material % of PPC*</u>
100	Coal receiving, handling	1.2
150	Limestone receiving, handling	1.2
250	Booster compression	2.4
300	KRW gasification	2.4
380	Recycle gas compression	2.4
400	Gas conditioning	2.4
500	External desulfurization	2.4
600	Sulfation	2.4
700	Sulfuric acid plant	2.4
900	Gas turbine system	0.9
1000	HRSG system	0.9
1100	Steam turbine system	0.9
1200	Ash and fines handling and disposal	1.2
1300	General facilities	0.9

Weighted Averages, Maintenance Materials (% of PPC)

Case 1	1.54
Case 2	1.52
Case 3	1.50
Case 4	1.61
Case 5	1.64
Case 6	1.43

* These percentages are applied to the sum of each area process plant cost and the prorated engineering fees, but excluding contingencies, to yield a total weighted average percent. The resulting value is multiplied by TPC to produce maintenance material costs. Maintenance labor costs are estimated based on plant staffing plans.

Insurance and local taxes are calculated as a percentage of the total plant cost. A 0.6-percent value, based on The Southern Company data, is used. In the current study, insurance and local taxes are included as an annual O&M charge for calculation of cost of electricity, as is stipulated in the DOE cost model. The Southern Company accounting normally treats these as components of the capital carrying charge, which is consistent with the EPRI TAG approach.

No annual royalty costs are assumed for cases in this study. (Prepaid royalties were included as a capital cost item.)

The final annual cost is general and administrative expenses. These are estimated as 10 percent of O&M labor and are a fixed O&M cost component.

7.2.2 Total and Net O&M Cost Results

The first-year O&M costs are summarized in thousands of dollars and in mills/kWh or \$/kW-yr in Table 7-9. Figure 7-4 shows the results in mills/kWh with O&M broken down into fuel, variable, and fixed components. All costs are in mid-1990 dollars. As noted, these costs are calculated for a capacity factor of 65 percent in keeping with DOE requirements.

The lowest first-year O&M cost is for Case 6 at 19.8 mills/kWh. This is primarily a result of lower fuel and variable costs. Case 6 has the best predicted heat rate which leads to the lowest fuel consumption and cost of the six cases (mills/kWh basis). Also, the predicted sulfur removal is better in the gasifier for Case 6, with less limestone consumption and no zinc ferrite system requirements; therefore, the variable costs are significantly less for Case 6. Case 1 has the next best O&M cost at 22.0 mills/kWh. This represents an increase of approximately 11 percent over Case 6.

The highest annual O&M cost belongs to the 100-MW ISTIG plant, Case 2, with 30.3 mills/kWh. This high cost is due primarily to its small size. Although the fuel and variable costs are reduced nearly proportionately to the size, the fixed costs are not reduced and become very significant for Case 2. For Case 4, which is a fourfold larger ISTIG plant than Case 2, the annual O&M cost is 23.8 mills/kWh. This represents a 22-percent reduction from the Case 2 results, due to economies of scale. For both ISTIG cases, the relatively higher production costs are also affected by the higher fuel costs for these designs. As explained previously, this is due to the lower thermal efficiencies of these plants relative to the IGCC designs. The lower efficiencies arise because the plant designs could not be optimized from a thermal standpoint due to gas turbine limitations.

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Table 7-9

First Year O&M Cost Summary*

	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Case 4</u>	<u>Case 5</u>	<u>Case 6</u>
<u>Net MW (90°F)</u>	458.4	99.6	231.7	399.1	420.2	409.6
<u>Fuel, \$ x 1000</u>	33,656	7,917	16,828	31,659	33,576	28,513
<u>Variable O&M, \$ X 1000</u>						
Limestone	2,885	679	1,441	2,712	0	1,182
Nahcolite	296	71	148	284	296	0
Zinc ferrite	1,648	299	824	1,193	9,653	0
Miscellaneous	595	243	360	582	1,449	703
Solids disposal	859	202	429	808	564	557
Total, \$ X 1000	6,283	1,494	3,202	5,578	11,961	2,441
<u>Fixed O&M, \$ X 1000</u>						
Operating labor	4,380	2,857	3,238	3,999	4,571	3,999
Supervision	1,127	697	815	1,051	1,165	1,051
Maintenance	8,639	2,941	4,953	8,558	9,839	7,189
Insurance/taxes	2,865	909	1,649	2,726	3,145	2,492
Other	564	348	407	526	583	526
Total, \$ X 1000	17,575	7,752	11,062	16,860	19,302	15,258
<u>Byproduct Credit, \$ X 1000</u>	0	0	0	0	3,471	0
<u>Total First Year O&M Costs, \$ X 1000</u>	57,514	17,162	31,093	54,097	61,368	46,212
<u>Fuel, mills/kWh</u>	12.89	13.96	12.76	13.93	14.03	12.23
<u>Variable w/o byproduct credit, mills/kWh</u>	2.41	2.63	2.43	2.45	5.00	1.05
<u>Variable w/ byproduct credit, mills/kWh</u>	2.41	2.63	2.43	2.45	3.55	1.05
<u>Fixed, \$/kW-yr</u>	38.3	77.8	47.7	42.2	45.9	37.2
<u>Total First Year O&M Costs mills/kWh</u>	22.03	30.26	23.57	23.81	25.65	19.81

* Based on mid-1990 dollars and 65-percent capacity factor. Capacity factor of 65 percent specified by DOE for comparison to other studies.

Figure 7-4 Comparison of O&M Cost Components

As shown in Figure 7-4, fuel is the single highest production cost for all cases. For all cases but Case 2, fuel costs range from 54 to 62 percent of the total O&M (mills/kWh basis). In Case 2, fuel and fixed costs are almost identical portions of the total at 46 and 45 percent, respectively. Case 5 possesses the greatest variable cost at 14 percent of the total annual O&M. Even though Case 5 is the only case with the byproduct credit and no limestone consumption, these reductions are insufficient to overcome the effect of the higher zinc ferrite consumption coupled with the high unit cost of zinc ferrite sorbent. Case 6 has the lowest proportion of variable costs at 5 percent of the total O&M. As described above, this is a result of the reduced limestone consumption requirements in the gasifier and elimination of nahcolite and zinc ferrite consumption.

8.0 ECONOMIC AND FINANCIAL ANALYSIS

8.1 COST OF ELECTRICITY DETERMINATION

8.1.1 Methodology

In order to evaluate the relative performance of the various study cases on a busbar cost basis, the levelized cost of electricity (COE) was calculated using the cost data described in Section 7. The U.S. Department of Energy's cost-of-electricity program, a Lotus 1-2-3™ spreadsheet developed at METC, was used to calculate the costs of electricity for all cases. The DOE spreadsheet embodies standard busbar electricity cost calculations. The DOE spreadsheet has a number of built-in costing assumptions which are consistent with experience and utility costing approaches and is generally consistent with EPRI's Technical Assessment Guide (TAG).

The first two portions of the DOE spreadsheet calculate the total capital requirement and the net annual operating costs, based on the input and calculation methods described in the preceding section of this report. Upon completion of the data entry, the COE is also automatically calculated and displayed. The COE is stated in terms of 10-year levelized costs, in both current and constant dollars. The COE is broken down into capital, fuel, and operating and maintenance cost components.

8.1.2 Assumptions

The basic financial parameters assumed in this study for COE calculation are given in Table 8-1. The project, book, and tax lives; the tax depreciation method; and the investment tax credit values were specified by the DOE COE method. The values listed for federal and state income tax rate and the financial structure are based on The Southern Company data base for Georgia Power Company requirements. The inflation and escalation values are from SCS economic forecasts.

One of the most significant assumptions is the capacity factor of 65 percent. As will be seen later in the sensitivity results, the capacity factor may substantially affect the COE results. The assumption of a 65-percent capacity factor was stipulated by DOE to facilitate comparison with other study results. The capacity factors for these plants in The Southern Company system would be substantially higher, as indicated by the system production cost analysis of Case 1 described in Section 8.3.

The levelizing factors listed in Table 8-1 are calculated from the financial assumptions given in the same table. These factors are shown for both current and constant dollars and are for 10-year levelized costs.

Table 8-1

**Base Case Economic Assumptions
for Cost of Electricity Calculation**

Project life	30 years
Book life	30 years
Tax life	30 years
Federal and state income tax rate (composite)	37.7%
Investment tax credit	0%
Tax depreciation method	ACRS
Inflation rate	4.5% per year
Real escalation rates (over inflation)	
Fuel	0.5% per year
Operating & maintenance	0% per year
Year dollar basis	mid-1990
Capacity factor	65%

Financial structure

<u>Type of Security</u>	<u>% of Total</u>	<u>Current Cost</u>	<u>Dollar. % Return</u>	<u>Constant Cost</u>	<u>Dollar. % Return</u>
Debt	45	10.3	4.6	5.6	2.5
Preferred stock	10	9.8	1.0	5.1	0.5
Common stock	45	13.8	6.2	8.9	4.0

Discount rate, % (cost of capital)	11.8	7.0
---------------------------------------	------	-----

<u>Levelizing factors, (10-year)</u>	<u>Current Dollar. %</u>	<u>Constant Dollar. %</u>
Capital carrying charge	0.169	0.113
Fuel	1.265	1.025
Operating & maintenance	1.234	1.000

8.1.3 Summary: Total Costs of Electricity

The COE results are presented in detail for each case in Appendix E and are summarized in Table 8-2. The COE results are shown as 10-year levelized mills/kWh costs, in both current and constant dollars, based on a 65-percent capacity factor. The comparisons discussed below are made on a constant dollar basis.

As with TCR and annual O&M results, Case 6 possesses the lowest COE with 45.6 mills/kWh. Case 1 is the next lowest with a COE of 49.1 mills/kWh, 8 percent greater than that for Case 6. Most of the difference in COE for Case 6 versus the other cases is in its lower O&M cost component, but also, to a lesser degree, its lower capital charges. The capital cost differences are a result of the predicted rapid gasification and sulfur capture reaction rates for the transport reactor, leading to reductions of the gasification island equipment and material costs, which in turn result in lower TCR and capital charges. In addition, the lower O&M costs for Case 6 result from higher gasifier efficiency, reduced limestone usage, and the elimination of nahcolite and zinc ferrite consumption.

The highest COE is for ISTIG Case 2 at 69.4 mills/kWh. This is a result of much greater capital charges and O&M components, primarily resulting from the smaller plant size. The COE for ISTIG Case 4, which is four times larger than Case 2, decreases to 53.2 mills/kWh, about a 23 percent reduction from Case 2 results. As mentioned in the previous section, the ISTIG case COE results are also adversely affected by the lower thermal efficiencies of these cases due to fundamental equipment limitations.

The COE for Case 5 is the highest among the COE results for cases with plant sizes near 400 MW. The Case 5 COE is 58.1 mills/kWh, which is 27 percent greater than that for Case 6. The higher COE for Case 5 is primarily a function of the larger capital charges (from a greater TCR) and higher O&M costs (mainly due to higher variable O&M costs from larger zinc ferrite consumption). Improvements in technology and design approach for IGCC with bulk ZnFe desulfurization and sulfuric acid production, as in Case 5, could potentially achieve a substantial reduction in COE. Possible improvements are discussed in some detail in Appendix B.

The COE results in constant mills/kWh, are also presented graphically in Figure 8-1. The figure illustrates that capital charges represent the single largest component of COE in all six cases. Capital charges range from 54 to 56 percent of the total COE. The fuel cost component is the second largest COE component in all the cases except for Case 2, where the O&M component is the second largest COE component. The dominance of capital charges in the COE results reinforce the importance of achieving capital cost reductions in gasification-based technologies.

Table 8-2

Levelized Cost of Electricity Summary*

	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Case 4</u>	<u>Case 5</u>	<u>Case 6</u>
CURRENT \$, mills/kWh						
Capital Charges	39.8	57.7	45.1	43.3	47.8	37.9
Fuel Costs	16.3	17.7	16.1	17.6	17.7	15.5
Operating & Maintenance	11.3	20.1	13.3	12.2	14.3	9.4
Total Cost of Electricity	67.4	95.5	74.5	73.1	79.9	62.8
CONSTANT \$, mills/kWh						
Capital Charges	26.7	38.8	30.3	29.1	32.1	25.5
Fuel Costs	13.2	14.3	13.1	14.3	14.4	12.5
Operating & Maintenance	9.1	16.3	10.8	9.9	11.6	7.6
Total Cost of Electricity	49.1	69.4	54.2	53.2	58.1	45.6

* Ten-year levelized at 65-percent capacity factor with mid-1990 dollar basis.

Case 1	KRW IGCC	458 MW
Case 2	KRW ISTIG	100 MW
Case 3	KRW IGCC	232 MW
Case 4	KRW ISTIG	399 MW
Case 5	KRW IGCC ZnFe Bulk Desulfurization	420 MW
Case 6	TRANSPORT IGCC	410 MW

**10-Yr Levelized mills/kWh at 65% Capacity Factor
Constant mid-1990 Dollars**

Figure 8-1 Cost of Electricity by Components for Gasification-Based Power Plants

8.2 COST-OF-ELECTRICITY SENSITIVITY STUDIES

Economic analyses generally are carried out using best estimates for all the factors which affect the final cost of electricity. It is also important to evaluate the uncertainties in the expected cost. Therefore, a simple sensitivity analysis for the COE was conducted to give some feeling for the range of uncertainty present in a cost estimate.

In this study, an attempt has been made to at least see that the ranges of the independent variables are realistic and correspond to the actual magnitude of uncertainty involving that variable. The major categories of uncertainty present in the estimated COE are:

- Cost of construction
- Fuel cost
- Capacity factor
- Efficiency of plant operation
- O&M costs

Each of these categories represents a different type of uncertainty and requires a different approach to reducing uncertainty. Some are amenable to additional engineering effort, and some are essentially outside the influence of the engineering organization. One objective of this sensitivity study is to help identify the areas where added engineering effort could have the greatest impact.

8.2.1 Sensitivity Studies Based on Case 1 (KRW IGCC, 458 MW)

8.2.1.1 Cost of Construction

There are many factors which can cause the actual cost of building a power plant to differ from the estimated cost, including such things as: inflation in equipment prices, interest rate on debt, labor rates, work stoppages, schedule delays due to permits, changing regulatory requirements, lack of detail or oversight in the engineering design, and inadequate design criteria. For first-of-a-kind plants, the last two factors are usually the most important.

In order to allow for uncertainties in preliminary-type cost studies, cost estimates typically include two contingency factors. One is the process contingency, which may be separately calculated for each section of the plant. Certain plant sections, such as coal receiving and the steam turbine system, are judged to be fully mature technologies requiring no process contingency. The power block, including the combustion turbines, heat recovery steam generator, and steam turbine, involves fairly well defined technologies for which reasonably reliable construction cost estimates can be made. Other sections carry substantially more risk and are thus assigned process contingencies of up to 15 percent.

In addition to the process contingencies for individual plant sections, the cost estimates in this study incorporate an overall project contingency calculated as a percentage (20 percent) of total process plant cost plus contingencies. This results in a contingency-upon-a-contingency, which may be too conservative. The general idea of a project contingency is to cover the cost of additional equipment or other costs which may arise from site-specific factors not known at the time of a cost estimate. For a cost estimate which is based on a specific site from the beginning, such as has been the case for the Plant Wansley site, it may be appropriate to reduce or drop the project contingency. The effect of removing all contingencies is shown for Case 1 in Figure 8-2. The leveled COE drops approximately 11 percent from 67.4 mills/kWh to 60.1 mills/kWh (44.1 mills/kWh in constant dollars). However, an overrun of 20-percent in estimated process plant costs would not be highly improbable. As seen in Figure 8-1, a 20-percent increase in process plant costs increases the expected COE by approximately 10 percent to 74.3 mills/kWh (53.9 mills in constant dollars).

Most of the uncertainty in capital costs is involved with the gasification island, because the major processes therein have little or no commercial history. Processes within the gasification island include KRW coal gasification, hot gas particulate removal, hot gas sulfur removal (zinc ferrite), and a sulfator unit. An increase of 40 or 50 percent in process plant costs for this block would not be out of the question. As illustrated in Figure 8-3, a 45-percent increase in gasification island costs results in an electricity cost of 74.9 mills/kWh, almost identical to the result obtained with an overall capital cost increase of 20 percent (Figure 8-1). Figures 8-2 and 8-3 may be considered as illustrating basically the same point.

8.2.1.2 Cost of Fuel

The relationship between cost of fuel and COE is straightforward. At a constant heat rate, an increase in coal cost creates a directly proportional increase in electricity cost. For the base case conditions, each \$2 per ton increase in coal cost raises the COE by approximately 1 mill per kWh.

This study is being concluded during yet another Middle East crisis (1990-91). It is unlikely that observed spikes in oil prices will have a large effect on coal prices as happened in the 1970s. There is little opportunity remaining in the United States to switch from oil to coal. Spot prices for coal remained mostly steady in the first 2 months following Iraq's invasion of Kuwait. This reflects an

Figure 8-2 Sensitivity to Capital Cost, Case 1 (KRW IGCC, 450 MW)

Figure 8-3 Sensitivity to Gasification Island Cost, Case 1

oversupply in the coal market which has existed for several years. However, the doubling of oil prices, if even partially sustained for an extended period, would exert an upward pull on coal prices, and the 0.5 percent per year coal price escalation which is built into the base case would then probably be too low. A doubling of coal prices, in real terms, over a 30-year period would not be unrealistic. This would correspond to an escalation rate of 2.5 percent per year. As seen in Figure 8-4, the COE is very insensitive to gradually escalating coal costs. High future costs are discounted and have a much smaller effect than immediate cost increases.

8.2.1.3 Capacity Factor

One parameter which has a dominant influence on electricity costs is the capacity factor at which the plant operates. Base case evaluations were all run at a capacity factor of 65 percent. However, the availability analysis indicates that the plant would be available 85 percent of the time, assuming a planned outage of only 412 hours per year. In The Southern Company, the average plant availability is currently running about 90 percent. Modeling the Southern electric system with the PROMOD simulator shows that the Plant Wansley IGCC facility would be dispatched as a baseload plant, and therefore the capacity factor could approach the equivalent availability factor.

Figure 8-5 illustrates the strong dependence of electricity cost on capacity factor. If a capacity factor of 85 percent can be achieved, the COE drops nearly 17 percent to 56 mills/kWh (41 mills on a constant dollar basis). However, if the plant is unable to operate as expected, the COE climbs sharply. At a capacity factor of 55 percent, electricity would increase 12.8 percent to over 76 mills/kWh (55 mills on a constant dollar basis).

8.2.1.4 Efficiency of Plant Operation

At the design point heat rate of 8,595 Btu/kWh, the overall net thermal efficiency (higher heating value basis) is 39.7 percent. The plant gross power output is 493.8 MW, of which 35.4 MW is consumed within the station, leaving 458.4 MW as the net power produced. There are a number of ways in which actual plant efficiency might fail to achieve the design value:

- Carbon conversion within the gasifier may be less complete than expected.
- Stoichiometric imbalances within the gasifier or other factors may result in a decreased cold gas efficiency at constant carbon conversion.

Figure 8-4 Sensitivity to Coal Escalation Rate

Figure 8-5 Sensitivity to Capacity Factor, Case 1

- Auxiliary power consumption within the plant may be higher than anticipated.
- Heat exchanger performance losses, high acid dewpoint, or other factors may lead to a higher stack temperature than expected.

Carbon conversion within the gasifier may be one of the largest unknowns in scaling up the KRW gasifier. There are two major responses to a decrease in carbon conversion within the gasifier. One would be to increase the coal feed to the gasifier to keep the same amount of fuel gas going to the turbine, with an associated increase in the amount of carbon in the gasifier char. The other would be to decrease the gas make in accordance with the drop in carbon conversion.

If high-carbon char from the gasifier could not be utilized, it would have to be disposed. For the design used in this study, char disposal probably represents an unrealistic case because of the fact that the system already includes a fluidized-bed combustor (the sulfator) which can be used to recover energy from any unconverted char exiting the gasifier system. However, a limitation on sulfator capacity or other design bottlenecks could make it necessary to discard char. A char disposal case was calculated by increasing coal feed so as to keep gas flow to the turbine constant, adding \$31 million in capital investment for char cooling and handling, and increasing ash disposal volume. The result is an increase of 3.9 mills/kWh (2.9 mills on a constant collar basis) if the carbon conversion in the gasifier drops from 96.5 to 86.5 percent. This represents approximately 5.8 percent.

An optimistic evaluation assumes that any unconverted char exiting the gasifier system can simply be burned up in the fluidized-bed sulfator to raise steam, and thereby recover the energy in the char. But if this steam is assumed to be worth only about 25 percent as much as producing more syngas to the turbine, coal feed rates must be increased by 75 percent of the char carryover to the sulfator to maintain output. Using this approach, the effect on the COE is less than 2 mills (3 percent) for a 10-percent decrease in carbon conversion. Therefore, lower coal conversion in the gasifier has only a minor effect if burnup can be completed in the sulfator and the system can handle the higher material flows. This can be seen as an important advantage of having the sulfator in the design.

A worst case evaluation of the effect of carbon conversion can be constructed by assuming that the sulfator can handle increased char burnup but that coal feed to the gasifier cannot be increased and plant output drops by 75 percent of the decrease in carbon conversion. In this case, if the gasifier carbon conversion falls from 96.5 percent to 86.5 percent, the COE increases by 5.7 mills/kWh (4.2 mills on a constant dollar basis), representing 8.5 percent above the base case COE.

Carbon conversion would be expected to influence the cold gas efficiency (thermal efficiency of converting coal to syngas feed to the turbine) more than any other factor. However, there may be independent effects, such as improper flow distribution within the gasifier, which could result in conversion to carbon dioxide instead of carbon monoxide, for instance. Other factors affecting the cold gas efficiency include the amount of recycle gas used to fluidize the cyclone diplegs and the agglomerated ash discharge stream. It is difficult to estimate how such changes would affect costs of operation without carrying out a process simulation. The maximum possible effect would be a drop in plant output proportional to the loss in cold gas efficiency. A lesser effect could be calculated by assuming that only the combustion turbine output is affected by the drop in cold gas efficiency, and that total output of the steam turbine remains constant. On this basis, a decline in cold gas efficiency from 69.7 to 68.7 percent would create a drop of 4.47 MW in plant output and increase the COE by 0.7 mills/kWh, or 1 percent of the base value.

Station service consumes 35.4 MW of plant power output. If the in-plant power needs turn out to be 20 percent higher than estimated, the result is an increase of 1.1 mills/kWh in the cost of electricity (0.7 mills on a constant dollar basis).

Heat exchanger performance may be degraded by any number of factors related to design or to operating conditions. Alternately, concerns about cold-end corrosion could limit stack gas heat recovery. A higher stack temperature is the result, with less steam being generated for the steam turbine. With a normal combustion turbine exit temperature of about 1,100°F and a stack temperature of 280°F, an increase of 20 degrees in stack temperature would result in a loss of 4.6 MW in output from the steam turbine. This translates into an increase of 0.7 mills/kWh in the COE, or again 1 percent.

It would be unlikely to experience decreased carbon conversion, additional decreases in cold gas efficiency, increased auxiliary power consumption, and higher stack temperatures all at the same time. However, it is of interest to see what magnitude of electricity cost would result from a combined effect situation. The data are plotted in Figure 8-6, on a cumulative basis. That is, each line in the figure represents the total effect of all parameters listed below the line. The effect of lower carbon conversion is based on a maximum decrease of 10 percent in gasifier conversion efficiency. The effect of lower cold gas efficiency is based on a maximum decrease of 3 percent in cold gas efficiency and application of the result only to the combustion turbine. The effect of increased auxiliary power is based on a maximum increase of 20 percent above the estimated value. The effect of high stack temperature is based on a maximum increase of 20 degrees.

Figure 8-6 Cumulative Sensitivity to Plant Efficiency Factors (Case 1)

The cumulative effect of potential lower plant efficiency factors (Figure 8-5) is an approximately 10-percent increase in the COE, to 73.9 mills/kWh. Lower carbon conversion in the gasifier accounts for about half the total. This margin can be viewed as an upper bound on the COE as affected by plant efficiency.

8.2.1.5 Operating and Maintenance Costs (Nonfuel)

Nonfuel operating and maintenance costs are subject to considerable uncertainties because they cover both labor costs and a wide variety of material costs. An increase of 25 percent in nonfuel O&M costs raises the COE by 2.9 mills/kWh (2.3 mills on a constant dollar basis), or 4.3 percent of the base value of 67.4 mills/kWh.

Solid waste disposal costs for ash and sulfur sorbent are one area of O&M costs with great potential risk for cost increases in the future. The base case estimate includes approximately \$3 per ton for disposal costs. If this should increase to \$15 per ton, the result would be an increase of 1.7 mills per kWh in the COE, to 69 mills per kWh (50.4 mills on a constant dollar basis). In other words, an increase in disposal costs by a factor of five would increase COE by only 2.5 percent.

8.2.1.6 Relative Risks

With the simplified approach to sensitivity studies used here, most economic sensitivity relationships are linear. As such, they can be completely represented by the slope of the line. Given the slope of the line and an estimate of the maximum probable range of the independent variable, it is possible to illustrate graphically the relative risks due to the different categories of uncertainty discussed in preceding sections. As seen in Figure 8-7, capacity factor is clearly the most important variable. It is influenced by design reliability and by system dispatching characteristics.

After capacity factor, it appears that roughly equal risks are due to uncertainties in capital cost and in the combined effect of thermal efficiency factors (i.e., carbon conversion, cold gas efficiency, auxiliary power consumption, stack gas temperature). Coal cost and other operating and maintenance costs represent a relatively lesser magnitude of risk.

8.2.2 Comparative Sensitivities

Sensitivity studies for the COE were also carried out for Cases 3, 5, and 6 in addition to Case 1. These results were combined to compare the sensitivity variables among the different case designs. When different systems are compared, however, the sensitivities must be shown as absolute values (i.e., as changes in the

Figure 8-7 Relative Risks - Case 1

COE, in mills per kilowatt-hour). Sensitivity analyses were not done for the ISTIG cases (2 and 4) due to their more conceptual nature and lack of optimal performance.

8.2.2.1 Capital Costs

Figure 8-8 illustrates the relative sensitivity to a 20-percent increase in capital cost for the four cases compared. Case 3, at half the size of Case 1, shows a larger effect simply because of the scale factor, which makes electricity from small plants more expensive than from larger plants. Because Case 3 is identical in design concept to Case 1, the COE sensitivities vary in direct proportion to the differences in base case values. Thus, Case 3 has a 14-percent higher capital cost per kilowatt and shows a 14-percent greater change in cost per kilowatt-hour for a given percentage change in capital cost.

Case 5, with the added cost of a sulfuric acid plant, has the highest capital investment per unit of output and shows the greatest effect. A 20-percent increase in capital costs would raise the COE by 8.3 mills/kWh. Starting from a higher base value means that any sensitivity (that is, response to a given percent change in the independent variable) is also likely to be higher for Case 5. Essentially because it has the lowest capital investment, Case 6 also shows the lowest sensitivity to a change in capital costs.

Although the capital sensitivities shown in Figure 8-8 are ordered in the same sequence as the capital costs themselves, that is not a preordained result. The changes in balance between capital costs and other costs for the four cases are simply not enough to outweigh the influence of capital cost itself. The relative sensitivities cannot be predicted just on the basis of relative capital costs. For instance there is a 17-percent difference in capital cost between Cases 1 and 6, but only a 3-percent change in the sensitivity. There is only a 10-percent difference in capital cost between Cases 1 and 5, but an 18-percent difference in sensitivity.

The relative rankings shown in Figure 8-8 are based on the same percent change in capital cost for each case. However, there would not be the same probability of this occurring for each case. Although Case 5 has the highest capital cost, a major reason for this is the cost of the sulfuric acid plant. This is standard, off-the-shelf technology with relatively low capital risk involved. On the other hand, Case 6 has the lowest capital cost because of the unproved assumption that the Kellogg transport gasifier will perform as expected and will have a high enough in-bed sulfur removal efficiency that a sulfur polishing step will not be needed. At this time neither the zinc ferrite sulfur removal step nor the transport gasifier has been tested on a sufficient scale, and it is impossible to assign a relative probability of exceeding the estimated cost of construction. An economic risk analysis would be desirable to estimate the probabilities.

Figure 8-8 Comparative Sensitivity to Capital Costs (For 20% Increase)

8.2.2.2 Cost of Gasification Island

The gasification island contains the process units with the most potential for exceeding estimated construction costs. The units included are:

- Coal receiving, handling, and preparation
- Limestone receiving and handling (where required)
- Booster compression
- Gasification
- Recycle gas compression
- Gas conditioning
- External desulfurization (where required)
- Sulfation (where required)

As seen in Figure 8-9, the ranking of the four cases for sensitivity to gasification island costs is the same as that for overall capital cost. However, Case 6 is the most attractive by a wider margin than before and appears to offer a significant advantage here.

8.2.2.3 Sensitivity to Carbon Conversion in the Gasifier

Because of basic differences in the flow sheets, the effect of low carbon conversion in the gasifier varies substantially between cases (Figure 8-10).

Case 6 shows by far the most desirable behavior. Any carbon unconverted in the gasifier is burned in the pressurized sulfator which exhausts to the gasifier outlet. Thus any carbon carryover from the gasifier is burned and the hot gas run through both the combustion turbine and the heat recovery steam generator. This generates approximately 89 percent as much electricity as if the carbon had been converted in the gasifier. Therefore Case 6 is not very sensitive to poor carbon conversion in the gasifier. This reduction in risk could be an important advantage of the hybrid technology in which both IGCC and PFBC operations are integrated.

At the other extreme, Case 5 has no provision for recovering energy from unconverted char, and the cost of electricity suffers accordingly. In a middle position are Cases 1 and 3, which recover some energy from the char, but less than in Case 6.

Figure 8-9 Comparative Sensitivity to Gasification Island Costs (For 50% Increase)

Figure 8-10 Comparative Sensitivity to Carbon Conversion in Gasifier

8.2.2.4 Sensitivity to Coal Cost

Sensitivity to coal cost depends only on the heat rate. A 25-percent increase in the cost of coal increases the COE by 4.2 mills for Cases 1 and 3, by 4.6 mills for Case 5, and by 4.0 mills for Case 6. Thus there is relatively little difference from one case to another.

8.2.2.5 Sensitivity to Capacity Factor

Other things being equal, the plant with the lowest fixed costs should show the lowest effect of operating at reduced capacity factor. Thus Case 6 sees the smallest increase in COE when capacity factor drops off, as shown in Figure 8-11. At 85-percent capacity factor, there is a difference of 14.4 mills between Cases 5 and 6, which widens to a difference of 22.4 mills at a capacity factor of 45 percent.

The Case 6 cost advantage over Case 1 is equivalent to a 7-percent increase in capacity factor. That is, Case 6 operating at 65-percent capacity factor is equivalent to Case 1 operating at over 72-percent capacity factor.

8.2.2.6 Auxiliary Power Consumption

The comparative sensitivity to increased in-plant power consumption showed that Cases 1 and 6 were equal, and Case 5 was again the worst. However, the total spread among all cases is only 0.6 mills, indicating that this is not a parameter of critical concern.

8.2.2.7 Sensitivity to Operating and Maintenance Costs

Sensitivity to nonfuel operating and maintenance costs is shown in Figure 8-12. The ranking of the four cases is the same as in the other charts, with Case 6 being the best and Case 5 the worst. The difference between these two is of about the same magnitude as was seen for capital cost (Figure 8-8). This indicates that attempts should be made to achieve the same degree of accuracy in estimates of operating and maintenance costs as is achieved in estimates of capital cost.

With regard to Case 5, byproduct credits from the sale of sulfuric acid are insufficient to offset the effect of higher capital costs. Sulfuric acid prices may fluctuate over a wide range, affecting byproduct credits. However varying sulfuric acid price over a total range of -30 to +30 percent results in a change of only one mill in the COE. Although operating costs receive the benefits of byproduct credits from the sale of sulfuric acid and the elimination of the need to buy limestone, net nonfuel operating and maintenance costs are still higher than for any other case. This is due to the extra cost of operating the sulfuric acid plant, the oxygen plant, and the larger zinc ferrite hot gas cleanup section.

Figure 8-11 Comparative Sensitivity to Capacity Factor - IGCC Cases

Figure 8-12 Comparative Sensitivity to Non-Fuel Operating And Maintenance Costs

8.2.2.8 Sensitivity to NO_x Emission Controls

A key uncertainty concerns future emission control regulations for nitrogen oxides. The base configuration for the economic analyses was no control. The estimated NO_x emissions from the heat recovery steam generator stack are shown in Table 4-22. Sensitivity cases were run with NO_x limits assumed to be as low as 25 ppm.

Nitrogen oxide emissions in Case 5 are much higher because no limestone is used in the gasifier bed. Case 6 emissions are higher than for Case 3 because of the lower operating temperature in the transport gasifier.

Costs were estimated for achieving the target limits by adding a selective catalytic reduction (SCR) process to the power plant, although less expensive processes may be available. It is unknown at this point whether the SCR catalyst would be fully compatible with coal-derived gases. At a control level of 75 ppm NO_x, the current new source performance standard for natural-gas-fired combustion turbines, only Case 5 would require controls. An SCR unit for 80-percent control is estimated to cost \$9.8 million with operating and maintenance costs of \$635,000 per year. The effect on COE is shown in Table 8-3, along with the effect on all cases if a 25-ppm standard were imposed. It must be pointed out, however, that these results could be subject to revision when and if commercial experience with SCR on IGCC plants is obtained.

8.2.2.9 Summary of Sensitivity Comparisons

Results of the comparative sensitivity studies suggest the following major conclusions:

- For an arbitrary percent change in any of the cost variables studied, Case 6 suffers the least impact on COE.
- Changes in coal cost or coal escalation rates affect all cases approximately the same. Sensitivity to these parameters is low.
- Capacity factor is the most important single cost parameter. A change of 7 percent (25 days per year) in availability would switch the economic ranking of Cases 1 and 6.
- Unexpectedly low carbon conversion in the gasifier would have dramatically different effects on the different designs. Case 5 suffers a significant cost increase but Case 6 is little affected.

Table 8-3

Effect of Imposing Nitrogen Oxide Emission Limits

	Increase in Cost of Electricity	
	<u>Current Mills</u>	<u>Constant Mills</u>
<u>75 ppm Limit</u>		
Case 1	0.0	0.0
Case 5	1.1	0.9
Case 6	0.0	0.0
<u>25 ppm Limit</u>		
Case 1	0.5	0.4
Case 5	1.5	1.1
Case 6	0.7	0.5

- Although capital cost estimates receive by far the most attention, uncertainty in nonfuel operating and maintenance costs also show an appreciable effect. More effort to refine these estimates could be worthwhile.
- Effects of higher disposal costs, higher auxiliary power usage, higher stack temperature, and higher limestone use do not show appreciable differences between the four cases.
- Based on current estimates, imposition of strict nitrogen oxide controls would increase the COE by less than 1 mill per kWh in all cases except Case 5. However, SCR has not been demonstrated commercially for IGCC, and therefore this conclusion could be significantly different if, for example, catalyst life is shorter than for natural gas applications.

8.3 SYSTEM PRODUCTION COST ANALYSIS FOR CASE 1

The foregoing analyses were based on COE as determined by standard leveled busbar cost calculations for single plants using an assumed capacity factor of 65 percent. However, in order to fully understand the potential economic performance of IGCC in The Southern Company, it is necessary to conduct a production cost analysis which fully accounts for The Southern Company characteristics. These characteristics include patterns of electric load or demand, future load growth projections, the current and projected generation mix, the performance and economic characteristics of individual plants, current and future financial parameters, and other system data.

To evaluate air-blown IGCC with hot gas cleanup for The Southern Company, a detailed system production cost analysis was conducted for IGCC based on Case 1 versus the current technology choice for new baseload generation, pulverized coal combustion with flue gas desulfurization (FGD). The technical approach and results for the system production cost analysis are described more in Section 8.3.

8.3.1 Technical Approach

An economic analysis of a Case 1 type IGCC plant was performed for the scenario in which an 811-MW pulverized coal-fired unit is replaced with 811 MW of IGCC capacity. This scenario assumes The Southern Company's 1990 base case economic projections and expansion plans (Table 3-2), which call for installation of an 811-MW pulverized coal-fired unit for baseload power generation with an in-service date of 2006. The year 2006 is the first year the expansion plan indicates a need for new baseload coal-fired capacity.

Although the detailed IGCC design and cost estimates developed in this study were based on a nominal 458-MW unit, there is an economy of scale for building larger IGCC plants. The capital cost of the IGCC plant was scaled to an 811-MW size to permit a consistent comparison with the pulverized coal-fired unit. A reduction of 11.9 percent was applied to the capital cost of the IGCC plant on a dollars-per-kilowatt basis to account for the economy of scale based on Electric Power Research Institute (EPRI) recommendations (Reference 9). The capital costs for both plants were based on installation at Plant Wansley. Reductions in the costs of the pulverized-coal-fired and IGCC plants were assumed where existing equipment and facilities at Plant Wansley could be used.

It was also necessary to develop scaling factors for the fixed O&M portion of the IGCC operating costs. As described in Section 7.0, the labor portion of the fixed O&M costs for the IGCC cases was calculated based on plant O&M staffing plans developed as part of the study. Specific plant staffing requirements (e.g., the number of O&M personnel per megawatt output) are known to decrease as plant size increases. In order to scale fixed O&M costs for Case 1 from a 458-MW installation to an 810-MW IGCC plant, O&M cost scaling parameters were developed from data available for The Southern Company coal-fired plants of various sizes. These factors were used to predict fixed O&M costs for an 810-MW IGCC unit. Variable O&M costs are expressed relative to kilowatt-hours of output and were not assumed to be affected by plant size.

It must be pointed out that both capital and operating costs for the pulverized coal plant used in the system analysis were not developed in the current study but represent costs used by system planners for internal system studies. Moreover, the adjustment of pulverized coal costs to a Plant Wansley basis must be considered approximate since a rigorous approach could not be undertaken during the timeframe of the current study. Therefore, the results of the analysis should be interpreted as indicative rather than absolute. Future efforts will be made to refine the pulverized coal plant costs.

8.3.2 System Expansion Plan Scenario

The projected expansion plan assumed for this study was presented previously in Table 3-2. Generating units presently committed are expected to supply adequate capacity through 1993. The next new capacity in the system is projected to be peaking capacity, followed by intermediate and baseload capacity. The 1990 base expansion plan shows combustion turbines being added as peaking capacity beginning in 1994, combined-cycle units as intermediate capacity beginning in 2004, and coal-fired units as baseload capacity beginning in 2006.

8.3.3 Financial Assumptions

The financial assumptions shown in Table 8-4 were used for the analysis of capital costs. These assumptions are generally consistent with those used for COE calculations (Table 8-1) with a few exceptions, such as the assumed plant life and the treatment of insurance and local taxes. Whereas insurance and local taxes are treated as an O&M cost in COE calculations, the system production cost analysis treats them as capital charges, which is consistent with The Southern Company accounting practices.

8.3.4 Plant Performance and Cost Assumptions

A summary of performance and cost assumptions for both the IGCC plant and the pulverized-coal-fired unit is shown in Table 8-5.

Production costs and operating characteristics were estimated using PROMOD III. PROMOD III is a computer software package developed by Energy Management Associates for simulating operation of electric utility generation systems. Input required for PROMOD includes forecast monthly energies and demands, unit forced outage rate projections, unit maintenance outage schedules, unit fuel cost rates, and operating and maintenance expenditure rates. Results from PROMOD include projected generation, fuel requirements, fuel cost, operation and maintenance cost, and production cost. Results are reported on a unit, plant, company, and system basis.

PROMOD was run for the period 2006 to 2010 in the Plant Wansley study. Detailed system modeling is not conducted beyond 2010, therefore the results were extended from 2010 to the end of the study period by assuming that production cost would increase at a rate of 4.5 percent per year, which is the projection for the Gross National Product Implicit Price Deflator.

The relative effects on revenue requirements of capital cost differences for the IGCC and pulverized coal plants are routinely evaluated by SCS using an in-house software application called PROVAL. Plant capital costs shown in Table 8-5 were input to PROVAL along with the financial assumptions shown in Table 8-4. The AFUDC output from PROVAL was added to the total plant capital cost to yield the in-service book value shown in Table 8-5.

Table 8-4

**Financial Assumptions for System
Production Cost Analysis**

<u>Source of Capital</u>	<u>Capitalization Ratio (%)</u>	<u>Cost (%)</u>
Debt	45	10.30
Preferred	10	9.80
Common	45	13.75
Composite	100	11.803
Discount rate (%)		10.0
AFUDC (%)	11.803	
Income taxes (%)		
State	5.66	
Federal	34.00	
Composite income tax rate	37.74	
Other taxes (%)	0.60	
<u>Plant Life</u>	<u>Life (years)</u>	
Useful life	45	
Book life	45	
Tax life	20	

Table 8-5

**Plant Performance and Cost Assumptions
for System Production Cost Analysis**

	<u>Pulverized Coal</u>	<u>Case 1 Air-Blown IGCC</u>
Net capacity (MW) ^a	811	811
Overnight plant capital cost (1990 \$) ^b	966,577,920	812,902,000
In-service book value including AFUDC (2006 \$) ^c	2,335,960,000	1,840,025,000
Fixed O&M expenses (1990 \$/kW-yr)	18.76	24.67
Variable O&M expenses (1990 \$, mills/kWh)	1.63	2.41
Full-load net heat rate (Btu/kWh, HHV basis) ^a	9581	8595
Equivalent forced outage rate (%)	6.0	11.43
Average planned maintenance outage (days)	28	17

a. At 90°F design point

b. Consists of total capital requirement less AFUDC

c. In-service date is 2006

8.3.5 Production Simulation Results

A summary of the PROMOD III simulation results for the IGCC plant for the year 2010 is shown in Table 8-6. By 2010 the Southern electric system will be operating at optimum risk and the results for 2010 are representative of the results over the life of the IGCC plant. The production simulation results from PROMOD indicate that the IGCC plant would operate as a baseload plant in the Southern electric system. In 2010 the IGCC plant is projected to run at an 83.6-percent capacity factor, and is projected to operate at maximum output during the times when the combined-cycle units and gasifier trains are available.

The time at maximum output is projected to be 6,469 hours. It is projected to operate an additional 1,781 hours at derated load due to partial forced outages. The IGCC plant is projected to be completely offline for 102 hours due to full-forced outage and 408 hours due to planned outage, for a total of 510 hours offline. The IGCC unit would not be equipped with backup fuel capability, therefore the combined cycle is not operational when the gasifier is unavailable. Due to concerns about bypass damper leakage, the design did not include a bypass duct for operating when the steam turbine was forced out. Therefore, there would be no operation in the simple-cycle mode. The average heat rate of the IGCC unit over all operating conditions is projected to be 8,621 Btu/kWh.

8.3.6 Projected Revenue Requirements

The economic competitiveness of installing an IGCC plant relative to a pulverized coal-fired plant was compared in the economic analysis. Capital costs and production costs were included in the comparison. Differences in projected annual revenue requirements for capital additions were determined by comparing the revenue requirements of the 811-MW pulverized coal-fired unit to the revenue requirements of 811-MW of IGCC capacity. A similar comparison was conducted for production costs. The sum of differences in capital and production costs for the two plants represents the total cost impact on the system due to the installation of the IGCC plant in place of the pulverized coal plant currently planned.

The projected capital revenue requirements are shown over the assumed plant life in Figure 8-13. Figure 8-13 shows cumulative revenue requirements expressed in mid-1990 present worth dollars. Present worth values were calculated using The Southern Company's projected after-tax discount rate of 10 percent. The results indicate that the IGCC plant would have lower capital revenue requirements by an estimated \$158 million in present-worth mid-year 1990 dollars.

Table 8-6

**Pulverized Coal Versus IGCC
PROMOD III Operating Results for Year 2010
(Typical year at optimum system generation mix)**

	<u>Pulverized Coal</u>	<u>IGCC</u>
Annual Service Hours	7602	8250
Availability Factor (%)*	86.8	94.2
Capacity Factor (%)	79.9	83.6
Average Heat Rate (Btu/kWh)	9608	8621

*** Defined as annual service hours divided by 8760 hours.**

Figure 8-13 Capital Revenue Requirements by Year in Cumulative Nominal Mid-Year Dollars
Expressed in Mid-1990 Present Worth

The difference in total production costs for the pulverized coal and IGCC plants was estimated using PROMOD III, as explained previously. Total production costs include fuel expenses and operation and maintenance expenses. The sum of the differences in production and capital costs of the two plants represents the difference in cost for a Case 1 IGCC plant compared to a coal-fired plant. Figure 8-14 shows the estimated differences in production cost, capital cost, and the sum of production cost and capital cost between the pulverized-coal-fired and IGCC plants. A positive difference indicates a savings for the IGCC plant. The IGCC plant has lower production costs than the coal-fired unit, primarily due to its lower heat rate as compared to the pulverized coal-fired plant. The production costs for the IGCC plant are \$463 million lower in cumulative nominal mid-year dollars present-worth to mid-1990.

When the production cost savings are added to the capital cost savings of \$158 million, the IGCC plant is projected to achieve a total savings of \$621 million in cumulative revenue requirements expressed as mid-1990 present-worth dollars, as shown in Figure 8-14. This is a substantial sum relative to the capital investment in the plant and indicates that the air-blown IGCC design with hot gas cleanup should be economically very competitive with the pulverized coal plant, if developed to full commercial potential.

8.3.7 Revenue Requirement Sensitivity to O&M Cost Assumptions

There is no commercial experience with the operation of a full-scale IGCC unit in the United States, particularly air-blown IGCC with hot gas cleanup. The fixed O&M costs for such a unit, including both labor and material requirements, are thus subject to some uncertainty. To address this uncertainty, two sensitivity cases to IGCC fixed O&M costs were conducted as part of the system production cost analysis. These cases involved: (1) assuming no difference in fixed O&M cost between the IGCC and the pulverized coal unit; and (2) assuming that the difference in fixed O&M is double the value estimated.

Figure 8-15 shows the impact of reducing the fixed O&M difference between the coal unit and the IGCC unit to zero. In this case, the mid-year 1990 present-worth production costs for the IGCC unit are \$501 million lower than those of coal-fired unit, and the total present worth cost savings is \$659 million.

In Figure 8-16, the difference in fixed O&M is doubled to \$11.82/kW-yr higher for the IGCC plant, resulting in mid-year 1990 present-worth production costs \$426 million lower than those of the coal-fired unit. The total present-worth cost savings for the IGCC plant is thus reduced from \$621 million (base case) to \$584 million.

Figure 8-14 Differences in Revenue Requirements (Coal Minus IGCC) by Year in Cumulative Nominal Mid-Year Dollars Expressed in Mid-1990 Present Worth

**Figure 8-15 Differences in Revenue Requirements (Coal Minus IGCC) by Year in Cumulative Nominal Mid-Year Dollars Expressed in Mid-1990 Present-Worth - Sensitivity Case:
Fixed O&M Cost Difference Twice as High for IGCC**

Figure 8-16 Differences in Revenue Requirements (Coal Minus IGCC) by Year in Cumulative Nominal Mid-Year Dollars Expressed in Mid-1990 Present-Worth - Sensitivity Case:
No Difference in PC and IGCC Fixed O&M

Based on these analyses, a significant increase in fixed O&M costs assumed for the Case-1-type IGCC plant would not change the conclusion that the IGCC plant has excellent potential to achieve cost savings over the pulverized-coal option. Again, this conclusion is subject to continuing engineering analysis during coming years.

Another O&M-related sensitivity parameter which was not investigated in the system analysis was the length of the planned outage for the IGCC plant. An aggressive schedule of 17 days per year has been assumed for the Case-1 IGCC, as shown in Table 8-5. The reasoning behind this assumption was discussed in Section 4. The assumption reflects the belief on the part of many experts that highly modular technologies should have inherently lower maintenance requirements than large-scale, non-modular lower generation systems.

The fixed O&M cost sensitivity analyses suggest that the basic conclusion of the system production cost analysis would not change over a reasonable range of uncertainty in the relative O&M costs. The large savings in fuel costs for the IGCC unit maintain a substantial advantage which overcomes increases in the other production costs.

8.3.8 Summary of System Production Cost Analysis

Section 8.3 has described a comparative analysis of air-blown KRW IGCC with hot gas cleanup versus a pulverized coal unit with FGD. This analysis suggests that advanced IGCC systems which take advantage of capital cost savings and efficiency improvements resulting from process simplification can potentially be extremely competitive with the pulverized coal option. In addition, the current study indicates that, even with hot gas cleanup, IGCC is environmentally superior to pulverized coal units.

On the other hand, the IGCC technologies described in this study have not yet been demonstrated at commercial scale. In order to realize the potential benefits of IGCC, further development, culminating in commercial demonstration, is required.

The system production analysis was conducted for a Case 1 type plant. A similar analysis for a Case 6 type plant would show even greater benefits for IGCC. However, the Case 6 technology is correspondingly further from commercialization than Case 1.

9.0 ENVIRONMENTAL HEALTH AND SAFETY AND SOCIOECONOMIC ASSESSMENT

All the IGCC and ISTIG plants considered in this study were designed to efficiently reduce all air emissions and effluents to regulatory limits or below, and to produce a solid waste suitable for landfill onsite. This section discusses the environmental effects of the IGCC and ISTIG plants, along with the safety and health effects, and the socioeconomic effects of construction of the plants.

9.1 ENVIRONMENTAL PERMITTING PROCESS AND COMPLIANCE PLAN

The environmental permits for operating the plant will be applied for 2 years prior to startup of the plant. Environmental permits will be required for air emissions, water discharge, and solid waste disposal. Each of these items is discussed in further detail in the following sections along with the key design features to ensure environmental compliance for the plant.

9.1.1 Air Emissions

Air emissions are primarily from three sources: fines lockhopper vent gas, flue gas from the HRSGs, and flue gas from the sulfator. The lockhopper vent gas contains fine particulate from the fines removal lockhopper. The flue gas from the HRSG contains SO_2 and NO_x . The hot particulate removal step removes virtually all the particulate from the gas stream that goes to the HRSG. The flue gas from the sulfator contains SO_2 , NO_x , and particulates produced during the oxidation of the sulfides and unburned carbon in the spent LASH.

9.1.1.1 Emissions Control

Vent Gas

The vent gas from the fines-removal lockhopper is filtered in order to meet particulate emission standards prior to being flared. Carbon monoxide in trace amounts is the only pollutant that remains in this gas after flaring. However, the stream flow of vent gas is very small. With the small flow and low concentration of carbon monoxide, the ambient air quality standard concentrations will be below NAAQS.

HRSG and Sulfator Flue Gas

SO_2 emissions are controlled by in-bed desulfurization in the gasifier and by removal of most of the remaining sulfur by a zinc ferrite sorbent, except for Case 5. In Case 5, zinc ferrite is used exclusively for desulfurization. Sulfur

removal efficiencies greater than 99 percent are achieved in all cases for the fuel gas to the combustion turbine.

Pilot plant measurements by MWK indicate that NH_3 in the fuel gas results in NO_x emissions ranging from 0.05 to 0.2 pound NO_x per MMBtu when limestone is used for in-bed desulfurization and 0.6 pound NO_x per MMBtu without limestone. However, SCR was needed in Case 5 (no limestone) to reduce the NO_x emission to a level below the NSPS for combustion turbines (75 ppm) because of the additional contribution of thermal NO_x from the combustion turbine.

Near absolute particulate removal is achieved in the fuel gas by the use of a cyclone and a ceramic candle gas filter. Pilot plant data from MWK indicate that particulate concentrations in the fuel gas are well below 10 ppm.

SO_2 and NO_x emissions from the sulfator are controlled by limiting the temperature in the sulfator. Sulfur released from the char in the sulfator is captured by the limestone and converted to calcium sulfate. In Cases 1 through 4, the sulfator flue gas is used for coal drying. The flue gas from the drying step is filtered in order to meet particulate emission standards prior to release.

9.1.1.2 Modeling

The Industrial Source Complex Short Term (ISCST) model was used to help determine the impact of the gasification-based power plant on air quality for all the cases. The maximum 1-hour concentrations for SO_2 , NO_x , and particulate emissions were estimated. Then 3-hour and 24-hour concentrations were determined by factoring the 1-hour concentrations as recommended by the EPA.

Initially, the model was run with "worst case" meteorological conditions to estimate the maximum 1-hour concentrations for SO_2 , NO_x , and particulate matter of size 10 micrometers or smaller for Cases 1, 5, and 6. For the cases without a separate discharge point for sulfator flue gas (Cases 5 and 6), the model indicates that short-term impacts are below the standards for SO_2 , NO_x , and particulates; therefore, it is very unlikely that National Ambient Air Quality Standards would be exceeded. However, for Case 1, the model indicates that short-term impacts of NO_x and particulates are less than the significant impact levels, but SO_2 exceeds the Prevention of Significant Deterioration (PSD) significant impact level for monitoring. A concentration of SO_2 higher than the significance level for Case 1 is due primarily to the emission from the sulfator, although the exceedance is not caused by a higher emission rate. In Cases 1 through 4 the sulfator flue gas is released from a separate stack and the flow from this stack is much less than the flow from the stacks at the HRSG. The plume from the sulfator stack cools rapidly due to the small flow. The rapid cooling results in a relatively small plume rise and consequently results in less dispersion of the gas.

Because of the exceedance for Case 1, the model was run for all cases using meteorological data from the region near Plant Wansley instead of the "worst case" data. The impacts predicted by the model for SO₂, NO_x, and particulate matter of size 10 micrometers or smaller, are listed in Table 9-1. The impacts for NO_x and particulates are below the monitoring significant impact level for all six cases. The monitoring significant impact level for SO₂ was exceeded in Cases 1, 3, and 4, but only slightly. Under PSD regulations, exceedance of the monitoring significant impact level may require preconstruction SO₂ monitoring for a period of 1 year at the plant site. This monitoring serves to determine the existing air quality at the plant site. Although the monitoring significant impact level for SO₂ may be exceeded, the actual concentration of SO₂ should be far below the NAAQS.

Since the exceedance of the SO₂ monitoring level is caused by the small flow from the separate stack, SCS would consider rearranging the plant or making other design changes to allow the sulfation gas to share a common stack with the flue gas from the HRSG. Modeling indicates that the significance level would not be exceeded if a common stack was used.

9.1.2 Solid Disposal

The solid wastes, including spent limestone, ash, spent zinc ferrite sorbent, and nahcolite sorbent, are disposed of in a dry landfill at the plant site. All of these wastes are nonhazardous. Various leaching studies of solid wastes from the KRW gasifier at Waltz Mill, Pennsylvania, were conducted using materials produced in the pilot plant with and without in-bed desulfurization. Also, samples of LASH material from the pilot plant in-bed desulfurization tests, both before and after sulfation treatment were tested. In every case the EP-toxicity test results of LASH showed compliance with the Resource Conservation and Recovery Act (RCRA) of 1976 requirements (Reference 10). Sulfation tests were conducted for MWK by Dorr-Oliver to determine the extent to which sulfides in the LASH could be converted to sulfate. Material from the Dorr-Oliver sulfation tests was tested by EP-toxicity tests with the results noted in Table 9-2.

The landfill for solids disposal is lined with a 3-ft thick clay liner and all leachate and surface runoff from the landfill is routed to a sedimentation pond (also lined with clay) for removal of entrained solids prior to discharge at an existing water discharge point. Four ground-water monitoring wells are located around the landfill. When ash is stacked to the maximum design height for the landfill, the material will be covered with a 2-ft layer of soil, and grass will be planted. Figure 9-1 shows the location of the landfill for Case 1, and a description of landfill operation can be found in Subsection 4.2.1.11.

Table 9-1

ISCST Model Predicted Impacts of IGCC Plants on Air Quality

	<u>3-hr</u>	<u>SO₂</u> <u>24-hr</u>	<u>Annual</u>	<u>PM-10</u> <u>24-hr</u>	<u>NO_x</u> <u>Annual</u>
Case 1 Impact	85	23	3.0	0.1	3.0
Case 2 Impact	44.8	10.9	3.0	0.1	3.0
Case 3 Impact	56.7	16	2.1	0.4	1.2
Case 4 Impact	85.9	23.1	3.0	4.6	3.0
Case 5 Impact	11.5	3.0	0.4	0.6	3.0
Case 6 Impact	39.3	10.4	1.2	0.5	0.6
Monitoring Sig. Impact Level	—	13	—	10	14
PSD Increments	512	91	20	30	25
NAAQS	1,300	365	80	150	100

Concentrations in micrograms/cubic meter

Predicted impacts are "highest-second-high" for 3- and 24-hour averaging times.

Table 9-2

**EP-Toxicity Test Results for Material from
the Dorr-Oliver Sulfation Treatment Test
(All quantities in mg/L)**

<u>Chemical Constituent</u>	<u>Maximum Allowable</u>	<u>Roasted LASH</u>	<u>Ratio</u>
Ag	5.0	0.0444	114
As	5.0	0.004	1250
Ba	100.0	0.530	1250
Cd	1.0	0.005	189
Cr	5.0	0.058	86
Hg	0.2	0.001	153
Pb	5.0	0.070	71
Se	1.0	0.003	333

9.1.3 Water Discharges

Most of the water streams to be discharged from the gasification-based plant are the same as streams in conventional power plants. Water streams treated by conventional techniques include coal pile runoff and structure runoff. Runoff from the limestone storage area is routed to an existing ash pond along with the coal pile runoff to allow sufficient time for solids to settle prior to discharge.

In Case 5, a weak acid stream from the sulfuric acid plant must be neutralized. A NaOH solution is used to neutralize the stream prior to discharge.

In all cases, leachate and surface runoff are collected from the ash landfill and routed to a sedimentation pond. The sedimentation pond and existing sedimentation ponds at the site have sufficient capacity to contain the 24-hour, 10-year rainfall runoff from the landfill. This scheme allows all solids to settle prior to discharge at an existing point. No further treatment is anticipated, but by collecting the water, additional treatment, such as pH adjustment, would be possible if it became necessary.

9.1.4 Minor Constituents

Environmental test programs were carried out in the KRW pilot plant to determine the fate of the minor constituents such as alkali and heavy metals, halides, and hydrocarbons that enter the gasifier in coal and limestone feed. Some of these minor constituents, such as alkali metals and halides, can be tolerated only at very low levels in the fuel gas to the gas turbine. A significant amount of analytical work related to the fate of these minor constituents was completed. The Radian Corporation performed two environmental sampling campaigns at the pilot plant. A 1983 campaign involved oxygen-blown gasifier operation with Pittsburgh No. 8, Wyoming, and North Dakota lignite coals (Reference 3). The second campaign was performed during a 1988 proprietary test campaign involving air-blown gasifier operation with and without in-bed desulfurization. Because the gasifier conditions do not vary significantly among the operating modes and the bituminous feedstocks, it is reasonable to draw general conclusions about the minor constituents as applicable to this study.

The minor constituents have been grouped together in five broad categories as described below:

Chloride

During the 1988 campaign, it was established that only about 15 percent of the coal-borne chloride is contained in the fuel gas. The bulk of the remaining chloride is contained in the fines that escape the KRW gasifier. Conservatively, it can be assumed that no more than 15 percent of coal-borne chloride will be contained in the fuel gas from the transport gasifier.

Alkali Metals

During the 1988 campaign, the sodium content of the fuel gas was found to be less than 0.1 ppm, which is below analytical equipment detection limits. The potassium level in the coal gas ranged from 6 to 12 ppb and is also below analytical equipment detection limits. During the 1983 campaign, it was determined that less than 0.1 weight percent of coal-borne sodium and potassium was contained in the fuel gas.

Trace Elements

During the 1988 campaign, calcium and vanadium levels in the fuel gas were found to be in the ranges of 10 to 20 ppb and 2 to 3 ppb, respectively. There appeared to be no increase in calcium content of the fuel gas when comparing similar operations with and without limestone addition to the gasifier bed.

N

ZONED GROWTH
FOR GRASSING

WASTE
AREA

(ORIGINAL 60x90)

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

**Figure 9-1 Solid Waste
Landfill Location for Case 1**

Several other trace elements were monitored during one of the 1983 campaigns. These are Ag, Al, As, B, Be, Cd, Co, Cr, Cu, Fe, Hg, Li, Mg, Mn, Mo, Ni, P, Pb, Sb, Se, Si, Sr, Ti, and Zn. For bituminous coals, less than 1 percent of the feed for all the foregoing elements, with the exception of mercury, was contained in the fuel gas. About 31 percent of mercury in coal was found in the fuel gas with the remainder being retained in the solids.

Volatile Organics

The benzene content of bituminous-coal-derived raw fuel gas averaged about 150 ppm. The product gas was also analyzed for toluene and xylene. Both toluene and xylene concentrations were below detectable limits during the bituminous coal tests (1983 campaign). Due to the lower operating temperature of the transport gasifier, the volatile organic content of the fuel gas to the gas turbine is projected to range from 3500 to 4000 ppm.

High-Molecular-Weight Organics

This category is for organic compounds with molecular weight greater than 128. Raw product gas from the gasifier contained about 18 ppm by weight in the coal feed) of the high-molecular-weight organics. In the pilot plant tests, raw gas containing particulates was scrubbed with water, resulting in about 90-percent removal of high molecular weight organics in water and particulate sludge (1983 campaign).

9.2 HEALTH AND SAFETY

The health and safety requirements applicable to operation of the proposed gasification power plant include the construction and general industry standards of the federal Occupational Health and Safety Administration in 29 CFR Parts 1910 and 1926, respectively. These standards include walking-working surfaces, means of ingress and egress, operation of powered equipment, adequate ventilation, noise exposure controls, fire protection, and electrical equipment safeguards. It is believed that current procedures are adequate to ensure that federal and state standards are met.

9.3 SOCIOECONOMIC EFFECTS OF CONSTRUCTION

The Wansley site is located in Heard and Carroll Counties at the intersection of the Heard, Carroll, and Coweta county lines. The labor-pool geographical source for construction workers and plant operations personnel is expected to be Carroll and/or Coweta counties. The required workers are expected to be available from the local area and no adverse effects to area public services are anticipated.

10.0 IGCC SYSTEM COMPARISONS AND SECOND LAW ANALYSIS

10.1 COMPARISON OF AIR-BLOWN IGCC WITH HOT GAS CLEANUP VERSUS OXYGEN-BLOWN IGCC WITH COLD GAS CLEANUP

An air-blown, hot-gas-cleanup IGCC system differs in many respects from the more conventional oxygen-blown gasifier, cold-gas cleanup system. The three areas of most interest for comparison are:

- Reliability and operability
- Environmental effects
- Economics

The two approaches were compared in these three areas based on the air-blown IGCC with hot gas cleanup of Case 1 and the oxygen-blown IGCC with cold gas cleanup from the EPRI site-specific study, referred to as the reference case in the following discussion. A brief description of each case is included in Table 10-1.

10.1.1 Reliability and Operability

10.1.1.1 Gasifiers

It is assumed that there is basically little difference in the reliability or operability of the gasifiers themselves. Although the oxygen-blown case has the added complexity of an oxygen plant, surge storage of liquid oxygen removes almost all reliability problems associated with this part of the facility. The air-blown system requires more gasifiers to handle the greater volume of gas associated with the use of air, although this effect is partially offset by the higher reactivity of the coal/limestone mixture.

One difference between Case 1 and the reference case as designed is that Case 1 includes a sulfator which burns up the carryover fines removed from the gas stream while the reference case does not. This gives Case 1 more operating flexibility because incomplete carbon conversion in the gasifier can be partially counteracted by burnup in the sulfator. This capability, however, is not an exclusive feature of the hot-gas versus cold-gas cleanup approach. A carbon burnup combustor could be added to the reference case as well. The addition of the sulfator does, of course, make operations more complex in Case 1.

In Case 1 there is also the added complexity of the limestone handling and feeding system. There is also increased performance variability resulting from geological variations in limestone, affecting the catalytic gasification reactions and the sulfur capture efficiency. The overall system design has the flexibility to

Table 10-1

Comparison of Hot-Gas and
Cold-Gas Cleanup Systems

	<u>Case 1</u>	<u>Reference Case</u>
Net output, MW	458.4	413.5
Oxidant	Air	Oxygen
Gasifier trains	4	3
Sulfur removal	Limestone, Zinc ferrite	Selexol®, Claus, SCOT
Capital requirement \$/kW	\$ 1,342	\$ 1,547
Operating costs, \$1000/yr	\$ 57,513	\$ 52,422
OPERATING AND MAINTENANCE COSTS		
Fixed O&M, \$/kW-yr	38.3	43.9
Variable O&M, mills/kWh	2.41	0.99
Fuel costs, mills/ kWh	12.9	13.6

counteract such variations through increased carbon burnup in the sulfator and increased sulfur removal in the zinc ferrite beds, respectively. Keeping the system balanced in the face of limestone variations, however, makes operation a more complex task for Case 1.

10.1.1.2 Gas Treatment Train

Both cases utilize a firetube heat exchanger as the primary high temperature gas cooler. Beyond that, the systems diverge, as summarized in Table 10-2.

The cold gas system becomes essentially a chemical processing plant, including the Selexol® acid gas removal process, the Claus process for recovering elemental sulfur from the acid gases, and the SCOT process for treating tailgas from the Claus plant. Although these are all commercial processes, they are complex chemical processes, and operate much better on a continuous, constant flow basis than during startups, shutdowns, and changing flow rates. If system load growth patterns are such that the IGCC facility would always be dispatched at maximum availability, then the operability of the cold-gas cleanup train under changing conditions is less of a concern.

Table 10-2

Comparison of Gas Treating Trains

Case 1
Hot Gas Cleanup

Fuel Gas Stream:

Particulate removal in ceramic filter
Chloride removal in guard bed
Sulfur removal in zinc ferrite bed

Reference Case
Cold Gas Cleanup

Fuel Gas Stream:

Low-temperature gas cooling
Particulate removal filter (metal)
Cooling and condensate removal
Acid gas removal
Reheating and saturation

Other:

Sour water stripping
Sulfur recovery from acid gas
Treating of sulfur plant tailgas

A significant difference between the two systems is that the Selexol®, Claus, and SCOT processes would all be built as single trains. For the hot-gas cleanup system, the ceramic candle filter, the chloride guard bed and the zinc ferrite sulfur removal bed would all be built as four trains, one for each gasifier. This provides the overall system with considerably more flexibility (to operate under part load) and reliability (to maintain some output in the face of difficulties in the gas cleanup system).

An intangible difference between the two cases is the character of the processes involved. In the hot-gas-cleanup system, the interactions are all between solids and gases, and there are no liquid transfers. Operating the solid ceramic filters, the solid chloride guard bed, and the solid zinc ferrite bed would be very similar to operating electrostatic filters and baghouses, with which the electric utility industry has long been familiar. On the other hand, operation of the cold-gas cleanup system is a chemical plant operation. Liquid handling systems are involved, with the attendant possibility of leaks and the need for operators with chemical plant backgrounds or experience.

Reliability of the components of the hot-gas cleanup system must be proved through demonstration runs of sufficient duration. Performance of the ceramic candle filter elements is a major concern. Potential problems include:

- Ceramic failure due to thermal shock
- Ceramic failure due to thermal fatigue

- Ceramic failure due to chemical (e.g., alkali and steam) attack
- Seal failures at ceramic-metallic interface
- Pore blinding and rise in cleaning pressure

The lifetime of the zinc ferrite sorbent is another major concern. Factors affecting sorbent life include:

- The ceramic filter must prevent clogging of the bed with char fines.
- The chloride guard bed must prevent chlorides from reacting with the zinc ferrite.
- Moisture must be added to the gas to prevent carbon deposition on the zinc ferrite.

Although pilot plant tests of the hot-gas cleanup concept have been promising, the ultimate reliability is still unknown.

10.1.1.3 Gas Turbine

One key development item associated with the hot-gas system is the fuel gas control valve, controlling the feed to the combustion turbine. Material concerns on this valve limit the allowable temperature of the fuel feed to the combustor. Operation of such valves in commercial sizes has not been demonstrated at temperatures above 600°F, although it is believed that 1,200°F is within the state of the art. With a cold-gas cleanup system, the gas control valve is not a concern.

A second key development item is the operation of the gas turbine with the very low-Btu gas which may be produced during some portions of the air-blown operating cycle. Use of oxygen-blown medium-Btu gas has been well demonstrated. Steady-state operation of industrial gas turbines on low-Btu fuel gas has been demonstrated. However, the addition of steam may drop the heating value of air-blown fuel gas to 100 Btu per scf or less. Combustion of gas of this type in a modern high-performance gas turbine has not been demonstrated over a wide range of operating conditions. The Department of Energy and EPRI are currently conducting tests to define any problems which may exist.

10.1.2 Environmental Comparison

IGCC systems in general have a low environmental impact compared to older coal-fired systems for power generation. This is due to a combination of high thermal efficiency, almost complete sulfur removal, and relatively low volumes of solid and liquid waste streams. Thus both Case 1 and the reference case offer good environmental performance, and the differences between them are small in comparison to their differences from other non-IGCC systems.

10.1.2.1 Thermal Efficiency

Overall thermal efficiency is environmentally important for several reasons:

- Higher efficiency means less coal is required.
- Less coal means less environmental impact due to coal mining.
- Less coal means less ash to dispose of.
- Higher efficiency means less carbon dioxide released into the atmosphere.

The overall thermal efficiency for Case 1 is 5.2 percent higher than for the reference case. Efficiency in the oxygen-blown, cold-gas cleanup case suffers from the addition of an oxygen plant and the need to completely cool the gas stream, then reheat it after removing sulfur. A 5.2-percent higher efficiency in Case 1 means that 5.2-percent less coal must be mined to produce the same amount of electricity, and environmental impacts should be reduced accordingly.

10.1.2.2 Carbon Dioxide Emissions

Carbon dioxide emissions are determined by two factors: (1) thermal efficiency, which sets the amount of coal burned, and (2) carbon dioxide evolved during the calcination of limestone in the gasifier. For Case 1, the total effect is 1.8 pounds of carbon dioxide emitted per kilowatt-hour of electricity generated.

In the reference case there is no limestone, which accounts for about 6 percent of the CO₂ emissions in Case 1. However, the lower efficiency of the reference case requires a 5-percent increase in the amount of coal burned. On the whole, therefore, the reference case emits about 1 percent fewer pounds of carbon dioxide per kilowatt-hour generated, despite its lower efficiency.

10.1.2.3 Sulfur Dioxide Emissions

Both cases easily meet New Source Performance Standards for sulfur dioxide emissions from coal-fired power plants. Emissions from the hot-gas cleanup system, however, are appreciably lower. Total sulfur emissions to the atmosphere are about 3.6 percent of the original sulfur in the coal for the reference case, and only 0.6 percent for Case 1. The higher emissions of sulfur dioxide in the reference case were due to the selection of a design basis of 95-percent sulfur removal for the IGCC plant with no COS hydrolysis step prior to acid gas removal. A cold-gas cleanup system that employs a COS hydrolysis step would be capable of removing over 99 percent of the sulfur in the fuel gas stream. This level of sulfur removal for a cold-gas cleanup system was demonstrated at the Cool Water IGCC demonstration facility. Therefore, both IGCC systems have similar capabilities to remove sulfur from the fuel gas to the combustion turbine, thereby minimizing the sulfur dioxide emissions from either plant.

In Case 1, 99.4 percent of the original coal sulfur ends up in the LASH solids discharged from the sulfator. All of this material is disposed of as solid waste. A key technical uncertainty regarding Case 1 is how well the sulfator will work in converting calcium sulfide to calcium sulfate. For disposal purposes, calcium sulfate may be considered inert. Calcium sulfide, as formed in the gasifier, will be classified a hazardous waste. Resource Conservation Recovery Act regulations classify as hazardous any sulfide-bearing waste which generates hazardous quantities of toxic hydrogen sulfide vapors if exposed to a pH of 2.0. Untreated LASH would fail this test. Pilot plant experiments with the sulfator concept showed only partial conversion of calcium sulfide to sulfate, but the conversion was of sufficient extent to pass the EPA reactivity test for sulfide wastes. If the LASH should be classified as hazardous waste, disposal costs would increase greatly. There is also the potential that future regulations on ash disposal will become more stringent.

10.1.2.4 Nitrogen Oxide Emissions

Nitrogen oxide emissions for the two cases are relatively low in that both would meet the existing gas turbine NSPS standard of 75 ppm in the turbine exhaust. However, the nitrogen flows are quite different. In the reference case, almost all the fuel-derived nitrogen compounds formed in the gasifier are scrubbed out in the gas quenching step and appear in the sour condensate water. Thus the only NO_x in the turbine exhaust is thermal NO_x derived from air in the turbine combustor. This probably amounts to less than 25 ppm. This is approximately the level achieved by the Cool Water demonstration plant.

In the hot-gas cleanup case, the presence of limestone in the gasifier accelerates cracking reactions and reduces the amount of fuel-bound nitrogen compounds (chiefly ammonia) in the gasifier product gas. However, all of these compounds which do remain are passed on to the gas turbine, contributing to a turbine exhaust level of about 60 ppm. New Clean Air Act amendments will probably require that the Administrator of the Environmental Protection Agency promulgate new NSPS standards for NO_x by January 1, 1994. If New Source Performance Standards are lowered to 25 ppm, then some NO_x control procedure, such as SCR, would have to be employed. Moreover, the current situation is that construction permits granted under Prevention of Significant Deterioration (PSD) regulations take precedence over New Source Performance Standards standards. Under PSD regulations, EPA makes case-by-case determinations of what is the required Best Available Control Technology (BACT) for that region. Cost/benefit tradeoffs in recent reviews have concluded that it is justified to require Selective Catalytic Reduction in achieving NO_x concentrations as low as 9 ppm in some areas, and even as low as 3.7 ppm in Southern California. Thus there is a reasonable probability that in The Southern Company system, stack concentrations may be limited to 25 ppm in the future. In that event, Case 1 would be at a disadvantage compared to the

BACT reviews require the use of SCR in all cases, the relative disadvantage would be less.

10.1.2.5 Air Toxics

Little quantitative information is available concerning the emissions of trace constituents from KRW gasifier systems. Some materials of concern to turbine operation, such as alkali metals, vanadium, and chlorides have been monitored. In the hot-gas cleanup system, chlorides will be removed ahead of the zinc ferrite bed, and concentrations of trace elements coming out of the gasifier are generally less than 1 percent of the material contained in the coal feed. Exceptions are the low-boiling elements mercury and lead.

Most alkali metals are expected to be condensed onto or adsorbed by the ash particles by the time the particles reach the particulate filters and will thereby be removed from the system. Although some heavy organic materials may survive the gasifier conditions, they would be expected to be destroyed in the turbine combustor. However, the low residence time in the combustor may leave that question open.

In the cold-gas cleanup system, cooling of the entire gas stream to 105°F ensures that almost all metals and heavy organic materials will be condensed and appear in the process condensate. In the hot-gas cleanup system there is some concern that such materials could pass on through the system and be emitted in the stack gas. An unknown fraction of these materials could be absorbed in the chloride guard bed and the zinc ferrite bed.

Complying with the air toxic regulations in proposed Clean Air Act amendments could be a problem for hot-gas cleanup systems. Standards which have been proposed would limit the emission of nearly 200 listed pollutants to no more than 10 tons per year for any one pollutant, and no more than 25 tons per year in total. Although most of the listed materials are organic compounds, some trace elements in coal are listed, such as selenium, zinc, cadmium, arsenic, mercury, tin, and lead. Using theoretical volatility calculations, DOE has estimated that only lead would be likely to exceed the 10-ton-per year limit. Definitive test data are sorely needed in this area.

10.1.2.6 Solid Waste Disposal

Major differences exist in the amount and character of the solid waste generated by the two cases. For the reference case, the solid waste stream consists of 301 tons per day of gasifier ash and 190 tons per day of coal fines. Leaching tests performed on the ash agglomerates and fines from the Waltz Mill pilot plant indicate that they are nonhazardous under RCRA regulations and are therefore suitable for landfill by conventional methods.

For Case 1, the solid waste stream consists of 1,231 tons per day of LASH. As discussed earlier, adequate conversion of sulfide to sulfate in the sulfator will yield LASH which is nonhazardous. There is, however, some uncertainty about whether this can be successfully achieved on a commercial scale. There is also the possibility that EPA test procedures will be revised to make it more difficult to meet the definition of nonhazardous.

Designation of the LASH waste stream as hazardous waste would have severe economic consequences. Qualifying a hazardous waste disposal site is expensive and also subjects the company to a whole new range of regulations and procedures. Therefore an additional treatment step to improve conversion of sulfide to sulfate would be the preferred solution.

10.1.2.7 Waste Water Discharge

In the reference case, the process condensate blowdown from the sour water stripper must be sent through a biological wastewater treatment system. The treatment system includes the unit operations of equalization, bio-oxidation, biological sludge digestion/dewatering, and final effluent filtration. Digested sludge solids are disposed of by landfill and the clarified waste water is discharged to a stream.

In Case 1, there is no condensate stream from the hot-gas cleanup system and therefore no need for a wastewater treatment system. This aspect of the design simplifies the operation of the plant, reduces capital cost and complexity, and could streamline the permitting process for certain sites.

10.1.3 Cost Comparison

The estimated cost of electricity from the air-blown, hot-gas cleanup system designated as Case 1 is 67 mills/kWh (49 mills on a constant dollar basis) versus 74 mills (53 mills on a constant dollar basis) for the oxygen-blown, cold-gas cleanup system designated as the reference case.

The major reason for the lower COE in Case 1 is a 15-percent lower capital cost. The biggest difference in capital cost is the need for an air separation plant in the reference case. Because of its higher thermal efficiency, Case 1 also requires less coal (5 percent). However, the larger number of gasifiers in Case 1, along with the limestone handling and hot-gas cleanup systems, results in a 16-percent higher cost for nonfuel operations and maintenance. The net result is a 9 percent lower COE for the air-blown, hot-gas cleanup case.

Although the estimated capital cost is lower for Case 1, there is also more uncertainty in the equipment cost estimate. This is because of the larger number of gasifier trains along with the ceramic hot-gas filters and the zinc ferrite

desulfurization system. Commercial experience is lacking for all these components. In the higher-cost reference case, the air separation plant, Selexol® acid gas removal, Claus sulfur recovery and SCOT tailgas cleanup processes are all commercially tested and costs can be estimated with a relatively low range of uncertainty. The range of uncertainty for Case 1 can be reduced only by continued development of the KRW gasification and hot-gas cleanup processes, with demonstrations on a larger scale than has been achieved to date. For the current study, cost uncertainties for Case 1 were accounted for by assigning higher process contingencies to higher-risk process areas. Thus, the Case 1 capital cost has been adjusted to reflect technology risks.

The modular construction philosophy used in the KRW system designs does not point to large economies of scale from designing bigger facilities. Gasifier vessel sizes were selected on the basis of shop fabrication capabilities and would not be made larger for larger facilities.

Although there would be little economy of scale associated with progressing to larger facilities, there should be learning curve effects as more units are built. This type of behavior has been seen, for instance, in building Lurgi gasifiers in South Africa. If large numbers of shop-fabricated KRW gasifiers were built it would be expected that unit costs would drop. However, this would have little effect on the difference between Case 1 and the reference case, because they both use essentially the same gasifier. We may conclude that there will always be a difference in capital cost, favoring Case 1.

Zinc ferrite sulfur removal is the first generation hot-gas cleanup process. Other more effective sorbents are under development. Costs for sorbent could eventually be reduced as much as 50 percent. Successful development of these materials would provide Case 1 with an even stronger cost advantage.

Because the success or failure of the KRW gasifiers would affect both Case 1 and the reference case essentially equally, the economic difference between them will be shifted only in response to the performance of the hot-gas cleanup system. Costs of the ceramic candle particulate filters and zinc ferrite sulfur removal beds would have to approximately double before Case 1 would cease to have an advantage in capital costs over the reference case.

10.1.3.1 Methodology and Assumptions

Before any cost comparison between Case 1 and the reference case could be performed, the oxygen-blown reference case costs had to be updated based on mid-1990 dollars and to correct for design and performance revisions from the original study. Major changes to the process area capital costs of the previous reference study are listed below:

- The two major changes are (a) escalation from December 1987 to mid-1990 dollars (about 10.3 percent increase) and (b) scaling in costs for the new, larger coal feed rate with the revised oxygen-blown reference case (about 7.1 percent cost increase based on new/old coal feed rate ratio of 3845/3459 tons/day and a scale exponent of 0.65). The process areas, for which these two factors constitute the sole cost revisions made, include the following: air separation plant; KRW gasification; acid gas removal; sulfur recovery; tail gas treatment; sour water stripping; and wastewater treatment.
- The oxygen-blown reference case costs for the coal receiving, handling, and preparation area are revised based on scaling the ratio of the new coal rate for the reference case versus the coal feed rate for Case 1 in this study by using a scaling exponent of 0.65, and multiplying the resulting factor by the current Case 1 costs. This method of scaling from Case 1 costs is used because the costs produced in the current study are from more recent and definitive work than those for the previous reference case study.
- The oxygen-blown reference case costs for the ash/fines handling and disposal area are revised based on scaling the ratio of the solids handling rate for the reference case versus the solids rate for Case 1 in this study by using a scaling exponent of 0.65. The resulting factor is multiplied by both the current Case 1 costs and the factor for the size increase of the revised reference case (a 7.1 percent increase).
- The revised oxygen-blown case costs for the combined-cycle power generation island are based on scaling the original value to mid-1990 dollars (10.3 percent increase) along with specific modifications for the overhead crane and the turbine costs, resulting from the more recent and detailed costs of Case 1.
- The revised oxygen-blown case costs for the general plant facilities are based on scaling to mid-1990 dollars with specific increases (based on costs developed for Case 1) for clearing and grubbing, grading, permitting, fuel storage, and general overheads. However, the costs for railroads and some electrical requirements are decreased based on the costs developed in Case 1.

The new process contingency for the revised reference case is estimated by simply multiplying the old process contingency by the ratio of the revised/original process area cost in each area and using the sum of the results. The project contingency is estimated as described previously (i.e., 20 percent), which is the same value as used in the current study. The AFUDC for the revised reference case is estimated as for Case 1 in this study, which results in a value equivalent to about 19.6 percent of the total plant cost.

The owner's costs for the reference case have also been revised. The initial catalyst and chemicals are modified by escalating for the different year dollar

basis (10.3 percent increase) and the coal feed rate increase (7.1 percent cost increase). Working capital is corrected by increasing the original value by 10.3 percent for putting on a mid-1990 basis, and by increasing the 60-day supply of coal by the proper difference between the original and revised coal feed rates. Revised startup costs are generated by escalating the original values for the different year dollar basis and, also, by applying 2 percent of the difference in the revised and original total plant investment. Spare parts for the revised reference case are estimated as usual, 0.5 percent of process plant costs, except for coal and ash handling, which is 7 percent.

The O&M costs for the oxygen-blown reference case are estimated at a 65 percent capacity factor and are expressed in mid-1990 dollars. The major revisions for O&M costs of the revised reference oxygen-blown case include the following:

- Annual fuel costs are escalated from the original based on the new coal rates and \$35/ton cost used in this study.
- The variable cost (consumables, solids disposal, and byproduct credit) is escalated from the original value based on the revised/original plant size and year dollar basis correction.

The other O&M costs for the revised reference case are estimated by the same methodology as for Case 1, which is described in detail in Section 7.2.1 and again briefly below:

Operating labor - 23 men/shift (same as Case 1)

Maintenance labor - 30 men (same as Case 1)

Supervision and labor - 20 percent O&M labor

Maintenance material - percent of TPC (1.55 percent of TPC)

Insurance and taxes - 0.6 percent of TPC

Other operating costs - 10 percent O&M labor

The methodology and financial assumptions for estimating the COE for the revised reference case are identical to those used in producing the other results

for all the cases in this study. This information is previously defined in Sections 8.1 and 8.2 of this report.

10.1.3.2 Cost Results

The total capital requirement (TCR) for the revised oxygen-blown reference case and Case 1 are compared in Table 10-3 and Figure 10-1. The TCR for the oxygen-blown case is about 1547 \$/kW compared to 1342 \$/kW for Case 1. The reference case TCR is about 15 percent greater than for Case 1, and most of the difference is in the plant construction costs. The requirements for the air separation plant and other cleanup systems (Selexol®, Claus, and Scot) with the oxygen-blown reference case more than offset the larger fuel gas handling requirements for the air-blown/hot-gas cleanup case. The reference case does have smaller process contingency and other costs.

The first-year O&M cost comparison between Case 1 and the revised reference case is shown in Table 10-4 and Figure 10-2. The lower O&M cost is for Case 1 at 22.0 mills/kWh, while this cost for the reference case is very similar at 22.3 mills/kWh. Case 1 experiences lower fixed and fuel costs due to better efficiency and an economy of scale. The reference case enjoys a variable cost advantage because of no limestone and zinc ferrite consumption as in Case 1, along with a substantial byproduct credit for sulfur.

The COE results for Case 1 (air-blown/hot-gas cleanup) and the revised reference case (oxygen-blown/cold gas cleanup) are presented in Table 10-5, in both current and constant dollars, and are expressed graphically in Figure 10-3, in constant dollars. The COE results are shown as 10-year levelized costs based on a 65-percent capacity factor.

On a constant-dollar comparison, Case 1 possesses the lower COE with 49.1 mills/kWh. The reference-case COE is 53.4 mills/kWh, 8.8 percent greater than for Case 1. In examining the component costs (capital charges, fuel costs, O&M), the single largest component in both cases is capital charges; and the major difference in COE between the two cases is for the capital charges. As a result of the greater TCR for the revised reference case, the capital charge component of the reference case COE is about 15 percent greater than that for the air-blown/hot-gas cleanup case. The fuel cost component is about 5 percent larger for the reference case, while the O&M component is about 5 percent larger for Case 1.

Table 10-3

TCR for Case 1 vs. Oxygen-Blown.

	All Costs in Mid-1990 Dollars			
	<u>Case 1</u>		<u>Oxygen Blown</u>	
New Power Output, MW	458.4		413.5	
	\$MM	\$k/W	\$MM	\$k/W
Plant Construction Cost	378	825	405	980
Process Contingency	20	43	15	36
Project Contingency	80	174	84	203
Total Plant Cost	477	1042	504	1218
AFUDC	94	204	99	239
Total Plant Investment	571	1246	602	1457
Owner's Cost	44	96	37	90
Total Capital Requirement	615	1342	640	1547

Figure 10-1 Total Capital Requirement of IGCC Cases

Table 10-4

**O&M Cost Summary
Mid-1990 Dollars and 65-Percent Capacity Factor**

<u>O&M Cost</u>	<u>Case 1</u>	<u>Oxygen Blown</u>
Fixed O&M*, k\$	17,575	18,156
\$/kW-yr	38.3	43.9
mills/kWh	6.73	7.71
Variable, k\$	6,283	2,337
mills/kWh	2.41	0.99
Fuel Cost, k\$	33,656	31,928
mills/kWh	12.89	13.56
Total, k\$	57,514	52,422
mills/kWh	22.03	22.26

* Includes insurance and taxes

Figure 10-2 Comparison of O&M Cost Components

Table 10-5
Levelized Cost of Electricity Summary*

	<u>Case 1</u>	<u>Oxygen Blown</u>
<u>CURRENT \$ mills/kWh</u>		
Capital Charges	39.8	45.9
Fuel Costs	16.3	17.1
Operating & Maintenance	11.3	10.7
Total Cost of Electricity	67.4	73.8
<u>CONSTANT \$ mills/kWh</u>		
Capital Charges	26.7	30.8
Fuel Costs	13.2	13.9
Operating & Maintenance	5.1	8.7
Total Cost of Electricity	49.1	53.4

-
- * Ten-year levelized at 65-percent capacity factor with mid-1990 dollar basis.

Figure 10-3 Cost of Electricity for IGCC Cases

10.1.4 Tradeoffs and Risks

As noted in the preceding discussion, development of an air-blown gasifier with a hot-gas cleanup system offers significant advantages when compared to an equivalent oxygen-blown gasifier with a cold-gas cleanup system. These advantages include:

- Lower capital investment
- Higher thermal efficiency
- Lower cost of electricity
- Simpler operating system
- More compatible with advanced turbine cycles

Arrayed against these advantages are potential disadvantages in the environmental area, including:

- Emissions of air toxics
- Greater solid waste disposal problems
- Higher NO_x emissions

The first two of these are potentially major problems in that solutions could turn out to be not easily achievable. However, the severity of these problems depends on regulatory developments, which are difficult to predict at the present time.

In order to reap the potential advantages listed above and avoid the potential disadvantages, all the areas of major uncertainty associated with the process must be resolved. These uncertainties are mapped in Figure 10-4. Although none of these uncertainties appear unreasonable, there is at least some element of risk involved with each. A well thought-out and consistent research and development program is needed to remove uncertainties.

10.2 EXERGETIC AND THERMOECONOMIC EVALUATION OF CASE 1 AND THE REFERENCE CASE

An important issue in the current study of air-blown IGCC is the potential for further improvements in plant design through design optimization. This section describes the results of detailed exergy (second-law) analyses and thermoeconomic analyses conducted to evaluate design optimization and understand the underlying performance differences between air-blown systems with hot-gas cleanup versus oxygen-blown systems with cold-gas cleanup. The analysis focuses on Case 1, two variations in Case 1 labeled Cases 1A and 1B, and the oxygen-blown reference case.

Figure 10-4 Areas of Major Uncertainty in Air-Blown Hot-Gas Cleanup System

Initially, a brief introduction to the thermoeconomic analysis is given. Subsequently, the results of the analyses are presented and design options that may lead to a reduction in the COE are discussed. Finally, the conclusions from the exergetic and thermoeconomic analyses are summarized. More detailed results than the ones shown in this section are presented in Appendix F.

10.2.1 Introduction to Thermoeconomic Analysis

The complete exergy/thermoeconomic analysis techniques are somewhat specialized and will not be described in detail in this section but will be summarized to the level necessary to present the key conclusions. Greater detail on the methodology and the results is contained in Appendix F.

The exergy analysis of a process such as a power plant requires the calculation of the exergy of all streams in the plant, the rate at which exergy is "destroyed" or wasted in each component, and what is termed the exergetic efficiency of each component. The term "destruction" is used to identify the unrecoverable loss of exergy within the component, as distinct from the loss of exergy in an output stream. To understand the term exergetic efficiency, it is helpful to think of each component as having a "product," which represents the desired result from the component, and a "fuel," which represents the driving force for the process, or the desirable resources used to obtain the product. Using this terminology, the exergetic efficiency is defined as the ratio of the exergy in the product to the exergy in the fuel. The difference between exergy in the fuel and exergy in the product is the exergy destruction in the component being considered. The greater the exergy retained in the product, and thus the lower the extent of exergy destruction, the higher the exergetic efficiency of the component.

Exergy costing is based on the finding that exergy is the only rational basis for assigning costs to streams and to "energy waste". In exergy costing, one calculates the cost required to generate each stream in the plant, as well as the following costs for each plant component: (a) the cost of providing the driving force (fuel) to the component, (b) the cost of generating the desired output (product) from the component, and (c) the cost due to exergy destruction within the component. The costs must include capital, fuel, and nonfuel O&M charges associated with a given component. In general, it is recognized that increased exergetic efficiency (reduced exergy destruction) requires higher capital costs for a given component. This tradeoff is a key aspect of the thermoeconomic optimization of a process.

Based on the variables described above, the thermoeconomic theory shows that an explicit relationship can be developed for the cost-optimal exergetic efficiency for each plant component. Under certain assumptions, the system of equations for optimal component exergetic efficiency can be solved so that the optimal exergetic

efficiency is a known quantity for each component in the plant. A comparison of the optimal efficiency with the actual one for a plant component indicates whether it is more cost-effective for the plant to increase the capital expenditures associated with the component and reduce the cost of exergy destruction in the component, or vice versa.

10.2.2 Thermoeconomic Evaluation of the Design of Case 1

The scope of the current study called for the thermoeconomic optimization of Case 1 using the general approach described above. For a full thermoeconomic optimization of a power plant, it would be necessary to develop a number of design and cost estimates over ranges of each of several optimization parameters. In order to conduct a manageable optimization study, it was decided to concentrate on a single optimization parameter, viz., gasifier temperature, and limit the number of design variations to three, corresponding to gasifier temperatures of 1800°F (Case 1A), 1900°F (Case 1), and 2000°F (Case 1B). For each design temperature, new heat and material balances were developed, key components were redesigned, and new costs were developed for the key components. Selected variables for the three designs plus the reference oxygen-blown case are compared in Table 10-6.

As stated, the determination of an overall thermoeconomic optimum for a plant design requires that the relationship of component capital cost to component characteristics, such as size and exergetic efficiency, be approximated by a certain mathematical function (Appendix F). This makes solution of the system of thermoeconomic equations readily achievable. An evaluation of the cost data developed for Cases 1, 1A, and 1B indicated that some data could not be expressed in the required mathematical form. For this reason, the optimal exergetic efficiency from the economic viewpoint was calculated for only a small number of plant components of the gasification island. Nonetheless, the exergy analysis reveals important information about the Case 1 design, as is discussed in the following section.

10.2.3 Results of the Thermoeconomic Evaluation of Case 1

Figures 10-5 through 10-7 are simplified flow diagrams for the Case 1 gasification island, gas turbine/HRSG system, and steam system. These diagrams will provide a useful reference in discussing the thermoeconomic results. Table 10-7 summarizes the key thermoeconomic measures developed for Case 1 and its two design variations. The three key measures in the table are:

Table 10-6
Comparison of Selected Variables for Four IGCC Design Configurations

VARIABLE	CASE 1A	CASE 1	CASE 1B	REF CASE
Gasifier operating temp. [F]	1800	1900	2000	1850
Coal flow rate [lb/hr]	308682	315975	325850	320402
Coal exergy flow rate [MW]	1149.96	1177.13	1213.92	1116.57
Gasification pressure [psia]	400	400	400	465
Temperature at the outlet of: - product gas cooler [F]	1225	1244	1213	460
Air to Gas Turb. syst. [lb/hr]	5994222	5978825	5958190	6031842
Gas turbine firing temp. [F]	2235	2235	2235	2235
Gas temp. at the HRSG inlet [F]	1098	1098	1098	1113
Ambient temperature [F]	90	90	90	90
Main steam temperature [F]	1000	1000	1000	970
Main steam pressure [psia]	1450	1450	1450	1350
Reheat steam pressure [psia]	350	350	350	350
Condenser pressure [psia]	1.72	1.72	1.72	1.72
Steam turbine gross power [MW]	175.02	182.20	191.65	159.08
Gas turbine gross power [MW]	311.64	311.64	311.64	298.80
Total net power [MW]	451.75	458.41	467.19	413.19
Ratio of electric power generated in the gas turbines and steam turbines	1.78	1.71	1.63	1.87
Thermal efficiency [%]	40.05	39.70	39.24	37.72
Total exergetic efficiency [%]	39.28	38.94	38.49	37.01
Gasification Island Exergetic Efficiency [%]	77.68	76.91	76.09	81.73
Heat Rate [Btu/kWh]	8520	8594	8696	9045

Figure 10-5 Simplified Flow Diagram of the Gasification Island in Case 1

Figure 10-6 Simplified Flow Diagram of the Gas Turbine System and the HRSG in Case 1

Figure 10-7 Simplified Flow Diagram of the Steam Cycle in Case 1

Table 10-7
Exergy Destruction, Dimensionless Ratio Between Exergy Destruction and Exergy
of Coal (Ratio) and Exergetic Efficiency for Cases 1A, 1, 1B, and the Reference Case

COMPONENT	CASE 1A				CASE 1				CASE 1B				REFERENCE CASE			
	ED [MW]	RATIO [%]	EXERG. EFF. [%]	ED [MW]	RATIO [%]	EXERG. EFF. [%]	ED [MW]	EXERG. EFF. [%]	ED [MW]	RATIO [%]	EXERG. EFF. [%]	ED [MW]	EXERG. EFF. [%]	ED [MW]	RATIO [%]	EXERG. EFF. [%]
OXID. FEED SYSTEM	8.61	0.75	86.6	9.36	0.80	86.3	9.96	0.82	86.5	17.49	1.57	39.8				
OXYGEN PREHEATER	--	--	--	--	--	--	--	--	--	0.22	0.02	70.9				
GASIFIER	206.05	17.92	82.9	215.00	18.27	82.5	229.07	18.87	81.9	135.42	12.13	89.3				
CYCLONES & MIX	1.63	0.14	99.2	1.71	0.15	99.2	1.49	0.12	99.4	0.33	0.00	99.9				
RADIANT COOLER	--	--	--	--	--	--	--	--	--	15.87	1.42	66.3				
CONVECTIVE COOLER	--	--	--	--	--	--	--	--	--	9.84	0.88	73.2				
PROD. GAS COOLER	19.72	1.71	66.7	23.95	2.03	66.4	33.71	2.78	65.8	--	--	--				
(INCLUDING DRUM)																
GAS CONDITIONING	21.68	1.89	88.4	22.15	1.88	88.7	22.36	1.84	88.9	--	--	--				
SULFATOR	13.81	1.20	56.9	14.14	1.20	56.9	14.60	1.20	57.0	--	--	--				
ZINC FERR SYSTEM	6.05	0.53	99.4	8.26	0.70	99.2	5.86	0.48	99.4	--	--	--				
EXIT GAS COOLER	6.82	0.59	73.5	6.96	0.59	73.9	7.40	0.61	73.5	--	--	--				
RECYCLE GAS CLR	1.43	0.12	71.9	1.42	0.12	71.8	1.17	0.10	77.1	--	--	--				
REC. GAS COMPR. 1	0.15	0.01	84.2	0.14	0.01	84.2	0.15	0.01	84.2	0.02	0.00	83.7				
REC. GAS COMPR. 2	0.06	0.01	75.6	0.06	0.01	75.6	0.07	0.01	75.6	--	--	--				
G.T. & AIR COMPR.	80.83	7.03	79.4	82.65	7.02	79.0	85.49	7.04	78.5	82.88	7.42	78.3				
COMBUST. CHAMBER	192.82	16.77	82.5	192.57	16.36	82.6	192.40	15.85	82.6	225.17	20.17	80.2				
HRSG	35.41	3.08	84.5	35.21	2.99	84.6	35.11	2.89	84.7	38.74	3.47	82.9				
STEAM TURBINES	28.97	2.52	85.8	30.10	2.55	85.8	31.56	2.60	85.9	27.97	2.51	85.0				
REM. STEAM CYCLE	19.05	1.66	--	19.68	1.67	--	21.18	1.74	--	14.72	1.31	--				
OTHER COMPONENTS	0.67	0.06	--	0.70	0.06	--	0.68	0.06	--	29.87	2.68	--				
SUM	643.75	55.98	--	664.06	56.41	--	692.26	57.03	--	598.54	53.61	--				
EXERGY LOSSES	56.36	4.90	--	56.56	4.80	--	52.62	4.33	--	110.44	9.89	--				
EXERGY OTHER THAN																
COAL EXERGY TO	(1.88)	(0.16)	--	(1.90)	(0.16)	--	(1.89)	(0.16)	--	(5.60)	(0.50)	--				
TOTAL PLANT	698.23	60.72	39.3	718.72	61.06	38.9	746.77	61.52	38.5	703.38	63.00	37.0				

Exergy Destruction (ED) - Unrecoverable loss of exergy in a given component, which is equivalent to the wasting of useful energy. Exergy destruction is often associated with less-than-ideal heat transfer, mixing, pressure losses, or chemical reactions. Thus, a high exergy destruction generally indicates areas in which the process could be improved by higher investment costs or a revised design. This term is expressed in megawatts to facilitate comparison to the plant electrical output.

Dimensionless Ratio of Exergy Destruction to Coal Exergy (Ratio) - Expressed in percent in the table, this represents the portion of the coal exergy which is wasted or destroyed in a given component. The exergy in the coal is similar to the "energy" in the coal—e.g., the higher heating value.

Exergetic Efficiency - Ratio of the exergy in the product from a component (or process) to the exergy contained in the fuel. A high exergetic efficiency indicates that the component is performing well and minimum exergy losses are incurred.

Using capital and O&M cost data in conjunction with the exergy analysis summarized in Table 10-7, it is possible to calculate another useful index of thermoeconomic performance, which is the cost due to exergy destruction. This index is simply the loss in dollars per hour for a given component due to the destruction of exergy (useful work) in that component.

To better illustrate the usefulness of the exergy analysis, Figure 10-8 shows for Case 1 the exergy destruction in certain key components and the corresponding exergetic costs or dollar losses associated with these components. It will be noted that the least efficient components in the process from an exergetic standpoint are the gas turbine combustion chamber, the gasifier, and the gas turbine/compressor. When costs are also considered, it is clear that the gas turbine components result in the highest dollar losses of any of the components.

Referring to the results in Table 10-7 and the graphical illustration in Figure 10-8, the following discussion addresses the thermoeconomic analysis of each of the major components in Cases 1, 1A, and 1B. Where applicable, equipment numbers corresponding to Kellogg process flow diagrams in Section 4 are listed.

Figure 10-8 Exergy Destruction, and Associated Costs for Selected Plant Components in Case 1

Gasifier

As the gasifier operating temperature increases from 1800°F (Case 1A) to 1900°F (Case 1) and then to 2000°F (Case 1B), the thermoeconomic variables vary as indicated in Table 10-7. In the three cases being considered here, the gasifier has the highest exergy destruction among all plant components. Approximately 18 to 19 percent of the coal exergy is destroyed in the gasifier alone. Out of the 215.5 MW of exergy destruction in Case 1, 28.3 MW and 8.0 MW are due to the preheating (from the inlet temperature to the gasifier operating temperature) of nitrogen and oxygen in the air, respectively, and 5.1 MW of exergy destruction is caused by the vaporization and superheating of the water contained in the coal (4.98 weight percent). The pressure drop in the gasifier causes approximately 2 MW of exergy destruction. The remaining 172.1 MW of exergy destruction is due to the irreversibility of the chemical reactions in the gasifier.

With increasing gasifier operating temperature, both the exergetic efficiency and the gasifier capital costs decrease, whereas the coal throughput, and consequently the exergy flow rates, increase. Based on capital cost data provided by Kellogg (Appendix F), the relationship between gasifier capital cost and the change in gasifier temperature does not exhibit the expected mathematical form. When going from a gasifier operating temperature of 1900°F to a value of 1800°F, capital costs appear to increase sharply for little change in gasifier efficiency, whereas in the range between 1900°F and 2000°F the cost change is much more gradual versus efficiency. Also, the effect of capacity on the investment costs is stronger in the former than in the latter interval of gasifier operating temperature. For these reasons, the optimal exergetic efficiency from the cost viewpoint could not be calculated for the gasifier using the standard thermoeconomic approach.

However, if the gasifier operating temperature is used as the only decision variable, calculation of the cost per unit of product exergy indicates that the cost minimum for the gasifier is obtained in the range between 1850°F and 1870°F, as shown by the minimum point in Figure 10-9. This optimum value is obtained by considering the gasifier only and does not necessarily need to be the optimum gasifier operating temperature when the total plant is considered. However, because of the position of the gasifier (near the total plant inlet) and its importance to the IGCC plant, the optimal value for the total plant should not be far from the range indicated. In fact, detailed simulations and economic analyses conducted for several total plant configurations (similar to those of cases 1A, 1, and 1B) in steps of 10°F in the temperature range of 1800 to 2000°F confirm this hypothesis. For a total plant capacity factor of 65 percent, the minimum cost of electricity is obtained at 1870°F while at a capacity factor of 85 percent the optimal gasifier operating temperature is 1855°F. The design changes discussed in the following discussion might slightly affect these optimal values.

Figure 10-9 Cost per Unit of Product Exergy in the Gasifier
as a Function of Operating Temperature in Case 1

Cyclone (320L) And Recycle Gas Mixing

The exergy destruction in the cyclone ranges from 2.4 MW in Case 1A to 2.75 MW in Case 1B. It is caused primarily by the mixing of the recycle gas with the solids removed from the gas stream. Although the cyclone is not an energy conversion component, the actual cyclone exergetic efficiency (99.2 percent to 99.4 percent) is very close to the calculated optimal one from the economic viewpoint.

Product Gas Cooler (360C) and Heat Recovery Steam Drum (360F)

Depending on the case being considered, the exergy destruction in the product gas cooler and the heat recovery steam drum varies from 20.12 MW (Case 1A) to 34.27 MW (Case 1B). This represents 1.75 percent and 2.82 percent of the coal exergy, respectively. The exergy destruction is mainly caused by the heat transfer from gas to steam.

With increasing gasifier operating temperature, both the investment costs and the amount of steam raised in the components being considered (and consequently the exergy flow rate of the product stream) increase, whereas the exergetic efficiency decreases. As in the case of the gasifier, the relationship of cost versus efficiency does not have the required mathematical form and thus the economically optimal exergetic efficiency cannot be calculated. A more complex cost equation must be assumed for the product gas cooler and steam drum, provided that the investment cost data supplied for Cases 1A, 1, and 1B are accurate.

The values of cost per unit of product exergy suggest that if (a) the product gas cooler and the heat recovery steam drum are considered isolated, and (b) the gasifier operating temperature is the only decision variable, then the cost performance of these components is best around 1900°F.

Mixing of Steam and Gas from the Product Gas Cooler

The exergy destruction in the mixing process of steam (stream 39) with the gas exiting the product gas cooler (stream 12) is considerable: 16.7 MW, 17.0 MW, and 17.0 MW for cases 1A, 1, and 1B, respectively. In a new design developed for Case 1, every effort should be made to reduce the irreversibility at this point by reducing the mass flow rate of steam to the absolute minimum required for the operation of the zinc ferrite unit and/or by cooling the gas to a lower temperature before the gas is mixed with the steam.

Cyclones and Filters (421L/F: 422L/F: 430L/F/V: 431L)

These components together with the chloride guard systems which are discussed below form Area 400, Gas Conditioning. The exergy destruction in these components, excluding the mixing process discussed above, is 3.5 MW, 3.6 MW, and 3.7 MW for Cases 1A, 1, and 1B, respectively. When the mixing of steam and gas from the product gas cooler is considered together with the cyclones and filters, the high value of exergy destruction and the high associated cost (\$333-\$340 per hour of plant operation) indicate that more attention should be devoted to this area in future studies.

Area 600 -- Sulfation

The exergy destruction in Area 600 is 13.8 MW, 14.1 MW, and 14.6 MW for Cases 1A, 1, and 1B, respectively. The exergetic efficiency for each of the three cases is approximately 57 percent. The calculated economically optimal efficiency is above 80 percent. Thus, it might be cost-effective to increase the capital expenditures for this area, if this would lead to an increase in the efficiency of this area. One option to accomplish this might be to reduce the temperature at which the sulfator flue gas leaves the area by generating more steam in the sulfator and then mixing this flue gas with the gas entering the HRSG. However, this is not consistent with the current approach of using the sulfator flue gas to provide the energy and inert gas for coal drying. Clearly, it does not seem cost-effective to first dry the coal in coal preparation and then to add high-pressure steam to the gas exiting the product gas cooler. This arrangement is dictated by the constraints of gasification, on the one hand, and protection of the zinc ferrite sorbent, on the other. Efforts should be made to reduce or eliminate these constraints.

Chloride Guard Reactors (410D, DA)

The exergy destruction in the chloride guards is on the order of 1.5 MW and the cost of exergy destruction is approximately \$29 per hour of plant operation. There are few obvious improvements that could be made to this system from a thermoeconomic standpoint.

Area 250 -- Booster Air Compression

The exergy destruction rate in the air recuperator (251C), BFW preheater (252CA), trim cooler (252CB) and booster air compressor (251J) is approximately 2.0 MW, 0.7 MW, 2.2 MW, and 4.5 MW, respectively, and the cost associated with the exergy destruction about \$55/hr, \$17/hr, \$63/hr, and \$139/hr, respectively. All components in this area operate near their economically optimal exergetic efficiency values.

Zinc Ferrite Reactors (510D, DA)

These reactors are responsible for 8.3 MW of exergy destruction and for an associated cost of \$117/hr. No specific recommendations based on the thermoeconomic analysis can be made for the zinc ferrite reactors.

Exit Gas Cooler (510C)

The exergy destruction in the exit gas cooler is about 7 MW and the associated cost \$196/hr. The actual exergetic efficiency of 73.9 percent is only slightly higher than the economically optimal efficiency of 70 to 71 percent at a plant capacity factor of 65 percent.

Recycle Gas Cooler (361C)

The recycle gas cooler operates at an exergetic efficiency of about 72 percent in Cases 1A and 1, and 77 percent in Case 1B. The exergy rate of the product is 3.7 MW, 3.6 MW, and 3.9 MW for Cases 1A, 1, and 1B, respectively, whereas the corresponding investment costs (materials plus subcontract) are 3.23, 3.19, and 3.23 million dollars. These numbers indicate that the cost data are not consistent with the thermodynamic data because case 1B has approximately the same investment costs with case 1A although both the exergetic efficiency and the mass flow rates in case 1B are higher than in case 1A for the recycle gas cooler. Thus, the economically optimal exergetic efficiency could not be calculated for this component.

Heat exchangers such as the recycle gas cooler must be evaluated in the context of the total plant heat exchanger network. The results of an analysis of the heat exchanger network based on the pinch method is shown in Figure 10-10. This analysis shows that, in the recycle gas cooler, some heat transfer occurs through the pinch, which is at $T_{hot} = 520^{\circ}\text{K}$ and $T_{cold} = 495^{\circ}\text{K}$. This ineffectiveness increases both the energy and capital costs and must be corrected in future IGCC designs. The lower part of the heat transfer in the recycle gas cooler should be used as an IP economizer to preheat the liquid water entering the IP drum up to the saturation point.

Recycle Gas Compressor (381J)

The exergy destruction in the recycle gas compressor is 0.2 MW and the associated cost is \$6.21/hr. A relatively high proportion of the cost of exergy destruction for the recycle gas compressor is due to the capital and operating costs of this component, which is typical for compressors and pumps.

Figure 10-10 Composite Curves for the Heat Exchanger Network of the IGCC Plant in Case 1

Area 900 -- Gas Turbine System

This area is responsible for the highest exergy destruction rate among all IGCC plant areas. A total of 275.2 MW of exergy (i.e., 23.4 percent of the coal exergy) is destroyed in this system. The estimated exergy destruction in the combustion chamber is 192.6 MW (16.4 percent of the coal exergy), which represents the second highest value of exergy destruction among individual plant components. The cost rate associated with the exergy destruction is \$3,881/hr (the highest among all plant components) for the combustion chamber and \$1,331/hr for the gas turbine/air compressor system. These numbers combined with the relatively low investment costs of the combustor indicate that the cost effectiveness of the total plant might be improved significantly if a gas turbine process with a reheat stage were used. There is an urgent need to study the economic feasibility of mature gas turbine system designs including a reheat stage.

Area 1000 -- Heat Recovery Steam Generator

The exergy destruction rate in the total HRSG varies from 35.4 MW in Case 1A to 35.1 MW in Case 1B. The corresponding exergetic efficiency is of the order of 84.6 percent. Significant exergy destruction occurs in the superheater/reheater/blast steam reheater (12.9 MW with an associated cost rate of \$250/hr), feedwater heater no. 1 (4.2 MW with an associated cost rate of \$80/hr), the HP1 economizer (4.0 MW with an associated cost rate of \$77/hr), feedwater heater no. 2 (3.5 MW and \$65/hr), the HP drum (3.4 MW and \$66/hr), and the IP drum (2.9 MW and \$74/hr).

In the IP drum, 1.3 MW of heat is transferred through the heat exchanger network pinch (495°K/520°K), which is shown in the composite curve of Figure 10-11. This inefficiency must be corrected in future design configurations of the HRSG.

Figures 10-11 and 10-12 show the composite curves and the temperature/heat transfer profile for the HRSG. The minimum temperature differences occur in the HP2 economizer, IP superheater, and HP drum. Estimates based on the values of the HRSG thermoeconomic variables and on a comparison of Figures 10-11 and 10-12 indicate that it might be cost-effective to increase the temperature differences in the HP2 economizer and IP superheater and decrease the minimum temperature difference in the HP drum. At this point it must be emphasized that the HRSG should not be optimized by itself. An optimization study must simultaneously consider the HRSG, all heat recovery devices in the gasification island, and the steam turbine performance. The temperature, pressure, and mass flow rate values of the main steam and the reheat steam must be kept variable in such an optimization study. Prior studies by TTU Center for Electric Power indicate that a good match of temperature profiles in the HRSG has the potential to increase the overall plant efficiency significantly.

Figure 10-11 Composite Curves for the HRSG in Case 1

Figure 10-12 Q-T Diagram for the HRSG in Case 1

Area 1100 Steam Cycle

The total exergy destruction in the steam cycle (including the exergy losses in the condenser) is 48.0 MW, 49.8 MW, and 52.7 MW for Cases 1A, 1, and 1B, respectively. The actual steam cycle design considered by itself is cost efficient. The cost optimal design for this area, however, can be obtained through an integrated optimization of the overall heat exchanger network and the steam turbine.

10.2.4 Comparison Between Case 1 and Reference Case

In addition to the optimization of a given power plant design, second law/thermoeconomic analysis techniques are useful for comparing two distinctly different plant concepts. A thermodynamic comparison was conducted for the air-blown plant in Case 1 versus the oxygen-blown reference case. The major differences between Case 1 and the reference case for the present analysis lie in the following:

- Oxidant feed
 - The air feed to the oxidant feed system was not taken from the gas turbine system in the reference case.
- Coal gasifier
 - An oxygen-blown gasifier is used in the reference case as opposed to an air-blown gasifier in Case 1.
 - The coal moisture at the gasifier inlet is 11.12 weight percent in the reference case versus 4.98 weight percent in Case 1.
 - The gasifier operating temperature is 1900°F in Case 1 versus 1850°F in the reference case.
- Solids to be disposed of
 - Stream No. 16 (from gas filter to fines handling) in the reference case contains 4.11 percent of the carbon supplied to the gasifier with the coal whereas stream No. 24 (solids to disposal) in Case 1 contains less than 0.07 percent of the carbon fed with the coal into the gasifier.

- Gas cleanup systems
 - Case 1 uses a hot-gas cleanup process whereas the reference case uses commercially available cold-gas cleanup processes. The temperature of the fuel gas at the inlet to the combustion chamber is, therefore, 534°F higher in Case 1 than in the reference case.
- Byproducts
 - Case 1 has no useful byproducts as opposed to the reference case in which high-purity nitrogen and elemental sulfur are generated.

The comparison between the reference case and Case 1 is conducted with the aid of: (a) the ratio "r" of the exergy destruction rate in a plant component divided by the exergy rate of the coal supplied to the power plant, and (b) the ratio of the exergy loss (the exergy rate of a stream leaving the total plant) to the exergy rate of the coal supplied to the power plant. From the thermodynamic viewpoint, these dimensionless variables represent objective measures of the component performance and the value of a stream, respectively. These variables allow for the comparison of power plants with different capacities, different design configurations, and different coal moisture contents. A comparison of the values of these dimensionless variables for Case 1 versus the reference case is shown in Table 10-8. A graphical comparison of second-law performance by plant system, based on the ratio "r", is shown in Figure 10-13.

The oxygen-blown KRW gasifier has a considerably higher thermodynamic efficiency than the air-blown KRW gasifier. This difference is caused in part by the heating of nitrogen up to the gasifier operating temperature, but mainly by the differences in the reactions occurring in the gasifier. The product gas from the reference-case gasifier contains relatively more carbon monoxide and hydrogen and considerably more methane than in Case 1.

From the thermodynamic viewpoint, the lower gasifier operating temperature of the reference case represents an advantage versus Case 1, while the higher coal moisture in the reference case represents a disadvantage. The thermodynamic advantage of the oxygen-blown KRW gasifier is, however, partially offset in the gas turbine system. In the reference case, a larger air flow rate than in Case 1 must be compressed in the air compressor and, more importantly, heated in the combustion chamber.

An important question regarding the difference in thermal efficiency between Case 1 and the reference case is how much of this difference is attributable to the use of oxygen versus air and how much is due to hot-gas cleanup. Using the

values shown in Table 10-8 and the exergetic efficiencies calculated for various groups of plant components, the effects of these two features can be roughly estimated independently. Assuming two IGCC power plants with hot-gas cleanup designed under identical technical approaches and assumptions, with the only difference being that one plant uses an oxygen-blown KRW gasifier and the other an air-blown KRW gasifier, the plant with the oxygen-blown gasifier should have a thermodynamic efficiency of the order of 0.2 to 0.8 percentage points higher than that of the air-blown plant.

The overall plant efficiency improvement that can be attributed to the high-temperature gas cleanup process used in Case 1 compared with the low-temperature gas cleanup process of the reference case is probably of the order of 0.4 to 0.8 percentage points. The relatively small value of the total efficiency improvement is attributed to the relatively high irreversibility in the areas of gas conditioning and external desulfurization for Case 1. As shown in Table 10-8, if only gas cleaning and low-temperature gas cooling are compared, the exergy destruction in Case 1 is considerably higher than in the reference case.

Of course, the cost effectiveness of both plant types must consider capital costs and nonfuel O&M costs as well as thermal efficiency.

The foregoing conclusions suggest that the choice of oxidant and the method of cleanup do not fully account for the thermal efficiency differences determined for Case 1 versus the reference case in the current study. In fact, the exergy losses associated with material streams exiting the boundaries of the plants are mainly responsible for the difference in the overall efficiency between the reference case and Case 1. In the reference case, approximately 3.7 percent of the total exergy supplied to the plant by the coal is rejected to the environment with the solids-to-disposal stream as opposed to 0.07 percent in Case 1. This loss, which is due to the unconverted carbon in the solid waste, could be reduced in future design options based on the oxygen-blown KRW gasifier.

A secondary factor relates to the steam system. The comparison of the values given for the HRSG in Table 10-8 shows that the HRSG design in Case 1 is thermodynamically more efficient than in the reference case. This is due in part to the fact that the temperature profiles in the HRSG can be better matched in Case 1 than in the reference case.

Table 10-8

Comparison of Case 1 (Air-Blown) Versus
Reference Case (Oxygen-Blown)

Compares: (1) the exergy destruction in selected groups of plant components, and (2) the exergy losses associated with streams leaving the plant for the reference case versus Case 1. The values in the first two columns are given as a percentage of the coal exergy supplied to the corresponding IGCC power plant.

<u>Variable</u>	<u>Reference Case</u>	<u>Case 1</u>	<u>Difference</u>
I. EXERGY DESTRUCTION			
Gasification system			
(incl. oxidant feed system)	13.71	19.22	-5.51
Gas turbine system	27.61	23.38	+4.23
High-temperature gas cooling	2.71	2.03	+0.68
Gas cleaning and low- temperature gas cooling	2.15	3.31	-1.16
Heat recovery steam generator	3.47	2.99	+0.48
Steam turbines	2.51	2.55	-0.04
Remaining steam cycle	1.31	1.67	-0.36
Other components	0.14	1.26	-1.12
Sum I	53.61	56.41	-2.80
II. EXERGY LOSS			
Solids to disposal	3.70	0.07	3.63
Flue gas	2.36	2.83	-0.47
Sulfur	1.39	—	1.39
Other exergy losses	2.44	1.90	0.54
Sum II	9.89	4.80	5.09
SUM OF TOTAL EXERGY DESTRUCTION AND EXERGY LOSS	63.50	61.21	2.29

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

CASE 1 EXERGETIC EFFICIENCY = 38.94

REFERENCE CASE EXERGETIC EFFICIENCY = 37.01

**COMPARISON OF THE RELATIVE EXERGY
DESTRUCTION IN VARIOUS PLANT AREAS**

**Figure 10-13 Comparison of the Relative Exergy
Destruction in Various Plant Areas**

* Ratio of energy destruction or waste in a given component to the
exergy (useful energy) in the feed coal

10.2.5 Conclusions from the Thermoeconomic Analysis

The exergy analysis and the thermoeconomic analysis applied in this study identify and evaluate the real energy waste and the real cost sources in an IGCC design option. They are very useful tools in evaluating the potential for improving plant efficiency and cost effectiveness. With the aid of these analyses, conclusions can be drawn without redesigning the total plant.

This study shows that the cost effectiveness of Case 1 might be improved if the following design changes were made:

1. Operate the gasifier at a temperature ranging from 1850°F to 1870°F, as opposed to the proposed 1900°F.
2. Redesign the recycle gas cooler, IP economizer, and IP drum to avoid heat transfer through the pinch. Optimize the heat exchanger network in the total plant in conjunction with the steam turbines. This optimization might result in a reduction of both capital expenditures and energy costs. In addition, a reduction in the overall pinch temperature from about 34°C (61°F) in the current design to about 25° to 30°C (45° to 54°F) might prove to be cost effective.
3. Study the economic feasibility of a gas turbine system with a reheat stage. The thermoeconomic analysis indicates that this change might result in considerable cost savings.
4. Enhance the heat recovery operations in the sulfation area at the expense of coal drying. Compare the cost effectiveness of coal drying versus steam raising.
5. Compare Case 1 with a modified Case 1 design in which maximum sulfur removal is achieved in the gasifier. In this design, the cost of exergy destruction, the capital costs, and the nonfuel O&M costs associated with the zinc ferrite unit, the chloride guard reactor, and the mixing of steam and gas from the product gas cooler are eliminated. The modified design should be similar to the Case 6 design.

The gasification island produces two products: The fuel gas, which is used in the gas turbine and in the HRSG, and the steam which is used in the steam turbines. The ratio of cost per exergy unit between these two energy carriers can be calculated for the gasification island and, assuming the same contribution to the electricity cost, for the power island. This ratio is approximately 2.1 in the gasification island and 3.7 in the power island. In an ideally optimized IGCC design these two values should be equal. This is an additional indication that the

overall design of Case 1 is not optimal and could be improved by some of the methods suggested above.

The comparison of the thermodynamic performance between Case 1 and the reference case shows that the oxygen-blown KRW gasifier has a higher overall efficiency than the air-blown KRW gasifier. Some of these thermodynamic gains in the gasifier are offset in the gas turbine system, which is more efficient for the air-blown plant. Also, an IGCC plant based on hot-gas cleanup is thermodynamically more efficient than a similar plant which uses cold-gas cleanup even when the irreversibility associated with the hot-gas cleanup is significant, as is true for Case 1.

The exergy losses in material flows in the oxygen-blown reference case have a considerable impact on the overall efficiency. The efficiency of this case can be significantly improved by reducing the losses associated with the solids-to-disposal stream and optimizing the heat exchanger network.

From the above discussion it is apparent that an IGCC power plant which uses:

- a. an oxygen-blown, KRW fluidized-bed gasifier employing in-bed desulfurization,
- b. an optimized hot-gas cleanup process, if necessary,
- c. a combustor to burn any residual carbon remaining in the char fines, and
- d. an optimized heat exchanger network integrated with the steam cycle

should demonstrate significant performance improvements compared with either of the two design options discussed in this study.

However, the issues of capital cost, process complexity, and environmental performance must also be evaluated in selecting the appropriate IGCC technology for The Southern Company. These and other issues are addressed in Section 11.

11.0 IMPLEMENTATION OF GASIFICATION-BASED POWER PLANTS FOR THE SOUTHERN COMPANY

11.1 POWER GENERATION EXPANSION FOR THE SOUTHERN COMPANY

The power generation expansion plan for The Southern Company was described in Section 3.0 in connection with the study criteria. A major feature of generation additions has been the changing nature of electricity demand in the Southeast. During the 1960s and 1970s, there was a large growth in the demand for baseload power in the Southeast, and thus a major power plant construction program was initiated. When demand growth declined rapidly in the wake of the oil price shock of the mid-1970s, the result was an overshoot of baseload power generating capacity as new generating plants came online. Consequently, baseload capacity currently exceeds the need within The Southern Company's service territory, and substantial quantities of electric power are presently dedicated to off-system power sales.

The predicted annual growth rate of baseload power in the Southeast over the next 20 years is approximately 3 percent. For a system capacity exceeding 30,000 MW, this growth rate will represent a large increase in demand over the coming years. For the next 10 to 15 years, baseload demand requirements can be adequately met with the excess baseload capacity currently dedicated to bulk power sales. The immediate need for The Southern Company in the near future is for peaking power generation. Gas turbines have been identified as the technology of choice for meeting the peaking power demand. After the peaking power demand is satisfied, the next expected generation expansion is for intermediate load. Technologies presently being considered for intermediate power generation are cycling coal and natural gas or oil-fired combined cycle.

The next baseload power plant is expected to be placed in service in the year 2006. This date has continued to be adjusted closer to the present as growth in power consumption exceeds projections. For planning purposes, the next baseload power plant is assumed to be a subcritical 811-MW pulverized-coal (PC) unit with a scrubber for SO₂ removal and, if required by regulation, selective catalytic reduction (SCR) for NO_x removal. This type of unit is very similar to the latest coal-fired units built by Georgia Power Company. Other power generation options under consideration for baseload capacity include a PC unit with a supercritical steam cycle, plus a scrubber for SO₂ removal, and oxygen-blown IGCC with conventional cold gas cleanup. Significant experience has been obtained with supercritical PC units within The Southern Company, and therefore

the selection of this alternative over a subcritical PC unit will rest primarily on an evaluation of the relative economics, including such factors as heat rate, fuel flexibility, and reliability. Selection of a gasification-based power plant for the next baseload addition would represent a considerable departure from The Southern Company base of experience and would call for evaluation of technical and economic risk in addition to busbar costs. Gasification-based power plants such as those evaluated in the current study have not been formally considered as baseload candidates to date, and represent a further step in terms of familiarity, risk, and uncertainty.

With this background, it is worthwhile to consider how a gasification-based power plant might be implemented on The Southern Company system. The following sections discuss in greater detail the selection process for baseload technologies, the major criteria against which a baseload technology would probably be evaluated for The Southern Company, and finally how the gasification-based technologies addressed in this study appear to stack up against the selection criteria and how these technologies might be implemented.

11.2 SELECTION PROCESS FOR BASELOAD POWER GENERATION

Currently within The Southern Company, decisions with regard to implementation of power generation technologies are made as part of an integrated resource planning process. This process, which is coordinated by SCS, involves representatives from all operating utilities as well as the service company. The integrated resource planning process receives input in many areas related to generation planning, including supply side factors, demand side factors, fuel projections, environmental factors, and financial criteria.

With regard to supply side factors, a systemwide Technology Panel plays a major role in the identification of candidate technologies to meet projected load growth. This panel is an ad hoc committee which meets annually to develop recommendations to the integrated resource planning process regarding technology options for generating capacity. Information on new or emerging technology options, such as gasification-based power plants, typically comes from the following sources:

- Screening studies
- Site-specific design studies
- Field research and development programs

These evaluations are led by SCS Research and Environmental Affairs (R&EA), with significant involvement from SCS engineering and SCS system planning. At this stage, operating company involvement focuses primarily on providing study criteria and reviewing study results. Screening studies are typically used to assess the performance and costs of generating technologies to see if they warrant

further investigation. Once a promising technology has been identified, site-specific design studies, such as this one, are conducted to better define the advantages and disadvantages of the technology, gain engineering experience, and identify the major barrier issues for the technology. Detailed engineering data, scaleup, and the resolution of barrier issues for promising technologies are addressed in either pilot plant or retrofit projects.

Technology Panel recommendations on generation options are factored into the integrated resource planning process along with all the other inputs mentioned. The product of this process is an integrated resource plan, which projects system load growth and identifies the best methods for meeting the load growth, including the types of new generation needed and the schedule for generation additions. Generating additions are classified as peaking, intermediate, or baseload capacity. In general, the plan will identify the particular type of technology (e.g., PC-fired plant with FGD) for each planned addition. If the addition is some years in the future, however, multiple candidate technologies will be carried as possible options. As the date for a given generation addition approaches, detailed engineering and planning studies will be commissioned to determine the technology of choice. This process recently occurred for The Southern Company with respect to additions of peaking capacity in the mid-1990s. The detailed studies confirmed the results from earlier screening studies, which indicated that combustion turbines were the best choice for peaking capacity additions.

Responsibility for approval of the completed integrated resource plan rests with a systemwide Operating Committee. This committee is comprised of key decision makers from all system operating utilities as well as SCS.

In summary, the process of selecting a baseload technology for implementation by an operating utility of The Southern Company involves a comprehensive planning process in which input from system planning, financial planning, research and environmental affairs, engineering, production, and other areas is evaluated. Some of the important technology-related criteria by which the selection of a baseload technology will be made are discussed in the following section.

11.3 SELECTION CRITERIA FOR BASELOAD POWER GENERATION

The relative merits of gasification technology will vary depending on the characteristics of different power producers. Consideration of gasification-based power generation for baseload additions in The Southern Company will depend on the relative advantages and disadvantages of this technology versus the most likely alternative, which is a PC plant with FGD. The following discussion is an attempt to examine some key selection factors in view of needs and characteristics of The Southern Company. Major factors affecting the choice of baseload technology include the following:

- Economic factors
- Environmental factors
- Resource requirements and siting
- System factors (load growth, fuel issues)
- Plant sizing and scale economies
- Other considerations

Each of these areas will be discussed briefly.

11.3.1 Economic Factors in Baseload Technology Selection

The primary criterion by which competing technologies of equal risk are judged remains busbar costs, although other factors are gaining in importance. A major strategic goal of The Southern Company is to be in the lowest quartile of electricity cost for the Southeast. Revenue requirements analysis therefore plays an important part in the screening and analysis of competing baseload technologies. The capital cost of a PC unit with FGD for The Southern Company is relatively low due to standardization of design, multi-unit sites, low regional labor rates, and substantial economies of scale. In addition, production costs for existing units and the next planned baseload unit (PC plus FGD) are also low compared to the national average. However, low production costs are due primarily to low fuel and manpower costs rather than to high thermal efficiency. The tendency for recent baseload units has been to spend capital dollars for reliability and fuel flexibility, rather than for increased efficiency. With the current excess baseload capacity, the value of reliability has declined, suggesting that there may be a willingness to sacrifice some reliability in the future in order to reduce capital costs. Another development favoring low capital cost is uncertainty regarding the recovery of capital investment in large plants. In fact, full cost recovery has not been allowed for the most recent units added to the system.

Considering all these factors together, The Southern Company economics with respect to the next baseload power plant would appear to favor minimization of capital cost by building large plants and sacrificing some plant performance. Moreover, in order to compete favorably with the PC option, a candidate technology such as IGCC should have production costs which are inherently low due to plant simplicity, scale economies, and O&M requirements similar to those of existing plants. Thermal efficiency is an important advantage if it can be achieved without substantial capital cost outlays.

11.3.2 Environmental Factors

Environmental criteria are expected to play an increasing role in the selection of future baseload power plants. New Source Performance Standards (NSPS) are expected to become more stringent and must, as a minimum, be met by any

candidate technology. Emissions caps on SO_2 will force offsets of emissions produced by new units, and therefore technologies that can do better than NSPS will have some additional value in the future. This aspect favors gasification-based technologies which can achieve sulfur reductions of 99 percent with little increase in capital cost. PSD reviews, including BACT analysis, are expected to become more stringent in the future, with the probability that some type of supplementary NO_x control will be required. In general, the cost of supplementary NO_x reduction--whether by combustion controls or post-combustion controls--is expected to be substantially less for gasification-based power plants than for PC power plants, resulting in an additional advantage for gasification systems. Air toxics legislation is projected to have a moderate effect on utility power plants in the near term. Conventional IGCC with cold-gas cleanup and a PC plant with FGD should both be good performers with respect to air toxics. Gasification-based power plants with hot gas cleanup, such as those in the current study, will probably be riskier with respect to air toxics. Unfortunately, it is difficult to predict whether this will be a barrier issue for the technology without further research work and good predictions of likely regulatory developments.

Notwithstanding the comments above, the service territories of The Southern Company utilities are in general less sensitive with respect to air emissions than many other parts of the country, notably New England, Wisconsin, and California. Most of the generating capacity of The Southern Company is sited in rural locations where ambient air quality is not degraded. (The exception to this is ozone nonattainment areas near major metropolitan centers such as Atlanta and Birmingham.) Therefore, PC plus FGD is projected to remain a permissible baseload technology in most parts of the region, in contrast to sensitive regions in which gasification-based power plants may be the only permissible coal-based technology. Gasification-based power plants may have a permitting advantage over PC with FGD near the large metropolitan areas.

An important air emissions issue now developing is global warming, which is linked by some to CO_2 emissions. It is too early to predict what regulations may be imposed on fossil fuel plants to deal with this issue. If specific limits are imposed on CO_2 emissions, then high-efficiency coal-based technologies such as gasification power plants are favored over PC plants with FGD. Among gasification power plants, air-blown plants with hot gas cleanup, perhaps excluding in-bed desulfurization, would be favored over conventional IGCC with cold gas cleanup. In general, higher efficiency reduces specific emission rates of all airborne pollutants, and thus combined-cycle technologies offer the most potential in this respect.

Environmental sensitivity is also increasing with respect to solid wastes and wastewater discharge. In the area of solid wastes, IGCC power plants based on entrained gasification with cold gas cleanup offer an apparent advantage over either a PC with FGD or gasification with hot-gas cleanup in the form of in-bed

desulfurization. The waste solids from gasification with in-bed desulfurization are disposed of in a dry landfill, which simplifies disposal versus PC with FGD. However, concerns about future rulings regarding the leachability and toxicity of the solids from systems with in-bed desulfurization may outweigh any such advantages. With respect to wastewater, gasification with hot gas cleanup has an advantage in that there are no substantial wastewater streams arising from gas scrubbing, and hence water permits should be easier to obtain. Considering both waste solids and wastewater, gasification-based systems would appear to have advantages—different ones depending on the system—over PC plus FGD.

11.3.3 Resource Requirements and Siting

Factors considered here include the availability of coal, water, and other materials; the availability of plant sites and any limits on space or land area; and any other constraints on the permissibility of coal-fired power plants. With respect to coal availability, The Southern Company is one of the largest coal burning utilities in the United States and both Alabama and Georgia have good transportation access to coals, particularly Eastern bituminous supplies. This fact favors neither PC nor gasification-based plants. However, it does require that a gasification technology selected for baseload addition have a demonstrated capability to burn bituminous coals. Limestone supplies are also readily available in The Southern Company service area, and therefore hot-gas cleanup systems that rely on limestone, like FGD, are not penalized.

The Southern Company service region receives relatively large amounts of rainfall and has substantial water resources such as rivers, lakes, and coastlines. Therefore, water is readily available at many plant sites. Nonetheless, some rivers are overused, and water availability is an issue in many parts of the region, notably northern Georgia. For sensitive sites, gasification-based plants offer reduced water consumption versus PC plants with FGD. Air-blown (or highly integrated oxygen-blown) IGCC with hot gas cleanup offers substantial reductions in water usage due to the fact that little or no water is needed for NO_x control, flue gas scrubbing, and solids disposal.

In general, siting of coal-fired power plants is less of an issue in The Southern Company than in many other utility systems. A majority of the plants in The Southern Company are located on large rural plant sites with room for expansion. Therefore, the plant footprint is not a very significant criteria as it may be for other locations in the U.S., such as the Northeast. Moreover, many of the plants have substantial room for solids disposal. Many of the sites are already permitted for coal which eases the permitting for adding additional coal capacity. With respect to plant siting, then, gasification power plants will not have a significant advantage over conventional systems such as PC plants at many potential sites.

11.3.4 System Factors (Load Growth, Fuel Issues)

Planning projections for The Southern Company call for moderately high load growth, moderate coal pricing, increasing gas and oil prices, and no significant problems with the availability of oil or gas. Under the scenario of higher than expected growth, baseload options with short lead times are favored, so that gasification-based power plants would be expected to have some advantage over PC plants. Also, under high growth, combined-cycle plants built for intermediate load would be constructed for efficient conversion to IGCC plants, again increasing the likelihood of implementation of IGCC on the system.

Under low-growth scenarios, new capacity additions are delayed as long as possible, and technologies with shorter lead times become very important. Baseload options with the potential for short lead times or staged construction, such as IGCC, are more likely to be selected for installation over conventional technologies.

Another important system planning consideration related to combustion turbine installations is uncertainty surrounding the availability of gas and oil. While supplies of these fuels are expected to be available, a large commitment to combustion turbines on the system may call for the construction of at least one baseload coal plant capable of producing liquid fuels. This would probably be an oxygen-blown IGCC plant with conventional cold gas cleanup and a once-through methanol unit. If premium fuel supplies are constrained, or prices rise substantially, any combined cycles fired with oil or natural gas would be converted to IGCC.

If coal costs increase more than expected, the total production costs of both PC and gasification power plants would increase.

Because the gasification-based plants have better heat rates than a PC unit, the increasing coal costs would favor selection of the gasification power plant. Conversely, lower coal costs would favor the conventional PC with FGD.

11.3.5 Plant Sizing and Scale Economies

The Southern Company is a large system with over 30,000 MW of generating capacity. This requires that additional power generation occur in large increments to match the demand in load growth. The construction of large plants allows the realization of substantial scale economies for conventional power plants. In this context, gasification-based technologies under consideration for The Southern Company should not be limited in cost-effective plant size due to constraints on scale-up, plant configuration, or other factors. In the current study, the number of KRW gasifiers (four to six) required for the larger plants due to scaleup and shipping limits raises some concerns about costs and complexity. Because of their

modular nature, the gasification power plants considered in this study do not show the degree of scale economies that would be expected from PC plants with FGD. Thus, while gasification-based plants may be substantially lower in cost than PC plants at small scale, it remains to be seen how well the advantage holds up for very large plants. The current study does not answer this question, since the maximum plant size investigated is nominally 400 MW. This issue needs to be addressed specifically in future studies.

Large increments of growth do not favor phased construction of an IGCC plant by first constructing the combined-cycle portion of the plant followed by the addition of the gasifier at a later date. Also, The Southern Company system has a substantial need for peaking and intermediate power generation, which makes it counter-productive to convert simple- or combined-cycle units to baseload, because the peaking and intermediate demand would still exist.

11.3.6 Other Considerations

There are a number of other considerations which could affect the implementation of gasification-based power plants within The Southern Company. Like many regulated utilities, The Southern Company must be considered risk averse, and thus favors demonstrated technology with a proven track record. The adverse experience in recovering investments in nuclear plants has heightened the emphasis on making decisions on baseload power generation that will be deemed prudent by the public service commissions. This position will tend to favor the lower risk technology, which is PC with FGD. Among gasification power plants, technologies which have undergone a significant U.S. demonstration will undoubtedly be favored in the selection of baseload technology. To date, experience favors conventional IGCC with cold gas cleanup, which has been successfully demonstrated at the Cool Water plant. It is unlikely that air-blown gasification with hot gas cleanup would be selected for the next baseload addition in The Southern Company unless a successful demonstration is mounted.

Other factors which will be important in selecting the next baseload plant deal with plant familiarity, operability, and reliability. System personnel are most familiar with pulverized coal power plants, and The Southern Company has a first-rate design and operating capability with this type of plant. This background would tend to make the PC plant a preferred option for new baseload addition to the system. In addition, The Southern Company has limited expertise in engineering and operating plants which have complex chemical processing steps, in part because the coal-fired plants in the system have no scrubbers for SO₂ removal. Also, the system has had little integration or intimate collaboration with chemical industries in its service territory.

For these reasons, gasification with hot gas cleanup may be more readily accepted as a baseload option because it has gas stream cleanup steps that are more

familiar to the utility industry than conventional IGCC systems. On the other hand, The Southern Company is a large system with demonstrated capability in designing and operating sophisticated plants, including several nuclear units. Therefore, there is no doubt that The Southern Company could adapt to the engineering and operation of a more complex gasification-based system if this system had clear advantages over other options.

11.4 IMPLEMENTATION OF GASIFICATION POWER PLANT FOR THE SOUTHERN COMPANY

Based on the foregoing discussion, it is evident that the selection of a gasification-based power plant as the next baseload addition for The Southern Company will involve a complex evaluation of many important tradeoffs. System criteria favor large central stations (> 800 MW) with substantial economies of scale, inherently low capital and operating costs, and systems which match the engineering and operations experience of a PC-based system. On the other hand, compared to utilities in heavily populated or environmentally sensitive regions, less value will probably be assigned to state-of-the-art environmental performance, low land area requirements, and special siting advantages, such as low resource requirements and minimal waste production. These observations suggest that, under expected scenarios of load growth and fuel supply, power plants based on air-blown gasification with hot gas cleanup could prove to be an attractive baseload option for The Southern Company. This assumes that this technology is demonstrated at a meaningful scale (> 200 MW) in time to include it in system planning.

On the other hand, a number of alternative scenarios would favor PC units or oxygen-blown IGCC with cold gas cleanup. Concerns about the supply of premium fuels may dictate that an oxygen-blown IGCC plant be installed to provide syngas for liquid fuel production. Repowering of oil- or gas-fired combined cycles, which could occur in the event of rapid load growth or high oil and gas prices, could also demand oxygen-blown technologies for compatibility. More stringent environmental regulations, particularly with regard to NO_x and air toxics, could preclude the selection of IGCC based on hot gas cleanup.

For either conventional IGCC or IGCC with hot gas cleanup, a number of steps need to occur to allow selection of these technologies over a PC unit by the year 2006. A conceptual schedule leading to implementation of IGCC as the next baseload unit is shown in Figure 11-1. Despite the fact that the next baseload unit is some 15 years away from entering commercial service, the likely implementation schedule is rather tight. Demonstrations of conventional IGCC in utility settings are expected to occur in time to maintain this schedule. However, it is not clear at present that air-blown IGCC with hot gas cleanup will be demonstrated at sufficient scale soon enough to reach commercialization by the year 2006.

Figure 11-1 Conceptual Schedule for Implementation of New Baseload

11.5 POTENTIAL BARRIER ISSUES TO IMPLEMENTATION OF AIR-BLOWN GASIFICATION WITH HOT-GAS CLEANUP

The current study has served to identify a number of important issues which must be addressed in the scaleup and demonstration of power plants based on air-blown gasification with hot-gas cleanup. These can be summarized under the following headings: gasification, hot-gas cleanup, combustion turbine, and other. The major development issues are listed in Table 11-1 and discussed in the following sections.

11.5.1 Gasification Development Issues

The KRW gasifier has undergone extensive testing at pilot scale, and therefore many potential barrier issues have been addressed. However, several issues remain. One important question is the scaleup of the KRW gasifier to the limit defined in this study, which is approximately 14 feet in diameter. The KRW gasifier relies on well-defined zones, feedpoints, and circulation patterns within the gasification and fluidization zone. Scaleup of the vessel diameter could therefore be a development hurdle. Some cold-flow modeling has been done to address these issues, but final demonstration is necessary.

Another issue with regard to the KRW gasifier is the effect of limestone on maximum fluidization velocity and thus gasifier throughput. This study assumes that the presence of limestone causes a denser bed, thus allowing higher fluidization velocities without excessive elutriation of fines. This assumption allowed a reduction in the number of gasifiers required for the nominal 400-MW plant versus the same size plant without limestone (e.g., four gasifiers in Case 1 versus six in Case 5), with significant impacts on cost. While this design assumption is based on some pilot plant data, more experience in a large operating system is required for confirmation.

Much more uncertainty exists concerning the ultimate performance of the transport gasifier/combustor technology assumed in Cases 6 and 7. The transport concept is based on the design of fluid catalytic cracking units, with which Kellogg has a substantial experience base. In addition, tests of the transport reactor design have been conducted by Kellogg at pilot scale at atmospheric pressure in a combustion mode. Kellogg is planning further tests using a small-scale transport reactor test unit which can operate at high pressure in both gasification and combustion modes. From this background, Kellogg has developed design models which predict the performance of the transport reactor in various modes of operation, and these models were used for the sizing and performance rating of the transport systems assumed in the current study. Most of the cost and performance advantages determined for Case 6 in the present study rely fundamentally on the predictions of the Kellogg transport reactor design models. These design models need to be verified with additional testing at small scale

Table 11-1

**Major Development Issues for Air-Blown Systems
With Hot Gas Cleanup**

KRW GASIFICATION

- Scaleup to commercial size
- Gasifier throughput

TRANSPORT GASIFICATION

- Kinetics of reactions
- Sulfur capture limits
- Carbon conversion
- Scaleup
- Capital costs

HOT GAS CLEANUP

- Reliability, performance, and cost of hot particulate filters
- Performance, costs, and system effects of zinc ferrite; improved sorbents
- Fate of fuel-bound nitrogen in fuel gas and effects on NO_x emissions
- Carryover of volatile alkali compounds to the combustion turbine
- Emissions of trace compounds (potential air toxics), notably mercury and lead

COMBUSTION TURBINE (CT)

- Combustion characteristics of low-Btu gas
- Extraction of gasifier air from the CT compressor
- Cooling of the CT components and blades
- Fuel control valve for high temperature fuel gas
- Control issues relating to integration between CT and the gasification island

OTHER ISSUES

- Characteristics of sulfide containing waste solids; potential disposal costs and liability
- Performance of materials and components under exposure to high temperatures and pressures (e.g., valves, piping, heat exchangers)

followed by scaleup to a prototype and demonstration sizes. Issues which must be addressed for the transport technology include:

- The kinetics of gasification and combustion for small particle sizes, which will largely determine the transport equipment size
- Limits to carbon conversion for the transport gasifier, including the effects of char reactivity loss and the role of coal type
- The kinetics, thermodynamics, and inherent limits for sulfur capture by limestone in gasification and combustion modes
- The production rates of tars and other condensible hydrocarbons for the transport gasifier, including the effects of limestone
- The feeding, withdrawal, and recirculation of the fine solids utilized for the transport concept
- Scaleup limits for the transport reactor
- The true equipment costs associated with a transport-type system

Until these and other issues are resolved, the superior cost and performance predicted for transport-based systems such as Case 6 must be considered less certain than for the KRW-based systems.

11.5.2 Hot-Gas Cleanup Development Issues

Hot gas cleanup appears to yield substantial advantages with respect to capital cost reduction and efficiency improvement for gasification power plants. A number of development issues must be addressed in order to realize these advantages. A pressing issue is the long-term reliability, performance, and costs associated with the hot particulate filtration devices necessary to achieve hot gas cleanup. To date, such devices have not been tested at sufficient scale, over sufficient operating periods, and with appropriate systems integration to consider them to be commercial for high temperature applications, especially with regard to gasification. Another key development area is desulfurization with zinc ferrite sorbent, as assumed for Cases 1 through 4 of this study. Long-term testing is required to fully determine the costs and performance of this system. The sorbent cost alone is a substantial operating expense, even under favorable assumptions. Adverse experience with sorbent life could make external desulfurization uncompetitive with other alternatives. In addition, substantial moisture addition to the fuel gas was required in the designs developed in this study in order to protect the zinc ferrite sorbent. This requirement has a detrimental effect on overall cycle efficiency, water consumption, and particularly the combustibility

of the low-Btu gas. Advanced sorbents such as zinc titanate may obviate this requirement. Finally, some concerns have been raised regarding the operability of a fixed-bed zinc ferrite system which depends on high-temperature valves for isolating vessels during regeneration. These concerns must be addressed at demonstration scale.

A major unresolved issue with respect to hot gas cleanup is the compounds that are not removed from the fuel gas due to the lack of a scrubbing step in the process. Fuel-bound nitrogen (chiefly in ammonia and hydrogen cyanide) are of concern because they yield NO_x when burned in the combustion turbine. None of the cases addressed in this study is predicted to achieve NO_x emissions significantly below the NSPS limit for combustion turbines (75 ppmv) without SCR. More stringent limits are being and will be required for future combined-cycle installations, and thus SCR should probably be considered a necessary component of hot-gas cleanup systems. Other compounds of concern include volatile alkalis, which can promote combustion turbine corrosion, and volatile metals which could be categorized as "air toxics." Some experience related to these types of compounds was cited for the KRW pilot plant in Section 9. Based on these data, the only coal-borne elements that would appear to report to the gas in significant levels would be chloride and mercury. More work needs to be done on understanding the trace emissions from systems using hot gas cleanup, and regulatory developments must be monitored closely.

11.5.3 Combustion Turbine Development Issues

Several issues were raised regarding combustion turbines (CT) in service in air-blown plants with hot gas cleanup, particularly as applicable to the GE 7F machine. The very low heating value (below 90 Btu/scf) of the fuel gas from the air-blown KRW gasifiers poses a concern to turbine manufacturers. Combustion tests to date have not furnished sufficient evidence that such gases can be burned reliably and cleanly in can-type combustors. Such gases could probably be burned successfully in the silo-type combustors used by some CT manufacturers. Another important issue is the extraction of air for gasification from the CT compressor. For the GE 7F, modifications must be made to ensure that air can be extracted without adversely impacting the compressor discharge flow. Also, cooling of metal components in the CT is complicated by the large quantities of extracted air, as well as by the high temperature of the fuel gas from the hot-gas cleanup system. Other CT issues include the operability of a fuel control valve at high gas temperatures (1000°F or higher) and general control issues related to the tight integration between the CT and the gasification island.

These and other turbine-related issues must be addressed before air-blown systems with hot gas cleanup can be commercialized. Many of the questions related to CT air extraction and combustion of low-Btu gas should be answered (at least for European-type turbines) in the course of upcoming demonstrations of

the "highly integrated" oxygen-blown gasification power plants. It is not clear that there will be sufficient near-term development activity in hot-gas cleanup systems to encourage U.S. CT suppliers to address these and other issues for the U.S.-type machines.

With regard to the ISTIG cases developed in this study, there are too many development and commercialization issues associated with these paper concepts to permit a full discussion here. Based on the studies in this project, the most important need would appear to be the development of an aeroderivative CT with sufficient mass flow capability to accept all steam generated in the gasification island. More flowsheet optimization and study work needs to be done for these systems but was not undertaken in the current study because the LM-8000 CT is no longer proposed for commercialization by GE.

11.5.4 Other Development Issues

Several other development issues currently limit the commercial availability of the systems analyzed in the study. A major issue is the processing and disposal of solid wastes from fluidized-bed gasification systems. While such wastes may meet current disposal criteria, it is not clear that this will be true in the future. In particular, the ability to attain high conversion rates of any sulfides in such ash is an important goal in order to ensure that fluidized-bed systems can be widely commercialized.

12.0 CONCLUSIONS AND RECOMMENDATIONS

12.1 STUDY APPROACH AND LIMITATIONS

This study has investigated a number of gasification power plant concepts based on air-blown gasification with hot-gas cleanup. Both IGCC and ISTIG cycles were investigated, with variations in plant size, gasification technology, and gas cleanup method. All plants were designed and costed for a specific site in The Southern Company, namely Georgia Power Company's Plant Wansley, near Atlanta, Georgia. From The Southern Company's perspective, the primary motivation for the study was to determine how air-blown gasification with hot-gas cleanup compares to both PC with FGD and oxygen-blown IGCC with cold-gas cleanup for baseload power production in The Southern Company. Important aspects of this comparison include capital cost, efficiency, operability, reliability, and environmental performance.

This study should be viewed as a follow-on to an investigation of oxygen-blown IGCC at Plant Wansley conducted by SCS and EPRI in 1989 (Reference 2). While every effort was made to conduct the studies on a comparable basis, technology changes since the conclusion of the EPRI study made it necessary to update the performance and costs of the EPRI IGCC case. Case 1 of the current study is considered to be most directly comparable to the updated EPRI case and has been used to analyze and point out essential differences in cost and performance between air-blown gasification with hot-gas cleanup versus oxygen-blown gasification with conventional cold cleanup. Both plants were sized at nominally 400 MW, with power islands comprised of two GE 7F gas turbines and one reheat steam turbine. The comparison of these plant designs, along with the other plant concepts presented in the study, has provided important information with regard to costs, performance, scale effects, technology options, and development needs. These comparisons are considered to be very accurate on a relative basis because they were developed using a common technical approach and the same key personnel.

The basis of comparison for the PC plant is an 811-MW plant design similar to the latest units added to Georgia Power Company's system, updated with the necessary environmental control improvements. According to the current integrated resource plan for The Southern Company, such a unit is the most likely candidate for the next baseload power plant, which would be commissioned in 2006. In this study, the comparison between the Case 1 IGCC plant and the PC plant was made specifically for The Southern Company system using a revenue requirements analysis. This necessitated applying scaling factors to the IGCC costs in order to put the IGCC and PC plants on a comparable size basis. Limitations of the IGCC versus PC comparison are these: (1) the PC plant costs were not developed as part of the current study (although the PC costs were

consulted for input), and thus need to be updated and reviewed to ensure comparability; and (2) the use of scaling factors to adjust for substantial differences in plant size is expected to reduce the accuracy of the comparisons. Development of IGCC plant costs for an 800-MW facility should be undertaken to improve confidence in the results.

12.2 PERFORMANCE AND COST COMPARISONS

Table 12-1 provides a comparison of the performance of the six cases analyzed in the study as well as the updated EPRI reference case. Table 12-2 provides a summary of capital costs, O&M costs, and calculated busbar costs for the same cases. Conclusions regarding these study results are discussed in the following sections.

12.2.1 IGCC Case Results

The features of the IGCC cases (1, 3, 5, and 6) are given in Table 12-1. Among the DOE cases, the best performance was achieved by the transport IGCC (Case 6), which uses a hybrid cycle. The heat rate of 8153 Btu/kWh (HHV) for this case would be better than virtually any other coal-fired plant (electric power only) currently in existence or planned for the next several years. This performance compares to heat rates ranging from 8595 to 9357 Btu/kWh for the other IGCC cases, and 9039 Btu/kWh for the EPRI oxygen-blown design. The substantial performance advantage of this case over the KRW-based designs is attributable in large measure to the high cold-gas efficiency predicted for the transport gasifier (82.1 percent versus 70 percent), which is due in turn to the good gasification kinetics and the low gasification temperature, and to the elimination of the hot-gas cleanup processes. While the transport IGCC case appears to be very attractive for future consideration, this case is by far the least developed of the IGCC technologies. Although the transport gasifier is conceptual in status, the potential reductions in costs and heat rate justify further testing at a proof-of-concept scale to demonstrate the technology.

Among the KRW-based IGCC cases, Case 1 has the best performance, with an efficiency advantage of approximately 5 percent over the oxygen-blown reference case. This improvement is due in part to the better energy utilization of a hot-gas cleanup system. However, as pointed out in the Second Law analysis of Section 10, the reference case efficiency is adversely affected by the lower total carbon conversion of that design, which does not have a sulfator as in the hot-gas cleanup cases. The Second Law analysis indicated that both cases could be improved by further optimization.

With regard to the relative costs of the IGCC cases (Table 11-2), transport IGCC again has a substantial advantage due to both high efficiency and low capital cost. The capital cost advantage, like the high efficiency, stems from superior

Table 12-1
Summary of Gasification Power Plant Performance

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	OXYGEN BLOWN *
NOMINAL PLANT SIZE (MW)	400	100	200	400	400	400	400
GASIFIER TYPE	KRW	KRW	KRW	KRW	KRW	TRANS.	KRW
CYCLE TYPE	IGCC	ISTIG	IGCC	ISTIG	IGCC	IGCC	IGCC
BULK SULFUR REMOVAL	IN-BED	IN-BED	IN-BED	IN-BED	ZnFe	IN-BED	SELEXOL
<u>GASIFIER FEEDS:</u>							
COAL (T/DAY) (as prepared)	3,792	892	1,896	3,567	3,783	3,213	3,845
LIMESTONE (T/DAY)	1,053	248	526	990	---	431	---
AIR (T/DAY)	12,888	2,984	6,444	11,936	13,274	11,334	---
OXYGEN (T/DAY)	---	---	---	---	---	---	2,347
<u>GASIFICATION ISLAND PRODUCTS:</u>							
FUEL GAS (LHV, MMBTU/DAY)	63,336	15,096	31,668	60,384	62,184	59,757	73,653
STEAM TO POWER ISLAND (MLB/DAY)							
1600 psia (sat.) (1500 psia for Oxy. Blown)	12,447	---	6,224	---	12,824	6,160	10,555
850 psia (sat.)	---	949	---	3,795	---	---	---
SOLID WASTE (T/DAY)	1,231	289	615	1,158	540	797	492
SULFURIC ACID (T/DAY)	---	---	---	---	280	---	---
SULFUR (T/DAY)	---	---	---	---	---	---	78
<u>GASIFICATION ISLAND PERFORMANCE:</u>							
COLD GAS EFFICIENCY - LHV (%)	69.7	70.6	69.7	70.6	68.6	77.6	82.1
CARBON CONVERSION (%)							
GASIFIER ONLY	96.5	97.1	96.5	97.1	95.2	98.2	95.8
GASIFICATION ISLAND	99.9	99.9	99.9	99.9	95.2	99.9	95.8
SULFUR REMOVAL (%)							
GASIFIER ONLY	86.5	86.5	86.5	86.5	----	---	---
GASIFICATION ISLAND	99.4	99.4	99.4	99.4	99.4	99.2	96.4
<u>R-A-M ANALYSIS:</u>							
TOTAL PLANT EQ. AVAIL. FACTOR, % ***	84.4	84.2	83.0	86.1	84.0	85.3	85.2
GAS ISLAND EQ. AVAIL. FACTOR, % ****	94.8	93.4	93.5	95.2	94.4	96.0	N/A
<u>OVERALL PLANT PERFORMANCE:</u>							
COAL ENERGY INPUT (MMBTU/HR)	3,940	927	1,970	3,706	3,932	3,339	3,737
GROSS POWER OUTPUT (MW)							
COMBUSTION TURBINE(S)	311.6	104.7	155.8	419.0	306.6	263.2	298.8
STEAM TURBINE(S)	182.2	---	94.0	---	153.4	178.4	159.4
TOTAL POWER	493.8	104.7	249.8	419.0	459.0	441.6	458.2
STATION SERVICE (MW)	35.4	5.2	18.2	19.9	38.8	32.1	44.7
NET POWER OUTPUT (MW)	458.4	99.6	231.7	399.1	420.2	409.5	413.5
NET HEAT RATE - HHV (BTU/KWH)	8595	9310	8504	9287	9357	8153	9039
THERMAL EFFICIENCY - HHV (%)	39.7	36.6	40.1	36.7	36.5	41.8	37.7

* Updated design based on BCS study of KRW-based IGCC, Electric Power Research Institute Research Project 2773-6

** 380 psig steam supply pressure is adequate for Case 6.

*** Includes planned overhaul outage of 412 hours/year.

**** Does not include planned overhaul outage.

Table 12-2
Summary of Gasification Power Plant Costs

	ALL COSTS IN MID-1990 DOLLARS						OXYGEN BLOWN
	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	
NET POWER OUTPUT, MW	458.4	89.6	231.7	399.1	420.2	409.5	413.5
<u>CAPITAL COST *</u>	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW	\$MM \$/kW
PLANT CONSTRUCTION COST	378 825	120 1206	219 944	356 893	412 982	329 803	405 960
PROCESS CONTINGENCY	20 43	6 61	10 45	22 55	24 58	17 42	15 36
PROJECT CONTINGENCY	80 174	25 254	46 197	76 190	87 208	69 169	84 203
TOTAL PLANT COST	477 1042	151 1521	275 1188	454 1138	524 1247	415 1014	504 1218
AFUDC	94 204	29 280	54 226	89 223	103 245	82 199	99 239
TOTAL PLANT INVESTMENT	571 1246	180 1811	329 1413	543 1362	627 1492	497 1213	602 1457
OWNER'S COSTS	44 96	14 136	25 107	40 99	50 120	27 66	37 90
TOTAL CAPITAL REQ'T	615 1342	194 1947	354 1520	583 1461	677 1612	524 1279	640 1547
<u>O&M COST **</u>							
FIXED O&M COSTS, \$/kW-YR	38.3	77.8	47.7	42.2	45.9	37.2	43.9
VARIABLE O&M, mills/kWh	2.41	2.63	2.43	2.45	3.55	1.05	0.99
FUEL COSTS, mills/kWh	12.9	14.0	12.8	13.9	14.0	12.2	13.6
<u>COST OF ELECTRICITY ***</u>							
CURRENT \$, mills/kWh	67.4	95.5	74.5	73.1	79.9	62.8	73.8
CONSTANT \$, mills/kWh	49.1	68.4	54.2	53.2	56.1	45.6	53.4

* PLANT CONSTRUCTION COST = PROCESS PLANT COST & GENERAL FACILITIES + ENGINEERING FEES

** FIRST-YEAR O&M COST

*** TEN-YEAR LEVELIZED COST AT 65% CAPACITY FACTOR

gasification and desulfurization performance assumed for the transport reactor, which reduces reactor size and eliminates the need for external desulfurization. Comparing Case 1 with the reference case, air-blown gasification with hot-gas cleanup is projected to have a 9-percent advantage in cost of electricity over oxygen-blown IGCC. An important conclusion of this study is that the major advantage of the Case 1 plant may be a substantial reduction in capital cost, amounting to 13 percent versus the oxygen-blown plant. As discussed in Section 11, a true capital cost savings would probably be of greater importance in The Southern Company than operating cost advantages. In particular, operating cost advantages arising from byproduct sales, as in the oxygen-blown reference case and DOE Case 5, would not be viewed as favorably as capital-derived savings because of market risks and added operating complexity. Based on the current study, full hot-gas cleanup using fixed beds of zinc ferrite sorbent coupled with sulfuric acid production (Case 5) would not appear to be an attractive IGCC alternative. Complete reliance on a zinc ferrite desulfurization system would not be acceptable to The Southern Company at the current level of development. Moreover, the higher capital cost and overall plant complexity for this case do not fit well with The Southern Company plant profile as developed in Section 11. Potential improvements in this technology, some of which are outlined in Appendix B, must be pursued if this plant design is to gain widespread acceptance.

With regard to scale economies, Cases 1 and 3 provide an excellent indication of how plant scale affects both capital and operating costs, since Case 3 is essentially a half-scale version of Case 1. The conclusion from this analysis is that the capital cost scale economies over this size range are rather low (exponent of approximately 0.8 based on megawatts), which is a reasonable conclusion for highly modular technologies. Nonetheless, smaller plants would not be of significant interest for The Southern Company because of the large growth increments of the 30,000-MW system.

12.2.2 ISTIG Case Results

The ISTIG cases developed in the project were of more interest to DOE than to SCS. The advantages cited for this cycle include higher efficiencies and reduced capital costs due to elimination of the steam turbine cycle. These advantages were not seen in the current study, as noted in Tables 12-1 and 12-2. The ISTIG cases were less efficient than expected because of a limitation in the ability of the LM-8000 aeroderivative gas turbine to accommodate the full steam raising potential of the gasification island. It is not clear how this limitation could be overcome with the equipment specified in the study.

From the standpoint of SCS, the ISTIG plant concept is also unattractive at the present time because of lack of development support from the private sector. The LM-8000-based ISTIG is no longer supported by GE, which is now proposing a

new ISTIG concept. In any case, GE has indicated in the past that it will not pursue development of these systems unless a threshold customer base is assured. SCS does not intend to pursue further studies of the ISTIG concept unless further progress toward commercialization is made. A potentially more attractive development along these lines is the Humid Air Turbine (HAT) cycle currently under study by EPRI.

12.3 COST SENSITIVITY STUDIES

A cost sensitivity analysis was performed to identify potential risk factors in the costs of air-blown gasification power plants. Cost risks investigated included those arising from increases in capital cost, fuel costs, plant capacity factor, plant thermal efficiency, and non-fuel O&M costs. Based on this analysis, plant capacity factor was found to be the most important variable in realizing a cost-effective plant. The heat rate advantages of the best cases in the study should ensure that these plants would dispatch at full availability in The Southern Company system. This suggests that a major emphasis should be placed on verifying the relatively high reliability predicted for air-blown gasification systems with hot-gas cleanup, and improving reliability as much as possible without increasing capital cost.

The next most important factors in cost uncertainty are capital cost and thermal efficiency. Both factors are substantially more important to the attractiveness of the study cases than either fuel costs or non-fuel O&M costs. Again, further cycle optimization and commercial demonstrations must be pursued to minimize uncertainties with regard to capital costs and plant performance of these plant concepts.

12.4 SECOND LAW ANALYSIS OF IGCC CONFIGURATIONS

A detailed Second Law analysis and comparison was conducted for two 400-MW IGCC designs: the air-blown system with hot-gas cleanup (Case 1) versus the oxygen-blown system with cold-gas cleanup (modified EPRI reference case). This analysis indicated that although the oxygen-blown KRW gasifier has a higher overall efficiency, this advantage is offset by lower efficiency in the gas turbine compared to Case 1. Both cases were found to be in need of further optimization. Several design changes were identified for the air-blown IGCC plant to improve thermal efficiency in a cost-effective way. These design changes include:

- Operating the gasifier in the temperature range of 1850 to 1870°F versus 1900°F as designed.
- Redesigning the recycle gas cooler, IP economizer, and IP drum to avoid heat transfer through the pinch point.

- Studying the economic feasibility of developing a gas turbine system with a reheat stage.
- Exploring further the option of utilizing heat recovered in the sulfation step for power production rather than coal drying.

SCS recommends that these and other design improvements be investigated further to determine their feasibility and impact on the capital and operating costs for the air-blown IGCC plant with hot-gas cleanup.

12.5 SYSTEM STUDY OF AIR-BLOWN IGCC WITH HOT GAS CLEANUP VERSUS A PULVERIZED COAL UNIT WITH FGD

A Southern Company production cost analysis was performed to compare the difference in system revenue requirements associated with the installation at Plant Wansley of 811 MW of IGCC capacity based on the Case 1 design versus an 811-MW PC unit with FGD. Either plant would be commissioned in the year 2006 in keeping with the system integrated resource plan. Cost differentials included costs due to both capital investment and operating expenses over a 45-year plant life.

This analysis showed 1990 present worth savings of over \$600 million if the air-blown IGCC unit were installed rather than the PC unit. Most of the savings were attributed to reduced system operating expenses, primarily due to lower fuel costs resulting from the higher efficiency of the IGCC plant. However, approximately \$150 million (1990 present worth) in capital cost savings were realized because of the lower capital cost and shorter lead time of the IGCC plant.

The conclusion from this analysis is that air-blown gasification with hot-gas cleanup appears to have a large economic advantage over a PC plant with FGD. The predicted economic advantage depends primarily on the thermal efficiency of the IGCC plant, which can be predicted with greater confidence than relative capital costs, which would appear to be roughly the same for both plant types, within the limits of estimating error. IGCC based on the transport reactor in a hybrid system (Case 6) would offer even greater cost savings than Case 1.

To better define the economic advantages of IGCC systems over PC plants, additional system analyses should be done addressing the following points:

- The SCS PC capital costs should be compared in detail to IGCC costs to make sure that the estimates are comparable. The capital costs of a full 811-MW IGCC unit should be developed rather than relying on scaling factors.

- Sensitivity analyses should be performed with respect to forced outage rate, planned maintenance outage hours, and capital cost for the IGCC plant of Case 1.
- System production cost analyses should be extended to include the modified IGCC reference case, the transport IGCC case (Case 6), and an improved KRW system based on Case 1.
- Comparison of various IGCC cases with a supercritical PC case should be made.

12.6 ENVIRONMENTAL PERFORMANCE OF AIR-BLOWN GASIFICATION POWER PLANTS WITH HOT-GAS CLEANUP

Environmental performance and environmental issues were evaluated for the study cases. Environmental performance of air-blown, hot-gas cleanup systems is predicted to be superior to PC plants with FGD with respect to SO_2 , NO_x , particulates, and CO_2 . IGCC with hot-gas cleanup would be comparable to PC in NO_x emissions if SCR were installed on both, but the costs of SCR would be much cheaper for the gasification-based systems. The gasification systems are expected to be equal to or better than PC plants with respect to solid waste and water discharges. An exception might be in the area of solid waste if sulfides in the waste cannot be readily oxidized. Gasification power plants with hot-gas cleanup would probably have higher emissions of certain of the most volatile "air toxic" compounds, notably mercury and chlorides, compared to a PC unit with a scrubber. On the whole, it is concluded that the gasification-based plants addressed in the study would, based on current environmental criteria, be evaluated as significantly superior to PC with FGD with regard to environmental performance.

It is recommended, however, that additional work be pursued on the emission of air toxic compounds from gasification plants with hot-gas cleanup. Further work on solid wastes from fluidized-bed gasification should also be pursued. The planned Wilsonville hot-gas cleanup test facility (DOE project DE-FC21-90MC25140) would provide an excellent and unique opportunity to conduct investigations in these areas.

IGCC with hot-gas cleanup is not as environmentally superior as IGCC with cold-gas cleanup due to the gas scrubbing that occurs in the latter design. With the exception of SO_2 , hot-gas cleanup systems would produce higher emissions of most pollutants, including potential air toxics, than systems based on cold-gas cleanup. However, hot-gas cleanup systems do substantially reduce wastewater treatment requirements, which are quite extensive for conventional IGCC systems. Based on environmental standards and requirements in The Southern Company service area, the environmental performance of IGCC with hot-gas cleanup should be

readily acceptable unless air toxics or solid waste regulations become more stringent.

12.7 IMPLEMENTATION OF IGCC WITH HOT-GAS CLEANUP FOR THE SOUTHERN COMPANY

In Section 11, a detailed analysis of the selection process and criteria for The Southern Company baseload power generation additions was presented. Based on this analysis, air-blown IGCC with hot-gas cleanup would appear to have many of the features that would favor implementation of this technology, including lower capital costs, inherently high thermal efficiency, shorter lead times, and emphasis on dry cleanup and solids handling over refinery-type gas cleaning operations. Features often cited for conventional IGCC plants such as lower space requirements, reduced solid waste volume, byproduct production, and highest environmental performance would not appear to outweigh the advantages of IGCC with hot-gas cleanup. However, special advantages of oxygen-blown IGCC with cold cleanup such as compatibility with liquid fuels production could favor this type of system.

The implementation of a new baseload technology such as IGCC by 2006 requires an accelerated schedule of R&D, technology transfer, and preliminary engineering activities in the immediate future. It is essential that IGCC with hot-gas cleanup be demonstrated at utility scale within the 1990s to allow this technology to be considered for the next baseload unit in The Southern Company. Barrier issues discussed in Section 11 should be addressed in near-term development and commercialization plans.

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LIST OF ACRONYMS AND ABBREVIATIONS

A	Availability
ABD	Availability Block Diagrams
AFUDC	Allowance for Funds During Construction
APHA	American Public Health Association
ASPEN	Computer Simulation Model
BACT	Best Available Control Technology
BOD	Biological Oxygen Demand
CDP	Compressor Discharge Bleedair Port
COD	Chemical Oxygen Demand
COE	Cost of Electricity
CT	Combustion Turbine
CTG	Combustion Turbine Generator
DOE	U.S. Department of Energy
DOE/METC	U.S. Department of Energy/Morgantown Energy Technology Center
EA	Equivalent Availability
ED	Exergy Destruction
EERC	Energy and Environmental Research Center
EFOH	Equivalent Forced Outage Hours
EFOR	Equivalent Forced Outage Rate
EP	Extraction Procedure
EPA	U.S. Environmental Protection Agency
EPRI	Electric Power Research Institute
FGD	Flue Gas Desulfurization
FICA	Federal Insurance Corporation of America
FOR	Forced Outage Rate
GE	General Electric Company
GNPIP	Gross National Product Implicit Price Deflator
GPC	Georgia Power Company

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

HAT	Humid Air Turbine
HHV	Higher Heating Valve
HP	High Pressure
HPC	High-Pressure Compressor
HRSG	Heat Recovery Steam Generator
IGCC	Integrated Gasification Combined Cycle
IP	Intermediate Pressure
IP/LP	Intermediate Pressure/Low Pressure
ISCST	Industrial Source Complex Short Term
ISTIG	Intercooled Steam-Injected Gas Turbine
JTU	Jackson Turbidity Units
KRW	KRW Energy Systems
LASH	Limestone-Containing Ash
LHV	Lower Heating Valve
LP	Low Pressure
MDT	Mean Downtime
MTBF	Mean Time Between Failure
MTR	Mean Time To Repair
MWK	The M.W. Kellogg Company
NAAQS	National Ambient Air Quality Standards
NSPS	New Source Performance Standards
O&M	Operation and Maintenance
PC	Pulverized-Coal
PFBC	Pressurized-Fluidized-Bed Combustion
POF	Planned Outage Factor
PPC	Process Plant Cost
PROMOD	Computer Simulation Model
PROVAL	Cost Analysis and Financial Computer Software Application
PSD	Prevention of Significant Deterioration

R&EA	Research and Environmental Affairs
RAM	Reliability-Availability-Maintainability
RCRA	Resource Conservation Recovery Act
R.G.	Recycle Gas
SCOT	Tailgas Treatment System
SCR	Selective Catalytic Reduction
SCS	Southern Company Services, Inc.
STG	Steam Turbine-Generator
SYNTHA	Steam Cycle Simulation Model
TAG	EPRI Technology Assessment Guide
TCLP	Toxicity Characteristic Leaching Procedure
TCR	Total Capital Requirement
TPC	Total Plant Cost
TPI	Total Plant Investment
TSS	Total Suspended Solids
TTU	Tennessee Technological University
UNIRAM	Software Application for Reliability, Availability, and Maintainability

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**ASSESSMENT OF COAL GASIFICATION/HOT GAS
CLEANUP BASED ADVANCED GAS TURBINE SYSTEMS**

**FINAL REPORT
APPENDIXES**

December 1990

Work Performed Under Contract No. DE-FC21-89MC26019

For

**U.S. Department of Energy
Office of Fossil Energy**

**Morgantown Energy Technology Center
Morgantown, West Virginia**

By

**Southern Company Services, Inc.
Birmingham, Alabama**

**The M.W. Kellogg Company
Houston, Texas**

**Tennessee Technological University Center for Electric Power
Cookeville, Tennessee**

**General Electric Company
Schenectady, New York**

APPENDIX A
GASIFICATION ISLAND EQUIPMENT DETAILS

Table A-1 (Sheet 1 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Cases 1 & 3

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
220B	STARTUP HEATER	RADIANT HEATER, INTERMITTENT SERVICE; 20 MMBTU/HR	750	410
300C	PRODUCT GAS COOLER	FIRE TUBE EXCHANGER DUTY = 102 MMBTU/HR	1,000	400
320D	GASIFIER	12.5 FT ID, 130 FT TALL, REFRACTORY LINED VESSEL	1,000	450
302F	PRESSURIZATION AIR RECEIVER	VOLUME = 1603 CU FT	105	600
310F	FEED SURGE DRUM	CONICAL VESSEL WITH CONED BOTTOM VOLUME = 3,600 CU FT	200	15
311F	FEED PRESSURIZATION HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 850 CU FT	200	450
312F	FEED HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 1100 CU FT	200	450
300F	STEAM DRUM	HORIZONTAL VESSEL VOLUME = 910 CU FT	605	1,600
301J	LOCKHOPPER AIR COMPRESSOR	400 BHP 3-STAGE RECIPROCATING	300	600
310L	COAL FEED VENT FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 311F. AVERAGE VENT RATE OF 5700 #/HR.	200	450
320L	CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 320 CU FT/SEC, 95% EFFICIENCY	1,000	400
310V	COAL FEEDER	ROTARY STAR-WHEEL FEEDER 90,920 #/HR COAL RATE; 20,000 #/HR LIMESTONE RATE	200	450
330V	ASH FEEDER	ROTARY STAR-WHEEL FEEDER 23,770 #/HR LASH RATE	500	450
251C	AIR RECUPERATOR	AIR/AIR SHELL/TUBE EXCHANGER DUTY = 42 MMBTU/HR	700	450
252CA	BFV PREHEATER	SHELL/TUBE EXCHANGER DUTY = 25 MMBTU/HR	450	195
252CB	TRIM COOLER	WATER COOLED SHELL/TUBE EXCHANGER DUTY = 20 MMBTU/HR	200	185
251F	KNOCK-OUT DRUM	CYLINDRICAL VESSEL VOLUME = 2,400 CU FT	105	180
251J	BOOSTER AIR COMPRESSOR	13,000 BHP CENTRIFUGAL	340	455

Table A-1 (Sheet 2 of 3)
Plant Wansley Study
Equipment List and Descriptions
Cases 1 & 3

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
361C	RECYCLE GAS COOLER	SHELL/TUBE EXCHANGER 9.8 MWBTU/HR	505	1,650
360F	RECYCLE GAS RECEIVER	HORIZONTAL VESSEL VOLUME = 3,700 CU FT	525	750
381J	RECYCLE GAS COMPRESSOR	TWO-STAGE CENTRIFUGAL MACHINE 690 BHP	525	750
410D	CHLORIDE GUARD REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING NAHCOLITE SORBENT 12 FT ID X 46 FT TALL	1,015	365
410DA	CHLORIDE GUARD REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING NAHCOLITE SORBENT 12 FT ID X 46 FT TALL	1,015	365
421F	CYCLONE FINES HOPPER	REFRACTORY-LINED CONED BOTTOM VESSEL; VOLUME = 210 CU FT	1,015	385
422F	CYCLONE FINES DEPRESSURIZATION HOPPER	REFRACTORY-LINED CONED BOTTOM VESSEL; VOLUME = 210 CU FT	1,015	385
430F	FILTER DEPRESSURIZATION HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 400 CU FT	500	365
421L	FINES CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE = 285 CUFT/SEC, 90% EFFICIENCY	1,015	380
422L	CYCLONE HOPPER FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 422F.	1,015	380
430L	GAS FILTER	CERAMIC CANDLE FILTER; REFRACTORY LINED VESSEL CONTAINING 5 FT LONG FILTER ELEMENTS FLOW RATE = 17,900 FT/MIN	1,015	375
431L	FILTER HOPPER FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 430F.	500	365
430V	LASH CONVEYOR	WATER-COOLED SCREW CONVEYOR SOLIDS RATE = 28,500 #/HR DUTY = 0.8 MWBTU/HR	500	15

Table A-1 (Sheet 3 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Cases 1 & 3

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
510C	EXIT GAS COOLER	FIRE TUBE EXCHANGER DUTY = 41 MMBTU/HR	1,000	305
540F	GUARD DRUM	REFRACTORY-LINED CYLINDRICAL VESSEL VOLUME = 1200 CU FT	1,200	305
510D	ZINC FERRITE REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING ZINC FERRITE SORBENT 13 FT ID X 42 FT TALL	1,200	345
510DA	ZINC FERRITE REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING ZINC FERRITE SORBENT 13 FT ID X 42 FT TALL	1,200	345
541L	GUARD CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 400 CUFT/SEC	1,200	305
601B	SULFATOR	REFRACTORY-LINED ATMOSPHERIC VESSEL 31 FT ID X 50 FT TALL (CASE 1) 27 FT ID X 50 FT TALL (CASE 3)	1,000	25
601C	SULFATOR HEAT RECOVERY	SHELL AND TUBE EXCHANGER DUTY = 70 MMBTU/HR (CASE 1) DUTY = 62 MMBTU/HR (CASE 3)	1,000	20
601F	LASH RECEIVER	SURGE FOR LASH VOLUME = 820 CUFT	500	15
602F	SOLIDS DISENGAGER	FLOW= 932 CUFT/SEC (CASE 1) FLOW= 700 CUFT/SEC (CASE 3)	1,000	20
603F	SULFATOR STEAM DRUM	HORIZONTAL VESSEL VOLUME = 235 CU FT	605	1,000
601J	SULFATION AIR COMPRESSOR	SINGLE STAGE CENTRIFUGAL 820 BHP (CASE 1); 620 BHP (CASE 3)	200	22
601LV	SOLIDS FEEDER	SCREW TYPE FEEDER LASH FLOW = 38,000 #/HR (CASE 1) LASH FLOW = 28,500 #/HR (CASE 3)	500	15
602V	SULFATOR SOLIDS COOLER	WATER COOLED SCREW CONVEYOR SOLIDS RATE = 38,000 #/HR DUTY = 5 MMBTU/HR	1,000	15

Table A-2 (Sheet 1 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Cases 2 & 4

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
220B	STARTUP HEATER	RADIANT HEATER, INTERMITTENT SERVICE; 20 MMBTU/HR	750	800
320D	GASIFIER	11.5 FT (MAX) ID, 142 FT TALL, REFRACTORY LINED VESSEL	1,900	800
302F	PRESSURIZATION AIR RECEIVER	VOLUME = 930 CU FT	300	800
310F	FEED SURGE DRUM	CONICAL VESSEL WITH CONED BOTTOM VOLUME = 3,025 CU FT	200	15
311F	FEED PRESSURIZATION HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 720 CU FT	200	800
312F	FEED HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 940 CU FT	200	800
301J	LOCKHOPPER AIR COMPRESSOR	400 BHP 4-STAGE RECIPROCATING	300	800
310L	COAL FEED VENT FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 311F. AVERAGE VENT RATE OF 6000 #/HR.	300	800
320L	CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 145 CU FT/SEC, 95% EFFICIENCY	1,900	750
310V	COAL FEEDER	ROTARY STAR-WHEEL FEEDER 80,300 #/HR COAL RATE; 22,300 #/HR LIMESTONE RATE	200	750
330V	ASH FEEDER	ROTARY STAR-WHEEL FEEDER 24,950 #/HR LASH RATE	500	800
251C	AIR RECUPERATOR	AIR/AIR SHELL/TUBE EXCHANGER DUTY = 28 MMBTU/HR	750	800
252CA	BFM PREHEATER	SHELL/TUBE EXCHANGER DUTY = 6.5 MMBTU/HR	450	475
252CB	TRIM COOLER	WATER COOLED SHELL/TUBE EXCHANGER DUTY = 10.6 MMBTU/HR	200	470
251F	KNOCK-OUT DRUM	CYLINDRICAL VESSEL VOLUME = 430 CU FT	105	500
251J	BOOSTER AIR COMPRESSOR	3,400 BHP CENTRIFUGAL	370	820
301C	RECYCLE GAS COOLER	SHELL/TUBE EXCHANGER 8.6 MMBTU/HR	605	1,650
300F	RECYCLE GAS RECEIVER	HORIZONTAL VESSEL VOLUME = 1,800 CU FT	600	1,450
301J	RECYCLE GAS COMPRESSOR	TWO-STAGE CENTRIFUGAL MACHINE 280 BHP	600	1,450

Table A-2 (Sheet 2 of 3)
Plant Wansley Study
Equipment List and Descriptions
Cases 2 & 4

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
4100	CHLORIDE GUARD REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING NAHCOLITE SORBENT 10 FT ID X 30 FT TALL	1,000	715
410DA	CHLORIDE GUARD REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING NAHCOLITE SORBENT 10 FT ID X 30 FT TALL	1,000	715
421F	CYCLONE FINES HOPPER	REFRACTORY-LINED CONED BOTTOM VESSEL; VOLUME = 140 CU FT	1,015	735
422F	CYCLONE FINES DEPRESSURIZATION HOPPER	REFRACTORY-LINED CONED BOTTOM VESSEL; VOLUME = 140 CU FT	1,015	735
430F	FILTER DEPRESSURIZATION HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 400 CU FT	500	725
421L	FINES CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 125 CUFT/SEC, 90% EFFICIENCY	1,015	735
422L	CYCLONE HOPPER FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 422F.	1,015	735
430L	GAS FILTER	CERAMIC CANDLE FILTER; REFRACTORY LINED VESSEL CONTAINING 6 FT LONG FILTER ELEMENTS FLOW RATE = 7,550 FT/MIN	1,015	730
431L	FILTER HOPPER FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 430F.	500	725
430V	LASH CONVEYOR	WATER-COOLED SCREW CONVEYOR SOLIDS RATE = 23,300 #/HR DUTY = 0.4 MMBTU/HR	500	15
510C	EXIT GAS COOLER	FIRE TUBE EXCHANGER DUTY = 35 MMBTU/HR	1,000	715
540F	GUARD DRUM	REFRACTORY-LINED CYLINDRICAL VESSEL VOLUME = 1,000 CU FT	1,200	715
510D	ZINC FERRITE REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING ZINC FERRITE SORBENT 12 FT ID X 30 FT TALL	1,200	700
510DA	ZINC FERRITE REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING ZINC FERRITE SORBENT 12 FT ID X 30 FT TALL	1,200	700
541L	GUARD CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 130 CUFT/SEC	1,200	700

Table A-2 (Sheet 3 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Cases 2 & 4

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
601B	SULFATOR	REFRACTORY-LINED ATMOSPHERIC VESSEL 22 FT ID X 50 FT TALL (CASE 2) 31 FT ID X 80 FT TALL (CASE 4)	1,600	25
601C	SULFATOR HEAT RECOVERY	SHELL AND TUBE EXCHANGER DUTY = 14 MMBTU/HR (CASE 2) DUTY = 28 MMBTU/HR (CASE 4)	1,600	20
601F	LASH RECEIVER	SURGE FOR LASH VOLUME = 800 CUFT	500	15
602F	SOLIDS DISENGAGER	FLOW= 550 CUFT/SEC (CASE 2) FLOW= 1,100 CUFT/SEC (CASE 4)	1,600	20
603F	SULFATOR STEAM DRUM	HORIZONTAL VESSEL VOLUME = 300 CU FT	605	1,600
601J	SULFATION AIR COMPRESSOR	SINGLE STAGE CENTRIFUGAL 360 BHP (CASE 2); 720 BHP (CASE 4)	200	22
601LV	SOLIDS FEEDER	SCREW TYPE FEEDER LASH FLOW = 24,000 #/HR (CASE 2) LASH FLOW = 48,000 #/HR (CASE 4)	500	15
602V	SULFATOR SOLIDS COOLER	WATER COOLED SCREW CONVEYOR SOLIDS= 31000 #/HR, 0.5 MMBTU/HR (CASE 2) SOLIDS= 62000 #/HR, 1.0 MMBTU/HR (CASE 4)	1,600	15

Table A-3 (Sheet 1 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Case 5

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
220B	STARTUP HEATER	RADIANT HEATER, INTERMITTENT SERVICE; 20 MMBTU/HR	750	450
300C	PRODUCT GAS COOLER	FIRE TUBE EXCHANGER DUTY = 70.8 MMBTU/HR	1,850	400
320D	GASIFIER	12 FT (MAX) ID, 78 FT TALL, REFRACTORY LINED VESSEL	1,850	50
302F	PRESSURIZATION AIR RECEIVER	VOLUME = 930 CU FT	300	450
310F	FEED SURGE DRUM	CONICAL VESSEL WITH CONED BOTTOM VOLUME = 1,735 CU FT	200	15
311F	FEED PRESSURIZATION HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 420 CU FT	200	450
312F	FEED HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE ; VOLUME = 540 CU FT	200	450
300F	STEAM DRUM	HORIZONTAL VESSEL VOLUME = 615 CU FT	605	1,000
301J	LOCKHOPPER AIR COMPRESSOR	400 BHP 3-STAGE RECIPROCATING	350	450
310L	COAL FEED VENT FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 311F. AVERAGE VENT RATE OF 3250 #/HR.	300	450
320L	CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 144 CU FT/SEC, 95% EFFICIENCY	1,300	400
310V	COAL FEEDER	ROTARY STAR-WHEEL FEEDER 53,000 #/HR COAL RATE	200	400
330V	ASH FEEDER	ROTARY STAR-WHEEL FEEDER 4,300 #/HR ASH RATE	500	450
251C	AIR RECUPERATOR	AIR/AIR SHELL/TUBE EXCHANGER DUTY = 39 MMBTU/HR	750	450
252CA	BFW PREHEATER	SHELL/TUBE EXCHANGER DUTY = 27 MMBTU/HR	450	175
252CB	TRIM COOLER	WATER COOL SHELL/TUBE EXCHANGER DUTY = 22 MMBTU/HR	200	170
251F	KNOCK-OUT DRUM	CYLINDRICAL VESSEL VOLUME = 2,500 CU FT	105	200
251J	BOOSTER AIR COMPRESSOR	14,200 BHP CENTRIFUGAL	370	470
301C	RECYCLE GAS COOLER	SHELL/TUBE EXCHANGER 9.8 MMBTU/HR	605	1,050
300F	RECYCLE GAS RECEIVER	HORIZONTAL VESSEL VOLUME = 3,000 CU FT	600	750

Table A-3 (Sheet 2 of 3)
Plant Wansley Study
Equipment List and Descriptions
Case 5

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
381J	RECYCLE GAS COMPRESSOR	TWO-STAGE CENTRIFUGAL MACHINE 400 BHP	600	750
410D	CHLORIDE GUARD REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING MAHCOLITE SORBENT 12 FT ID X 30 FT TALL	1,000	365
410DA	CHLORIDE GUARD REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING MAHCOLITE SORBENT 12 FT ID X 30 FT TALL	1,000	365
421F	CYCLONE FINES HOPPER	REFRACTORY-LINED CONED BOTTOM VESSEL; VOLUME = 140 CU FT	1,015	385
422F	CYCLONE FINES DEPRESSURIZATION HOPPER	REFRACTORY-LINED CONED BOTTOM VESSEL; VOLUME = 140 CU FT	1,015	385
430F	FILTER DEPRESSURIZATION HOPPER	CONICAL VESSEL WITH 70 DEGREES CONE ANGLE; VOLUME = 100 CU FT	500	375
421L	FINES CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 180 CUFT/SEC, 90% EFFICIENCY	1,015	385
422L	CYCLONE HOPPER FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 422F.	1,015	385
430L	GAS FILTER	CERAMIC CANDLE FILTER; REFRACTORY LINED VESSEL CONTAINING 6 FT LONG FILTER ELEMENTS FLOW RATE = 9,050 FT ³ /MIN	1,015	380
431L	FILTER HOPPER FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 430F.	775	375
430V	LASH CONVEYOR	WATER-COOLED SCREW CONVEYOR SOLIDS RATE = 7,500 #/HR DUTY = 0.8 MMBTU/HR	775	15
510C	EXIT GAS COOLER	FIRE TUBE EXCHANGER DUTY = 25 MMBTU/HR	1,000	365
540F	GUARD DRUM	REFRACTORY-LINED CYLINDRICAL VESSEL VOLUME = 1,200 CU FT	1,200	365
510D	ZINC FERRITE REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING ZINC FERRITE SORBENT 12 FT ID X 42 FT TALL	1,200	350
510DA	ZINC FERRITE REACTOR	REFRACTORY-LINED FIXED-BED REACTOR CONTAINING ZINC FERRITE SORBENT 12 FT ID X 42 FT TALL	1,200	350
541L	GUARD CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 210 CUFT/SEC	1,200	350

Table A-3 (Sheet 3 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Case 5

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
511C	REGENERATION GAS STEAM GENERATOR	STEAM/STEAM GENERATOR DUTY = 197.6 MMBTU/HR 1 COMMON FOR 6 TRAINS	1200	450
512C	REGENERATION GAS TRIM COOLER	STEAM/BFW PREHEATER DUTY = 9.42 MMBTU/HR 1 COMMON FOR 6 TRAINS	500	450

Table A-4 (Sheet 1 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Case 6

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
220B	STARTUP HEATER	RADIANT HEATER, INTERMITTENT SERVICE; 7.2 MMBTU/HR	750	350
360C	PRODUCT GAS COOLER	FIRE TUBE EXCHANGER DUTY = 79 MMBTU/HR	1,455	300
320D	TRANSPORT GASIFIER	REFRACTORY LINED VESSEL 5.5 FT ID, 70 FT TALL	1,455	320
302F	AIR RECEIVER	VOLUME = 972 CU FT	200	600
310F	COAL FEED SURGE DRUM	VESSEL WITH CONED BOTTOM VOLUME = 4,160 CU FT	200	15
311F	COAL FEED PRESSURIZATION HOPPER	VESSEL WITH CONED BOTTOM VOLUME = 925 CU FT	200	350
312F	COAL FEED HOPPER	VESSEL WITH CONED BOTTOM VOLUME = 1,200 CU FT	200	350
313F	LIMESTONE FEED SURGE DRUM	VESSEL WITH CONED BOTTOM VOLUME = 200 CU FT	200	15
314F	LIMESTONE FEED PRESSURIZATION HOPPER	VESSEL WITH CONED BOTTOM VOLUME = 85 CU FT	200	350
315F	LIMESTONE FEED HOPPER	VESSEL WITH CONED BOTTOM VOLUME = 80 CU FT	200	350
320F	GASIFIER STANDPIPE	REFRACTORY LINED VESSEL 8 FT ID, 45 FT TALL	1,455	300
360F	STEAM DRUM	HORIZONTAL VESSEL VOLUME = 500 CU FT	605	1,600
301J	LOCKHOPPER AIR COMPRESSOR	300 BHP 3-STAGE RECIPROCATING	430	600
310L	COAL FEED VENT FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 311F. AVERAGE VENT RATE OF 3350 #/HR.	200	350
311L	LIMESTONE FEED VENT FILTER	SINTERED-METAL TYPE FILTER; CYCLIC OPERATION DURING PRESSURIZATION AND DEPRESSURIZATION OF 314F. AVERAGE VENT RATE = 225 #/HR	200	350
320LA,B	PRIMARY CYCLONES	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE= 12,050 ACFM	1,455	300
321LA,B	SECONDARY CYCLONES	REFRACTORY- LINED CYLINDRICAL VESSEL FLOW RATE = 12,050 ACFM	1,455	300
322LA,B	GASIFIER STANDPIPE CYCLONES	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE = LATER	1,455	300
310V	COAL FEEDER	ROTARY STAR-WHEEL FEEDER 110,000 #/HR COAL RATE;	200	350
311V	LIMESTONE FEEDER	ROTARY STAR-WHEEL FEEDER 15,000 #/HR LASH RATE	200	350

Table A-4 (Sheet 2 of 3)
Plant Wansley Study
Equipment List and Descriptions
Case 6

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
251C	AIR RECUPERATOR	AIR/AIR SHELL/TUBE EXCHANGER DUTY = 40 MMBTU/HR	740	350
252CA	BFV PREHEATER	SHELL/TUBE EXCHANGER DUTY = 14 MMBTU/HR	385	185
252CB	TRIM COOLER	WATER COOLED SHELL/TUBE EXCHANGER DUTY = 17 MMBTU/HR	260	185
251F	KNOCK-OUT DRUM	CYLINDRICAL VESSEL VOLUME = 2,237 CU FT	185	185
251J	BOOSTER AIR COMPRESSOR	17,500 BHP CENTRIFUGAL	290	355
301C	RECYCLE GAS COOLER	SHELL/TUBE EXCHANGER 1.6 MMBTU/HR	600	1,650
300F	RECYCLE GAS RECEIVER	HORIZONTAL VESSEL VOLUME = 1,500 CU FT	520	700
301J	RECYCLE GAS COMPRESSOR	TWO-STAGE CENTRIFUGAL MACHINE 320 BHP	520	700
430L	GAS FILTER	CERAMIC CANDLE FILTER; REFRACTORY LINED VESSEL CONTAINING 5 FT LONG FILTER ELEMENTS FLOW RATE = 19,950 ACFM	1,000	290
601B	TRANSPORT SULFATOR	REFRACTORY-LINED VESSEL 2 FT ID , 60 FT TALL	1,650	290
601CA, B	BAYONET BOILERS	REFRACTORY-LINED SHELL/TUBE EXCHANGER DUTY = 13 MMBTU/HR	1,650	290
602C	FLUID BED HEAT EXCHANGER	COILS INSIDE REFRACTORY-LINED VESSEL DUTY = 33 MMBTU/HR	1,650	290
604C	HIGH PRESSURE ASH COOLER	COILS INSIDE REFRACTORY-LINED VESSEL DUTY = 10 MMBTU/HR	1,650	290
603FA, B	ASH DEPRESSURIZATION HOPPERS	VESSEL WITH CONED BOTTOM VOLUME = 220 CU FT	330	300
604F	BLOWBACK AIR RECEIVER	VOLUME = 775 CU FT	520	700
610F	SULFATOR STEAM DRUM	HORIZONTAL VESSEL VOLUME = 290 CU FT	605	1,600
601J, JA	BLOWBACK AIR COMPRESSOR	SINGLE STAGE RECIPROCATING 20 BHP	520	700
601LA, B	SULFATOR PRIMARY CYCLONES	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE = 1,083 ACFM	1,650	290
602LA, B	SULFATOR SECONDARY CYCLONES	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE = 1,033 ACFM	1,650	290
603L	SULFATOR BARRIER FILTER	CERAMIC CANDLE FILTER; REFRACTORY-LINED VESSEL CONTAINING 5 FT. LONG FILTER ELEMENTS FLOW RATE = 2,100 ACFM	1,650	290

Table A-4 (Sheet 3 Of 3)
Plant Wansley Study
Equipment List and Descriptions
Case 6

EQUIPMENT NUMBER	EQUIPMENT NAME	DESCRIPTION AND CAPACITY PER TRAIN	OPERATING CONDITIONS (HIGHEST OPERATING)	
			TEMP, F	PRESS, PSIA
610L	FLUID BED PRIMARY CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE = 2,100 ACFM	1,650	290
611L	FLUID BED SECONDARY CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE = 2,100 ACFM	1,650	290
620L	ASH COOLER CYCLONE	REFRACTORY-LINED CYLINDRICAL VESSEL FLOW RATE = 2,100 ACFM	1,650	290
602V	ASH SCREW CONVEYOR	WATER COOLED SCREW CONVEYOR SOLIDS RATE = 29,500 #/HR DUTY = 1 MMBTU/HR	330	15

APPENDIX B
GENERAL ELECTRIC ENVIRONMENTAL SERVICES, INC.,
CONCEPT FOR
REVISED CASE 5

**A GENERAL ELECTRIC ENVIRONMENTAL SERVICES, INC.
REVIEW OF SOUTHERN COMPANY SERVICES/DOE
IGCC SYSTEMS STUDY - CASE 5**

INTRODUCTION

The Southern Company Services (SCS) IGCC Systems study has been performed to evaluate the performance and cost of coal fired power plants based on Integrated Gasification Combined Cycle (IGCC) power plants utilizing air blown, fluidized bed gasifiers with either internal or external desulfurization of the fuel gas. An oxygen blown case with external, cold gas cleanup has been added for comparison. The gasifier technology has been based on KRW designs with the exception of Case 6 which utilizes an advanced fluid bed concept. Cases 1 through 6 all utilized high temperature fuel gas desulfurization with zinc ferrite as the regenerable mixed metal oxide sulfur "sorbent" material in fixed bed reactors. Cases 1, 2, 3, 4, and 6 employ the high temperature desulfurization process in a polishing mode with the majority of the sulfur removal being accomplished within the gasification process through limestone addition to the coal feed. Case 5 has been constructed around totally external desulfurization.

Case 1 is most directly comparable to Case 5, both cases being nominal 400 MWe configurations utilizing air blown KRW gasification technology, the primary discriminant being predominantly internal (Case 1) versus totally external desulfurization. In Case 1, four (4) gasifier/HGCU trains are used to fuel two (2) parallel gas turbine combined cycles to produce a net output of 458.4 MWe. In addition to limestone addition to the gasifier, the external HGCU system contains cyclones and ceramic filters for particulate removal, fixed bed reactors containing Nahcolite for halogen removal and two (2) fixed bed reactors per gasifier train for sulfur removal and sorbent regeneration.

Case 5 utilizes six (6) gasifier/HGCU trains to fuel two (2) parallel gas turbine combined cycles to produce a net output of 420.2 MWe. All sulfur removal is accomplished external to the gasifiers utilizing two (2) fixed bed reactors per gasifier train for absorption and regeneration duty. Fixed bed reactors containing Nahcolite are also used for halogen removal and cyclones and ceramic filters for particulate removal. In both Cases 1 and 5, high pressure, high temperature steam is used as a once-through diluent for temperature control in the highly exothermic regeneration process. In Case 1 as well as Cases 2, 3, 4, and 6) the regeneration off-gas which is largely steam, N_2 , and SO_2 , is recycled to the gasifier to allow further in-bed sulfur capture with the ultimate fate of the sulfur being $CaSO_4$ following further external roasting of the gasifier ash/slag. Another option was chosen for Case 5 in which the time varying regeneration off-gas flow was directed to a commercial sulfuric acid plant with appropriate upstream modification.

As a consequence of predetermined system study rules and assumptions, GEESI has concluded that Case 5 has been penalized with regard to performance and cost and that substitution of GEESI's moving bed HGCU system for the two vessel, fixed bed system utilized in the study could offer substantial improvements. Since limited time and level of detail of the information provided by Southern Company Services precluded a complete analysis of the overall system performance and its basis for cost data, GEESI has examined the consequences of retrofitting its moving bed HGCU system within the existing overall system's process configuration while only making changes in process or equipment cost where the changes were decoupled or resulted in elimination with respect to the existing study. The changes and comments have been grouped by major process areas and the overall effect of the changes is presented in Table 1 as Cases 5' and 5" relative to Cases 1 and 5 of the current study. Figures 1 and 2 have been included to illustrate the major process and equipment differences between the HGCU and sulfuric acid plant configurations being considered.

GASIFICATION

Six (6) KRW fluid bed gasifiers have been required to provide fuel for the two (2) gas turbines of Case 5 while four (4) gasifiers are required for Case 1. The reason for the extra capacity was stated as a requirement for lower superficial gasification velocity due to the lack of limestone in the fluid bed. GEESI has pointed out that sized inert material such as sand could be used to restore bed density and allow increased velocity reducing the number of gasifiers to four (4). This presumably would have an effect on capital cost and, more important, would reduce the requirement for HGCU system trains from six (6) to four (4). As shown in Figure 1, such a reduction would have a significant cost impact due to the high temperature valve and piping differences between the fixed and moving bed system. However, since the choice of gasifier superficial velocity derives from another vendor (KRW), GEESI has not taken any credit for cost in this area in the construction of its alternate cases (5' and 5") and the six (6) flow train configuration has been retained.

CHLORIDE REMOVAL

It has been assumed in the current SCS systems study that chloride removal will be required, upstream of the zinc ferrite absorption system, for the purpose of prevention of formation of zinc chloride. This has been accomplished through the use of two (2) fixed bed reactors per gasifier flow train with non-regenerable consumption of Nahcolite and hot flow alternation between vessel. A capital cost for this system has been given in the study (Table 3.6) as \$14.7M, presumably based on Case 1 which requires only four (4) flow trains and eight (8) chloride vessel reactors and associated valving.

GEESI moving bed HGCU system, currently under demonstration testing, has incorporated a proprietary (U. S. Patent applied for) halogen removal concept which utilizes the existing sulfur removal system equipment and will require only inert, low pressure storage and metering of Nahcolite. The additional equipment costs are projected to be less than \$0.5M at the nominal 400 MWe power plant size being studied. It is likely that the capital cost for the

SCS study's chloride guard shown for Case 1 as \$ 14.7M was significantly higher for the six (6) train arrangement of Case 5 even though the fuel gas flows are similar for the two cases. There was not sufficient information available for GEESI to determine allocation of costs for process area 400 for Case 5 so a conservative approach was taken in constructing Case 5' by deleting the \$14.7M capital cost associated with the four (4) train system of Case 1 and ignoring the relatively small added cost (\$ $\leq$ 0.5M) for halogen removal in the moving bed system.

ZINC FERRITE BASED FUEL GAS DESULFURIZATION

The basic building blocks for the fixed bed and moving bed HGCU systems are shown in Figure 1. Apart from differences in total capital requirement for the two configurations, the selection of the steam regenerated, fixed bed, zinc ferrite system predetermines some operational and system performance limitations. The current SCS study has prescribed gasification, absorption and regeneration system operation at 450 psia, presumably set by gasifier size and performance optimization. Further, high pressure, high temperature steam that could be used for power generation is taken from the power island for regeneration. The SCS study requires 328,000 lb/hr of superheated steam at 450 psia and 950 F to be utilized for dilution of the regeneration air flow for temperature control. This steam appears to have been throttled from the 1605 psi drum pressure level and would be available to the steam turbine cycle at approximately 1525 psia and 950 F ($h = 1460$ B/lb). Assuming steam turbine expansion to 2.5 in. Hg. abs (1.23 psia) at a turbine efficiency of 0.85, this steam could generate 28.83 MWe. It should be noted that steam diluted regeneration is also required for Case 1 and accounts for an equivalent power generation loss of approximately 4 MWe.

The moving bed HGCU system utilizes regeneration off-gas as a diluent and requires a recycle compressor. For the coal flow and coal sulfur content shown for Case 5, the sulfur flow through the system is 8459 lb/hr requiring 60,266 lb/hr of air on stoichiometric basis. In the GEESI process, regeneration would have been accomplished at a significantly lower pressure; most likely 150 psia or less. There are several reasons for this. First,

zinc sulfate formulation is undesirably increased at higher pressure. Second, the regeneration vessel, valves and piping are dedicated to regeneration and do not have to be designed for fuel gas absorber pressure. Additionally, the recycle heat recovery steam generator, heat exchangers and recycle compressor can be cost-optimized at lower pressure and air compression work is reduced. If the GEESI process recycle system is configured for 150 psia operation with a recycle compressor flow of 348,337 lb/hr ($N_2 + SO_2$), a pressure loss of 7 psi and an inlet temperature of 150 F, the required compressor power will be less than 0.3 MWe.

Regeneration air is provided for the SCS study's fixed bed HGCU process (Cases 1 and 5) at 450 psia. This is gas turbine compressor extraction flow which has gone through additional boost compression with intercooling. Since the GEESI process only requires air at 150 psia and 200 F, this would be supplied directly from 200 psia compressor discharge. Boost compressor inlet temperature is likely to be 300 F or lower in the SCS study based on recuperative heat exchange, feed water heating and "trim cooling" considerations. If the 300 F inlet temperature is used, then the avoided air compression power for regeneration air due to use of the GEESI moving bed system will be 1.2 MWe.

Finally, the heat of combustion of the sulfur in going from the zinc and iron sulfides to SO_2 is available to the steam bottoming cycle with the GEESI process rather than being discarded with the steam diluent as in Case 5 of the SCS study. This has the effect of adding approximately 3.0 MWe to the net plant output using a 30 percent steam cycle efficiency.

TAIL GAS TREATMENT - SULFURIC ACID

Sulfuric acid has been selected as the final by-product of the external, fixed bed HGCU process for Case 5 of the SCS study. GEESI has favored this tail gas treatment concept for the moving bed system, primarily because of ease of direct integration with a significantly simplified and low cost sulfuric acid plant. A revised acid plant configuration with respect to the SCS study's design is shown in Figure 2. Since the GEESI process can provide an 800 F,

dry, continuous stream of approximately 12 percent SO_2 in N_2 , it is possible to present this gas stream directly to the converter without preconditioning or flow smoothing. Added sulfur combustion is not required nor is an oxygen plant as shown in the SCS study. Removal of the oxygen plant requirement allows a further increase in net station power of 1.5 MWe based on the study data presented for Case 5. The cost for the study's acid plant is given as \$27,048,000 with zero process contingency. Applying the 20 percent project contingency and a 19.6 percent AFUDC gives a Total Plant Investment line of \$38,826,268. Startup costs (2 percent) and a pro rata working capital amount of \$748,493 brings the Total Capital Requirement for the study's sulfuric acid plant to \$40,350,205. It is assumed that the cost of the oxygen plant has been included in this total.

It is GEESI's opinion, based on prior analyses for similar studies, that a fully integrated sulfuric acid plant operating downstream of the moving bed HGCU system would require approximately one-half of the \$40,350,205 Total Capital Requirement or approximately \$20M. This corresponds to \$44/KWe for the nominal 450 MWe power plant and compares favorably with a value of \$82.4/KWe (escalated from \$67.8/KWe in 1986 dollars) for the 50 MWe scale, 99.7 percent conversion acid plant utilized in prior GEESI studies. A scaling exponent of 0.76 provides the cost ratio of 0.53 for this plant size ratio of 9.0 and is probably conservative as sulfuric acid plants are highly scale sensitive with respect to capital cost.

SUMMARY AND CONCLUSIONS

GEESI has attempted to provide an alternate to the SCS study's Case 5 which incorporates the moving bed HGCU system and to assess net power output improvements and capital cost reductions associated with the substitution. To the extent allowed by the process flow and system cost and performance data supplied by Southern Company Services, GEESI has made a quantitative analysis of Case 5 and generated Case 5'. As can be seen from Table 1, Case 5' is a significant improvement over Case 5 in both cost and performance and the primary factors for these improvements are; use of regeneration recycle gas instead of process steam for a regeneration diluent, incorporation of the heat

of combustion of the sulfur in the coal in the coal in the steam bottoming cycle, elimination of a separate chloride removal subsystem (through incorporation in the moving bed HGCU system), simplification of the sulfuric acid plant for tail gas treatment and reduction of the regeneration pressure from 450 psia to 150 psia.

There are further benefits to be incorporated through the use of the moving bed system which GEESI has not been able to fully quantify in the manner used in generating Case 5'. Figure 1 has been provided to help in the comparison of the two systems. The two systems shown in the figure are the basic HGCU desulfurization subsystems that are associated with each single gasifier flow train. The fixed bed system differs from that shown in the SCS study's process flow sheets as it recognizes the need for double block and bleed valving for fuel gas isolation as well as the smaller, but necessary, regeneration gas valving. It is GEESI's opinion that the SCS study, in its current form, may have underestimated the cost of the valving and piping required for operation of such a fixed bed system.

The moving bed system will not require large diameter gas stream valves and will utilize 8 inch solids flow valving in its full scale form. Double block and bleed valving around the fuel gas containing vessels is shown for the solids flow system although it can be argued that the nitrogen purged lockhoppers provide sufficient isolation. GEESI's current process development unit in Schenectady employs single, 6 inch ball valves in these locations. It can also be noted that the large fuel gas piping is minimized in the moving bed system and that the regeneration system operates at only 150 psia (or lower).

Prior systems study comparisons made by GEESI and GE-CRD between the fixed and moving bed systems have resulted in a Total Capital Requirement cost ratio of approximately 0.6, favoring the moving bed system. Putting aside concerns over operability of a fixed bed system (time varying flow, control of exothermic reaction waves in a large enclosed volume of solids, blinding of fixed beds with particulate material, shutdowns for periodic sorbent replacement), GEESI has assigned a factor of 0.6 to the SCS study's Total

Capital Requirement for the fixed bed desulfurization system of Case 5' accordingly giving a Total Capital Requirement for Case 5" of \$1336/KWe. This figure is similar to that for Case 1 and \$249/KWe lower in cost than the SCS study's original Case 5.

Clearly, the absolute accuracy of any of the cost numbers in Table 1 depends on site specifics and the level of design detail available for cost estimation. As has been mentioned earlier, Case 5" still carries an undetermined penalty, relative to Case 1, of the requirement for six (6) rather than four (4) gasification/HGCU flow trains. GEESI believes that if an IGCC system is to be designed with "polishing" external desulfurization utilizing a regenerable metal oxide process, this still requires utilization of piping, valves and vessels suited to the full gasifier fuel flow. If totally external desulfurization is considered, instead, inert bed material can be utilized in a fluid bed gasifier to modify the flow area required for superficial velocity considerations providing an equivalent number of flow trains to the partially internal redulfurization system. This reasoning leads one to believe that some further reduction in Total Capital Requirement for Case 5" may be achievable relative to Case 1.

TABLE 1. COMPARISON OF SCS STUDY CASES 1 AND 5 WITH GEESI REVISION OF CASE 5 (CASES 5' AND 5") UTILIZING MOVING BED HOT GAS CLEANUP SYSTEM

	<u>SCS SYSTEMS STUDY</u>		<u>GEESI REVISIONS TO CASE 5*</u>	
	<u>CASE 1</u>	<u>CASE 5</u>	<u>CASE 5'</u>	<u>CASE 5"</u>
NET POWER OUTPUT - MWe	458.4	420.2	454.4	454.4
NET PLANT HEAT RATE - BTU/KW-HR	8595	9356	8651	8651
THERMAL EFFICIENCY - %	39.7	36.5	39.5	39.5
TOTAL CAPITAL - REQUIREMENT - \$/KWe	1341	1585**	1389	1336

* Cases 5' utilizes direct substitution of the GEESI moving bed system for the fixed bed system utilized in Case 5. Credits for net output increases and Total Capital Requirement reduction have been applied, quantitatively, for Case 5'. Case 5" differs from Case 5' only through the application of a GEESI-derived cost ratio between the moving bed and fixed bed HGPU systems. A factor of 0.6 has been applied.

** TCR of Case 5 was subsequently increased to 1612 \$/KW

FIGURE 1. A comparison of moving and fixed bed regenerable metal oxide hot gas desulfurization systems suited to pressurized regeneration.

FIGURE 2. A simplified sulfuric acid plant designed to be compatible with the 800 F, 12 percent SO₂, dry tail gas stream from the GEESI moving bed hot gas desulfurization process.

APPENDIX C
GASIFICATION ISLAND COMPONENTS
RANKED BY CRITICALITY

APPENDIX INFORMATION
TABLE
COMPONENT RELIABILITY

		Hours			
<u>Equipment Number</u>	<u>Equipment Name</u>	<u>Case 1 & 3 (MTBF/MTR)</u>	<u>Case 2 & 4 (MTBF/MTR)</u>	<u>Case 5 (MTBF/MTR)</u>	<u>Case 6 (MTBF/MTR)</u>
<u>Coal Pressurization</u>					
301-J	Lockhopper Air Compr	26,316/40	26,316/40	26,316/40	26,316/40
302-F	Press Air Receiver	99,967/34	99,967/34	99,967/34	99,967/34
310-F	Coal Feed Surge Drum	10,989/4	10,989/4	10,989/4	10,989/4
	310-F Solids Valve	24,400/2	24,400/2	24,400/2	24,400/2
	310-F Pressure Valve	66,667/2	66,667/2	66,667/2	66,667/2
310-L	Coal Feed Vent Filter	99,967/8	99,967/8	99,967/8	99,967/8
311-L	Limestone Feed Vent Filter	N/A	N/A	N/A	99,967/8
310-V	Coal Feeder	1,610/2	1,610/2	1,610/2	1,610/2
311-V	Limestone Feeder	N/A	N/A	N/A	1,610/2

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

		Hours			
<u>Equipment Number</u>	<u>Equipment Name</u>	<u>Case 1 & 3 (MTBF/MTR)</u>	<u>Case 2 & 4 (MTBF/MTR)</u>	<u>Case 5 (MTBF/MTR)</u>	<u>Case 6 (MTBF/MTR)</u>
311-F	Coal Feed Press Hopper 311-F Valve	4,602/3 66,667/2	4,602/3 66,667/2	4,602/3 66,667/2	4,602/3 66,667/2
312-F	Coal Feed Hopper	7,513/4	7,513/4	7,513/4	7,513/4
313-F	Limestone Feed Surge Drum	N/A	N/A	N/A	10,989/4
	313-F Solids Valve	N/A	N/A	N/A	24,400/2
	313-F Pressure Valve	N/A	N/A	N/A	66,667/2
314-F	Limestone Feed Press Hopper	N/A	N/A	N/A	4,602/3
	314-F Valve	N/A	N/A	N/A	66,667/2
315-F	Limestone Feed Hopper	N/A	N/A	N/A	7,513/4
<u>Gasification</u>					
320-D	Gasifier	40,000/252	40,000/252	40,000/252	40,000/252
320-F	Gasifier Standpipe	N/A	N/A	N/A	40,000/212
320-L	Cyclone	40,000/120	40,000/120	40,000/120	N/A
320-LA,B	Primary Cyclones	N/A	N/A	N/A	22,000/150
321-LA,B	Secondary Cyclones	N/A	N/A	N/A	22,000/150
322-LA,B	Gasifier Standpipe Cyclones	N/A	N/A	N/A	40,000/100
330-V	Ash Feeder	1,610/2	1,610/2	1,610/2	N/A
<u>Fines Removal</u>					
421-F	Cyclone Fines Hopper	10,989/8	10,989/8	10,989/8	N/A
	421F Solids Valve	24,400/2	24,400/2	24,400/2	N/A
	421F Pressure Valve	66,667/2	66,667/2	66,667/2	N/A

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

		Hours			
<u>Equipment Number</u>	<u>Equipment Name</u>	<u>Case 1 & 3 (MTBF/MTR)</u>	<u>Case 2 & 4 (MTBF/MTR)</u>	<u>Case 5 (MTBF/MTR)</u>	<u>Case 6 (MTBF/MTR)</u>
421-L	Fines Cyclone	40,000/120	40,000/120	40,000/120	N/A
422-F	Cyclone Depress Hopper	10,989/8	10,989/8	10,989/8	N/A
	422F Valve	66,667/2	66,667/2	66,667/2	N/A
422-L	Cyclone Hopper Filter	99,978/8	99,978/8	99,978/8	N/A
430-F	Filter Depress Hopper	10,989/8	10,989/8	10,989/8	N/A
	430F Valve	66,667/2	66,667/2	66,667/2	N/A
430-L	Gas Filter	33,333/40	33,333/40	33,333/40	33,333/40
	430L Valve	66,667/2	66,667/2	66,667/2	24,400/2
430-V	LASH Conveyor	11,236/18	11,236/18	11,236/18	N/A
431-L	Filter Hopper Filter	17,520/8	17,520/8	17,520/8	N/A
<u>Chloride Guard</u> (Parallel Path)					
410-D	Chloride Guard Reactor	5,300/24	5,300/24	5,300/24	N/A
	410D Valve	17,520/6	17,520/6	17,520/6	N/A
	410D Valve	17,520/6	17,520/6	17,520/6	N/A
410-DA	Chloride Guard Reactor	5,300/24	5,300/24	5,300/24	N/A
	410D Valve	17,520/6	17,520/6	17,520/6	N/A
	410D Valve	17,520/6	17,520/6	17,520/6	N/A
<u>Zinc Ferrite Desulfurization</u>					
510-C	Exit Gas Cooler	99,864/34	99,864/34	99,864/34	N/A
510-D	Zinc Ferrite Reactors	70,000/250	70,000/250	70,000/250	N/A
	Sorbent	37,840/48	27,500/48	13,140/48	N/A
	Inlet Valve	21,900/6	21,900/6	21,900/6	N/A
	Outlet Valve	21,990/6	21,990/6	21,990/6	N/A
	Regeneration Valve	17,520/6	17,520/6	17,520/6	N/A
540-F	Guard Drum	87,600/34	87,600/34	87,600/34	N/A
541-L	Guard Cyclone	87,600/34	87,600/34	87,600/34	N/A

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

		Hours			
<u>Equipment Number</u>	<u>Equipment Name</u>	<u>Case 1 & 3 (MTBF/MTR)</u>	<u>Case 2 & 4 (MTBF/MTR)</u>	<u>Case 5 (MTBF/MTR)</u>	<u>Case 6 (MTBF/MTR)</u>
220-B	Startup Heater Hx Tubing	99,864/120 33,333/32	99,864/120 33,333/32	99,864/120 33,333/32	99,864/120 33,333/32
<u>Booster Air Compressor</u>					
251-C	Air Recuperator	99,864/34	99,864/34	99,864/34	99,864/34
251-F	Knock-Out Drum	91,104/16	91,104/16	91,104/16	91,104/16
251-J	Booster Air Compressor	26,316/40	26,316/40	26,316/40	26,316/40
252-CA	BFW Preheater	99,864/34	99,864/34	99,864/34	99,864/34
252-CB	Trim Cooler	99,864/34	99,864/34	99,864/34	99,864/34
<u>Sulfator</u>					
601-B	Sulfator Hx Coil	40,000/252 8,772/60	40,000/252 8,772/60	N/A N/A	40,000/252 N/A
601-C	Sulfator Heat Recovery	33,288/34	33,288/34	N/A	N/A
601-CA,B	Bayonet Boilers	N/A	N/A	N/A	22,000/150
601-F	LASH Receiver	10,989/80	10,989/80	N/A	N/A
601-J	Sulfation Air Compr	26,316/40	26,316/40	N/A	26,316/40 (Note 1)
601-JA	Sulfation Air Compr	26,316/40	26,316/40	N/A	26,316/40 (Note 1)
601-L	Baghouse Filter	26,280/50	26,280/50	N/A	N/A

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

		Hours			
<u>Equipment Number</u>	<u>Equipment Name</u>	<u>Case 1 & 3 (MTBF/MTR)</u>	<u>Case 2 & 4 (MTBF/MTR)</u>	<u>Case 5 (MTBF/MTR)</u>	<u>Case 6 (MTBF/MTR)</u>
601-LA,B	Sulfator Primary Cyclones	N/A	N/A	N/A	22,000/150
601-LV	Solids Feeder	17,520/18	17,520/18	N/A	N/A
601-V	Ash Screw Conveyor	N/A	N/A	N/A	66,700/39
602-C	Fluid Bed Heat Exchanger	N/A	N/A	N/A	22,000/225
602-F	Solids Disengager	87,600/34	87,600/34	N/A	N/A
602-LA,B	Sulfator Secondary Cyclones	N/A	N/A	N/A	22,000/150
602-V	Sulfator Solids Screw	11,236/18	11,236/18	N/A	N/A
603-F or 610-F	Sulfator Steam Drum	66,667/39	66,667/39	N/A	40,000/225
603-FA	Ash Depressurization Hopper	N/A	N/A	N/A	10,989/8
603-FB	Ash Depressurization Hopper	N/A	N/A	N/A	10,989/8
603-L	Sulfator Barrier Filter	N/A	N/A	N/A	9,000/130
603-V	Flue Gas Solids Screw	66,667/39	66,667/39	N/A	N/A
604-C	High Pressure Ash Cooler	N/A	N/A	N/A	40,000/225604-
F	Air Receiver	N/A	N/A	N/A	99,967/8
610-L	Fluid Bed Primary Cyclones	N/A	N/A	N/A	40,000/100
611-L	Fluid Bed Secondary Cyclones	N/A	N/A	N/A	40,000/100
612-L	Ash Depressurization Vent Filter	N/A	N/A	N/A	43,800/16

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

		Hours			
<u>Equipment Number</u>	<u>Equipment Name</u>	<u>Case 1 & 3 (MTBF/MTR)</u>	<u>Case 2 & 4 (MTBF/MTR)</u>	<u>Case 5 (MTBF/MTR)</u>	<u>Case 6 (MTBF/MTR)</u>
620-L	Ash Cooler Cyclone	N/A	N/A	N/A	40,000/40
<u>Heat Recovery</u>					
360-C	Product Gas Cooler	8,772/60	N/A	8,772/193	8,772/60
360-F	Steam Drum	52,356/50	N/A	52,356/50	52,356/50
<u>Recycle Gas</u>					
361-C	Recycle Gas Cooler	91,100/18	91,100/18	91,100/18	91,100/18
380-F	Recycle Gas Receiver	91,100/18	91,100/18	91,100/18	91,100/18
381-J	Recycle Gas Comp	13,158/40	13,158/40	13,158/40	13,158/40
N/A --- Not Applicable					

Notes:

1. For Case 6 601-J/JA are blowback air compressors. Air to the sulfator is supplied from the booster air compressor.

APPENDIX D
UNIRAM ANALYSIS DETAILS

Integrated Gasification Combined Cycle (IGCC)
Systems Assessment Project

DE-FC21-89MC26019

Reliability-Availability-Maintainability (RAM) Analysis

DRAFT REPORT

August, 1990

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Reliability Availability Maintainability (RAM) Analysis

Introduction

This section of the IGCC Systems Assessment Project Report shows the results of RAM analyses on each of six IGCC plant designs. The results were compared to an equivalent availability (EA) goal of 85%.

Also included in this section are the plant reliability simulation models, failure rate and repair data for plant components, and the major assumptions.

The EPRI software package, UNIRAM, was used for the analysis tool.

The RAM model and analysis have been used not only to provide estimates of expected plant performance but also as an interactive tool in refining the plant designs to assist in optimizing the designs by balancing cost with performance (equivalent availability).

Plant RAM Simulation Models

Models for each of the six different plant designs are depicted with simplified availability block diagrams (ABD). Some blocks on the ABD represent more than one component. The ABDs are attached in Appendix 1.

An ABD is a combination of series and parallel trains, composed of blocks representing major plant equipment or subsystems. If a component in series fails, the entire train in that series becomes unavailable. When a component in parallel fails, however, the other components in parallel with the failed component can carry all (100%) or a portion (less than 100%) of the plant rated capacity. The percentages printed above the blocks show the percentage of plant capacity that the block can carry.

Please note that although the ABD can not always accurately portray the exact physical layout or flow of the plant, the ABD logic of series and parallel relationships does accurately model the plant availability.

In general, the gasification island portion of the model is based on design drawings provided by and discussions with M. W. Kellogg; the power island, coal, limestone and ash handling systems, and balance of plant auxiliaries were modeled with information based on designs by Southern Company Services (SCS).

The plant design configurations, as represented by the ABDs, were modeled with the UNIRAM computer software program. UNIRAM was developed for EPRI by ARINC Research Corporation and has been in production since the early 1980's. The basic objective of UNIRAM is to relate failures or combinations of failures at the component level to the resultant curtailment, if any, of the power plant. UNIRAM calculates predicted estimates of equivalent availability (EA) and other performance measures.

The UNIRAM program outputs for the analyses of all six cases are presented in Appendix 2.

Data

The UNIRAM modeling program requires input data for each component in the form of mean time between failures (MTBF - inverse of the failure rate) and mean downtime (MDT).

Special components, such as material storage bins, provide surge time. Some components upstream (or downstream) of the surge components are affected in that a failure of these upstream (or downstream) components have a lesser impact on plant EA because the surge time provides time for repair of upstream (or downstream) components before the plant must be derated or shut down.

Other inputs to the UNIRAM model are expected annual planned outage time and reserve shutdown hours, if any.

Surge time, planned outage time and reserve shutdown time are discussed in the following Assumptions section. Appendix 2 contains a list of components with MTBF and MDT data. This is part of the UNIRAM output. Note that some component input data have been adjusted by UNIRAM to account for any surge time allowances.

The source of the data was mainly published EPRI reports. All gasification designs from M. W. Kellogg arrived complete with MTBF and MDT data for every component. The M. W. Kellogg data were derived from several EPRI and industrial data base sources. Most of the components in the SCS design used EPRI data from the following two reports:

Component Failure and Repair Data: Gasification-Combined-Cycle Power Generation Units (EPRI AP-2205)

and

Component Failure and Repair Data for Coal-Fired Power Plants (EPRI AP-2071).

A few components' data were developed using SCS engineering estimates and/or manufacturer information.

Assumptions

The plant is baseloaded, with no reserve shutdown hours. On average, the plant will required 412 hours per year for planned overhauls which reduces plant EA by 4.7%. The gross output of the plant varies with the design case under study, and is presented elsewhere in this report. The plant has been designed without backup fuel capability.

The heat recovery steam generators (HRSG) include deaerators and feedwater heaters.

The boiler feedwater preheater in the booster air compressor system can be bypassed.

The coal handling magnetic separator can be bypassed.

The coal handling silos provide 14 hours of surge time. The coal handling and limestone handling holding bins provides two hours surge time.

Except for the pneumatic limestone conveyers, the limestone handling facility was not included in the model. The limestone handling system has sufficient redundancies and surge capacities to consider it available 100% of the time. The milled limestone silo has 24 hour surge capacity. In addition, milled limestone can be trucked in and blown-in on an emergency basis. Twenty-four hours is sufficient lead time.

The ash handling silo capacity permits 20 hours of surge time for components downstream of the silo. The ash collection surge bin provides one hour surge time.

The gasification feed surge drums have capacity for 1-1/2 hours of surge time.

The zinc ferrite reactor sorbent changeout would occur during planned plant overhauls. There would, therefore, be no unplanned unavailability associated with this component.

The sulfator barrier filters (Case 6) will be replaced annually during the planned maintenance overhaul. Also, if a filter element would fail in service, a blowback operation will be performed to mitigate the impact. Therefore, the availability impact on the plant caused by the barrier filters was assumed to be insignificant.

The oxygen plant (Case 5) has 24 hours storage (surge time).

The sulfuric acid plant (Case 5) has very high reliability. The redundancies are sufficient to assume no negative impact on availability.

Results

The expected equivalent availability (EA) for the six IGCC plant designs analyzed are

EQUIVALENT AVAILABILITY RESULTS

	<u>EA</u>
Case 1	84.40%
Case 2	84.17%
Case 3	82.97%
Case 4	86.10%
Case 5	83.96%
Case 6	85.27%

Compared to the EA goal of 85%, two cases (4 and 6) have met the goal. The other four cases are within 1 or 2 percentage points of the goal.

The EA can be defined as the percentage of the plant's maximum energy production which was available during a time period.

The annual planned overhaul of 412 hours reduces the EA by 4.7 percentage points. This reduction is included in the resultant EA figures above. With the plant scheduled to be in service the balance of the year, the expected annual generation can be estimated. For example, in Case 1, a 400 MW plant, the expected generation is

$$(400 \text{ MW}) (8760 \text{ HOURS/YEAR}) (84.40\%) = 2,957,376 \frac{\text{MWHR}}{\text{YEAR}}$$

The more complete table, below, lists other performance indices for each case:

PLANT WANSLEY IGCC PROJECTED PERFORMANCE INDICES

	<u>EA(%)</u>	<u>A(%)</u>	<u>EFOR(%)</u>	<u>FOR(%)</u>	<u>Days Per Year At Maximum Output</u>
Case 1	84.40	93.06	11.43	2.35	236
Case 2	84.17	84.18	11.67	11.67	307
Case 3	82.97	88.12	12.93	7.53	279
Case 4	86.10	94.67	9.65	0.56	229
Case 5	83.96	92.98	11.89	2.44	215
Case 6	85.27	92.28	10.52	3.17	253

Where

- EA = Equivalent availability
- A = Availability
- EFOR = Equivalent forced outage rate
- FOR = Forced outage rate

Availability measures the percentage of time during which the plant is capable of operating (at any level of generation). Since there is no reserve shutdown time expected, the annual service hours can be estimated using the availability/. In Case 1 the expected annual service hours are

$$(8760 \text{ hours/year}) (93.06\%) = 8152 \text{ service hours per year}$$

The EFOR represents the percentage of the plant maximum generation that was desired during a certain time period but was not available because of forced events (both full forced outages and forced deratings).

The FOR is the percentage of time that the plant's service was desired within a certain time period but was unavailable due to full forced outages only.

Critical components

The three component types causing the most forced lost generation were generally the combustion turbine generators (CTG), heat recovery steam generators (HRSG) and the steam turbine generator (STG). Together, these components accounted for 42% to 59% of the equivalent forced outage hours (EFOH), depending on the case. Cases 2 and 4 have no STG.

Although the gasification island had no particular components contributing to a large percentage to the total forced outage problems, the gasification island, in total, contributed significantly because it has many components. The gasification island accounted for 36% to 55% of the EFOH. The sulfator system (part of the gasification island) accounted for 40% to 59% of the gasification island's forced outage problems. The three sulfator components causing the most lost generation were the LASH receiver, the sulfator and the sulfator's heat exchanger coil. The exception was Case 5 which has no sulfator. In Case 5, the product gas cooler in the gasification heat recovery system was the second largest contribution to plant EFOR - second only to the CTG. The product gas cooler has a mean down time (MDT) of 193 hours in Case 5 compared to an MDT of only 60 hours in Cases 1, 3 and 6.

A complete list of all components ranked by critically is included in Appendix 2 for all cases. This is part of the UNIRAM output reports. A summary listing appears below.

PLANT WANSLEY IGCC - CRITICAL COMPONENTS

<u>Component</u>	<u>Percentage of total EFOH</u>					
	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Case 4</u>	<u>Case 5</u>	<u>Case 6</u>
CTG	27	27	25	32	26	30
HRSG	15	15	14	18	14	17
STG	12	---	10	---	11	13
Gasification Island/ Sulfator System	42/22	55/23	48/20	46/27	---	36/14
Gasification Island/ Product Gas Cooler	---	---	---	---	44/15	---

Bypass Option

The option of installing HRSG bypass stacks was investigated. Each CTG has a dedicated HRSG which saves CTG exhaust heat for further electricity production, usually by expanding the steam through a steam turbine. Without bypass stacks, when an HRSG fails, its corresponding CTG must be shut down. With a bypass stack, the CTG can continue to operate and produce electricity during an HRSG forced outage. The CTG exhaust will be sent up the stack. If the STG fails, both HRSGs can be bypassed, and the CTGs operated.

The Case 1 model was modified to include HRSG bypasses. The EA improved two percentage points from 84.40% to 86.40%.

HRSG bypasses do, however, have some negative impacts which should be considered before deciding to install bypasses. The initial cost, the negative heat rate impact, the maintenance cost and maintenance work conditions are all factors to consider. Efficiency losses may even occur during normal HRSG operation because the bypass stacks may leak.

Sensitivity to Changes in Planned Overhaul Maintenance Schedules

The basic assumption presented in the analysis assumes an aggressive, optimistic planned overhaul schedule. The 412 hours per year or 17 days per year (4.7 percent planned outage factor) is consistent with the assumption used in the earlier EPRI Plant Wansley IGCC RAM analysis. The "highest availability plan" was developed by Fluor Engineers and Constructors. It is a 10-year plan which assumes shutting down the entire plant once per year to perform the longest duration maintenance activity and all shorter duration activities at the same time. Sufficient personnel, material and parts are assumed to be available.

This optimistic plan is presented along with four other more conservative overhaul plans in the EPRI Report AP-5276, Availability Analysis of an Integrated Gasification - Combined Cycle, June 1987.

Together, these five plans have planned outage factors (POF) ranging from 4.7 percent to 8.8 percent. As a sensitivity check, the EAF for six cases were recalculated using the most conservative planned overhaul schedule (8.8 percent POF developed by ARINC Research Corporation). This conservative plan averages 32 days per year overhaul downtime which approximates the projections used for some Southern electric system large coal-fired steam plants. In every case, the EAF falls about 3.6 percentage points. The results of this sensitivity check are presented below:

EFFECT OF PLANNED OVERHAUL SCHEDULES ON EQUIVALENT AVAILABILITY

	EQUIVALENT AVAILABILITY	
	OPTIMISTIC PLAN (17 DAYS PER YEAR)	MORE CONSERVATIVE PLAN (32 DAYS PER YEAR)
Case 1	84.40%	80.77%
Case 2	84.17%	80.55%
Case 3	82.97%	79.41%
Case 4	86.10%	82.40%
Case 5	83.96%	80.35%
Case 6	85.27%	81.61%

Since every case has a different design configuration, overhaul maintenance plans should be tailored to each individual plant design. This was not considered. A blanket assumption for overhaul downtime was used for all six cases.

Other factors affecting the choice and impact of overhaul plans are the availability of backup fuel, equipment bypass capability and whether the overhaul plan is staggered or a complete yearly plant shutdown.

Summary

Six different plant designs for the Plant Wansley Integrated Gasification Combined Cycle Systems Assessment Project were modeled. The analysis showed that equivalent availability expectations ranged from 82.97% to 86.10%. Two of the designs met the EA goal of 85%.

The RAM analysis was just one part of a research study which considered many other factors including cost and efficiency, in determining whether any of these technologies merit further consideration.

The UNIRAM models have been used interactively to assist design optimization. The models are still available for measuring the impacts of proposed changes to design, equipment, or maintenance plans on plant performance.

APPENDIX 1

AVAILABILITY BLOCK DIAGRAMS

Models for each of the six different plant designs are depicted with simplified availability block diagrams (ABD). Some blocks on the ABD represent more than one component.

An ABD is a combination of series and parallel trains, composed of blocks representing major plant equipment or subsystems. If a component in series fails, the entire train in that series becomes unavailable. When a component in parallel fails, however, the other components in parallel with the failed component can carry all (100%) or a portion (less than 100%) of the plant rated capacity. The percentages printed above the blocks (on the left) show the percentage of plant capacity that the block can carry. The numbers printed above the blocks (on the right) correspond to the identification number(s) of the component(s) with respect to the UNIRAM program outputs in Appendix 2.

Please note that although the ABD can not always accurately portray the exact physical layout or flow of the plant, the ABD logic of series and parallel relationships does accurately model the plant availability.

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM --- CASE 1 (page 1 of 2) (Gasification and Power Island)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM --- CASE 1 (page 2 of 2) (Coal, Limestone, and Ash Handling)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM -- CASE 2 (page 1 of 2) (Gasification and Power Island)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM --- CASE 2 (page 2 of 2) (Coal, Limestone, and Ash Handling)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM -- CASE 3 (page 1 of 2) (Gasification and Power Island)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM -- CASE 3 (page 2 of 2) (Coal, Limestone, and Ash Handling)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM -- CASE 4 (page 1 of 2) (Gasification and Power Island)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM --- CASE 4 (page 2 of 2) (Coal, Limestone, and Ash Handling)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM --- CASE 5 (page 1 of 2) (Gasification and Power Island)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM --- CASE 5 (page 2 of 2) (Coal and Ash Handling)

(Gasification and Power Island)

INTEGRATED GASIFICATION COMBINED CYCLE POWER PLANT AVAILABILITY BLOCK DIAGRAM --- CASE 6 (page 2 of 2) (Coal, Limestone, and Ash Handling)

UNIRAM PROGRAM OUTPUTS

WANSLEY IGCC STUDY II FOR D.O.E. (CASE 1)

>>> SCHEDULED OUTAGE HOURS = 412.0
 >>> RESERVE SHUTDOWN HOURS = .0
 >>> SIZE = 400.0 MW

 BASIC SUBSYSTEM INFORMATION

NUMBER OF BASIC SUBSYSTEMS = 21

ID#	NAME	THROUGHPUT CAPACITY	NUMBER OF GATES	GATE NUMBER	GATE TYPE	GATE ABOVE
1	BOOST_AC	50.0	1			
2	CATCHALL6	29.7	1	1	-1	0
3	SULFATOR	33.3	1	1	-1	0
4	CHLORIDEGD	29.7	1	1	-1	0
5	RGCOMPRESS	29.7	0			
6	STARTHEATR	100.0	1	1	-1	0
7	ZINCFERR	29.7	1	1	-1	0
8	DUMMY	100.0	0			
9	CTHRSG50	50.0	1	1	-1	0
10	BOPSTG	100.0	3	1	-1	0
				2	2	1
				3	2	1
11	BASICBFP	50.0	1	1	-1	0
12	DISCHARGE	33.3	1	1	-1	0
13	BAGHOUSE	100.0	0			
14	VACUUM	100.0	1	1	-1	0
15	SIL0	100.0	0			
16	AERATION	100.0	1	1	-1	0
17	CONDITION	100.0	1	1	-1	0
18	DRYUNLOAD	100.0	1	1	-1	0
19	COAL100	100.0	2	1	-1	0
				2	2	1
20	MILLINGSYS	33.3	1	1	-1	0
21	LASTFEED	33.3	1	1	-1	0

 COMPONENT INFORMATION

TOTAL NUMBER OF COMPONENTS = 102

 SUBSYSTEM ID#/NAME

COMPONENT ID#/NAME	GATE ABOVE	MTBF (HOURS)	MDT (HOURS)
1 BOOST AC			
1 BFWPREHEATER	1	10000000.0	1.0
2 TRIMCOOLER	1	99864.0	34.0
3 KNOCKOUTDRUM	1	91104.0	16.0
4 BOOSTERAC	1	26316.0	40.0
5 AIRRECUPERATOR	1	99864.0	34.0
2 CATCHALL6			
6 FEEDSURGEDRUM	1	10989.0	4.0
7 VALVECP1	1	24400.0	2.0
8 VALVECP2	1	66667.0	2.0
9 PRESSZN.AC	1	26313.0	40.0
10 PRSZN.AIRRECEIVER	1	99967.0	34.0
11 COALFEEDVENTFILTER	1	99967.0	8.0
12 PRSZN.FEEDHOPPER	1	4602.0	3.0
13 VALVECP3	1	66667.0	2.0
14 FEEDHOPPER	1	7513.0	4.0
15 COALFEEDER	1	1610.0	2.0
16 GASIFIER	1	40000.0	252.0
17 CYCLONE	1	40000.0	120.0
18 ASHFEEDER	1	1610.0	2.0
19 PRODUCTGASCOOLER	1	8772.0	60.0
20 STEAMDRUM	1	52356.0	50.0
21 FINESCYCLONE	1	40000.0	120.0
22 CYC.FINE.LOCKHOPPER	1	10989.0	8.0
23 VALVEFR1	1	24400.0	2.0
24 VALVEFR2	1	66667.0	2.0
25 CFDEPRESSLOCKHOPPER	1	10989.0	8.0
26 CYCLOCKHOPPERFILTER	1	99978.0	8.0
27 VALVEFR3	1	66667.0	2.0
28 GASFILTER	1	33333.0	40.0
29 VALVEFR5	1	66667.0	2.0
30 G.F.DEPRSSLOCKHOPPER	1	10989.0	8.0
31 GFLOCKHOPPERFILTER	1	17520.0	8.0
32 VALVEFR6	1	66667.0	2.0
33 SOLIDSCONVEYOR	1	11236.0	18.0
34 RECYCLEGASCOOLER	1	91100.0	18.0
35 RECYCLEGASRECEIVER	1	91100.0	18.0
3 SULFATOR			
36 LASHRECEIVER	1	10989.0	80.0
37 SCREWFEEDER	1	17520.0	18.0
38 SULFN.AIRCOMPRESSOR	1	26316.0	40.0
39 SULFATOR	1	40000.0	252.0
40 SULFATORINTERNALS	1	8772.0	60.0
41 SOLIDSDISENGAGER	1	87600.0	34.0
42 SULFATORHEATRECOVERY	1	33288.0	34.0
43 SULFATORSTEAMDRUM	1	66667.0	39.0
44 FLUGASBAGHOUSEFILTER	1	26280.0	50.0
45 FLGASOLIDSCREWCOOLER	1	66667.0	39.0
46 SULFSOLIDSCREWCOOLER	1	11236.0	18.0
4 CHLORIDEGD			
47 VALVECG1	1	17520.0	6.0
48 CHLORIDEGUARDDRUM	1	5300.0	24.0
49 VALVECG2	1	17520.0	6.0
5 RGCOMPRESS			
50 RECYCLEGASCOMPRESSOR	0	13158.0	40.0
6 STARTHEATR			
51 STARTUPHEATER	1	99864.0	120.0
52 ST.UP.HEAT.INTERNAL	1	33333.0	32.0
7 ZINCERR			
53 VALVEZFDS1	1	21900.0	6.0
54 ZINCERRITEREACTOR	1	70000.0	250.0
55 ZNFERRITESORBENT	1	10000000.0	1.0

56	VALVEZFDS2	1	21900.0	6.0
57	VALVEREGENER3	1	17520.0	6.0
58	VALVEZFDS4	1	21900.0	6.0
59	ZINCFERRITEREACTOR2	1	70000.0	250.0
60	ZNFERRITESORBENT2	1	1000000.0	1.0
61	VALVEZFDS5	1	21900.0	6.0
62	VALVEREGENER6	1	17520.0	6.0
63	GUARDCYCLONE	1	87600.0	34.0
64	GUARDDRUM	1	87600.0	34.0
65	EXITGASCOOLER	1	99864.0	34.0
8	DUMMY			
	66 DUMMYCOMP	0	59475.6	18.0
9	CTHRSG50			
	67 CTG	1	4300.0	146.7
	68 HRSG	1	1321.0	25.0
10	BOPSTG			
	69 CONDPUMP	2	33333.0	31.0
	70 CIRCWPUMP	3	100000.0	29.0
	71 CONDENSER	1	25000.0	36.0
	72 STG	1	5076.0	71.0
11	BASICBFP			
	73 BFP	1	10000.0	54.0
12	DISCHARGE			
	74 SURGEBIN	1	10989.0	4.2
	75 OUTCYLVALVE	1	33744.2	4.0
13	BAGHOUSE			
	76 BAGHOUSES	0	730.0	2.0
14	VACUUM			
	77 FILTER	1	16404.0	5.3
	78 VACPUMP	1	5711.0	28.2
15	SIL0			
	79 ASHSILO	0	116508.0	45.5
16	AERATION			
	80 BLOWER	1	17786.0	28.2
17	CONDITION			
	81 CVALVE	1	3900298.0	4.0
	82 ASHCONDITIONER	1	869481.5	4.3
18	DRYUNLOAD			
	83 DVALVE	1	3900298.0	4.0
	84 UNLOADERBLOWER	1	17786.0	28.2
19	COAL100			
	85 RECLAIMCONVEYOR	1	37651.4	18.3
	86 TRANSFERCONVEYOR1	1	37651.4	18.3
	87 TRANSFERCONVEYOR2	1	37651.4	18.3
	88 TRANSFERCONVEYOR3	1	37651.4	18.3
	89 TRANSFERCONVEYOR4	1	37651.4	18.3
	90 CASCADECONVEYOR	1	37651.4	18.3
	91 PNEULIMECONVEYORS	2	27561.7	42.0
	92 HOLDINGBIN	1	15705.9	4.2
20	MILLINGSYS			
	93 EXHAUSTFAN	1	9408.6	28.0
	94 BLOWFAN	1	9408.6	28.0
	95 BAGFILTER	1	9521.3	24.0
	96 CYCLONE	1	49631.9	16.0
	97 COALSILO	1	109090.9	11.1
	98 COALFEEDER1	1	22716.3	7.7
	99 COALMILL	1	31856.4	15.5
21	LASTFEED			
	100 LIMESTONEFEEDER	1	21288.1	7.7
	101 COALFEEDER2	1	21288.1	7.7
	102 PNEUMATICELEVATOR	1	27235.5	42.0

NESTED SUBSYSTEM INFORMATION

NUMBER OF NESTED SUBSYSTEMS = 20

ID#	NAME	STRUCTURE	CODE
22	NCHLRGD100	PARALLEL-IDENT	2
23	NRGCOMPR	PARALLEL-IDENT	2
24	NGAS	SERIAL	4
25	NGASTRAIN2	PARALLEL	3
26	NGASTRAIN	SERIAL	4
27	NGASTRAINS	PARALLEL	3
28	NBOOSTAC	PARALLEL-IDENT	2
29	NSULFATOR	PARALLEL-IDENT	2
30	NCTHRSG	PARALLEL-IDENT	2
31	N3BFP50	PARALLEL-IDENT	2
32	NSURGE	PARALLEL-IDENT	2
33	NBAG	PARALLEL-IDENT	2
34	NVAC	PARALLEL-IDENT	2
35	NAER	PARALLEL-IDENT	2
36	NCONDIT	PARALLEL	3
37	NASHSYSTEM	SERIAL	4
38	NMILL	PARALLEL-IDENT	2
39	NFEED	PARALLEL-IDENT	2
40	COALLIME	SERIAL	4
41	NPLANT	SERIAL	4

SUBSYSTEMS DIRECTLY CONTAINED				
ID#	NAME	STRUCTURE	CODE	QUANTITY
4	CHLORIDEGD	BASIC	1	2
5	RGCOMPRESS	BASIC	1	3
2	CATCHALL6	BASIC	1	1
7	ZINCFERR	BASIC	1	1
22	NCHLRGD100	PARLL-IDNT	2	1
24	NGAS	SERIAL	4	1
24	NGAS	SERIAL	4	1
25	NGASTRAIN2	PARALLEL	3	1
23	NRGCOMPR	PARLL-IDNT	2	1
26	NGASTRAIN	SERIAL	4	1
26	NGASTRAIN	SERIAL	4	1
1	BOOST_AC	BASIC	1	2
3	SULFATOR	BASIC	1	3
9	CTHRSG50	BASIC	1	2
11	BASICBFP	BASIC	1	3
12	DISCHARGE	BASIC	1	4
13	BAGHOUSE	BASIC	1	2
14	VACUUM	BASIC	1	2
16	AERATION	BASIC	1	2
17	CONDITION	BASIC	1	1
17	CONDITION	BASIC	1	1
18	DRYUNLOAD	BASIC	1	1
15	SILO	BASIC	1	1
32	NSURGE	PARLL-IDNT	2	1
33	NBAG	PARLL-IDNT	2	1
34	NVAC	PARLL-IDNT	2	1
35	NAER	PARLL-IDNT	2	1
36	NCONDIT	PARALLEL	3	1
20	MILLINGSYS	BASIC	1	4
21	LASTFEED	BASIC	1	4
19	COAL100	BASIC	1	1
38	NMILL	PARLL-IDNT	2	1
39	NFEED	PARLL-IDNT	2	1
6	STARTHEATR	BASIC	1	1
10	BOPSTG	BASIC	1	1
27	NGASTRAINS	PARALLEL	3	1
28	NBOOSTAC	PARLL-IDNT	2	1
29	NSULFATOR	PARLL-IDNT	2	1

30	NCTHRSG	PARLL-IDNT	2	1
31	N3BFP50	PARLL-IDNT	2	1
37	NASHSYSTEM	SERIAL	4	1
40	COALLIME	SERIAL	4	1

 ***** RUN 1 *****

BASELINE RUN

FAULT TREE EVALUATION

SUBSYSTEM	MEAN DOWNTIME (HOURS)	MEAN TIME BETWEEN FAILURES (HOURS)	AVAILABILITY
1 BOOST AC	34.41	14471.07	.9976
2 CATCHALL6	12.60	384.21	.9682
3 SULFATOR	56.15	1909.91	.9714
4 CHLORIDEGD	17.23	3302.13	.9948
5 RGCOMPRESS	40.00	13158.00	.9970
6 STARTHEATR	54.05	24991.30	.9978
7 ZINCFERR	28.12	2790.02	.9900
8 DUMMY	18.00	59475.55	.9997
9 CTHRSG50	54.25	1010.55	.9490
10 BOPSTG	65.16	4218.22	.9848
11 BASICBFP	54.00	10000.00	.9946
12 DISCHARGE	4.15	8289.48	.9995
13 BAGHOUSE	2.00	730.00	.9973
14 VACUUM	22.31	4236.19	.9948
15 SILO	45.50	116508.00	.9996
16 AERATION	28.24	17785.99	.9984
17 CONDITION	4.28	710984.00	1.0000
18 DRYUNLOAD	28.13	17705.25	.9984
19 COAL100	14.30	4481.55	.9968
20 MILLINGSYS	23.10	2367.84	.9903
21 LASTFEED	17.35	7653.10	.9977

STATE DEFINITIONS WITH CORRESPONDING OUTPUT CAPABILITIES

SUBSYSTEM	CURTAIL- MENT LEVEL	STATE AVAILABILITY (PERCENTAGE)	STATE OUTPUT CAPACITY (PERCENTAGE)	STATE DEFINITIONS (SUBSYSTEM IDENTIFICATION NUMBERS AND CURTAILMENT LEVELS)
22 NCHLRGD100				4
	0	98.9648	59.	0
	1	1.0325	30.	1
	2	.0027	0.	2
23 NRGCOMPR				5
	0	99.0935	89.	0
	1	.9037	59.	1
	2	.0027	30.	2
	3	.0000	0.	3

-----				-----			
24 NGAS				2 7 22			
-----				-----			
	0	95.8558	30.	0	0	-2	
	1	4.1442	0.	*			
-----				-----			
25 NGASTRIN2				24 24			
-----				-----			
	0	91.8833	59.	0	0		
	1	3.9725	30.	0	1		
		3.9725	30.	1	0		
	2	.1717	0.	*			
-----				-----			
26 NGASTRIN				25 23			
-----				-----			
	0	91.8808	59.	0	-2		
	1	7.9449	30.	1	-3		
		.0025	30.	0	2		
	2	.1717	0.	*			
-----				-----			
27 NGASTRINS				26 26			
-----				-----			
	0	84.4208	119.	0	0		
	1	7.3022	89.	0	1		
		7.3022	89.	1	0		
	2	.1578	59.	0	2		
		.6316	59.	1	1		
		.1578	59.	2	0		
	3	.0136	30.	1	2		
		.0136	30.	2	1		
	4	.0003	0.	*			
-----				-----			
28 NBOOSTAC				1			
-----				-----			
	0	99.5261	100.	0			
	1	.4734	50.	1			
	2	.0006	0.	2			
-----				-----			
29 NSULFATOR				3			
-----				-----			
	0	91.6742	100.	0			
	1	8.0857	67.	1			
	2	.2377	33.	2			
	3	.0023	0.	3			
-----				-----			
30 NCTHRSG				9			
-----				-----			
	0	90.0693	100.	0			
	1	9.6711	50.	1			
	2	.2596	0.	2			
-----				-----			
31 N3BFP50				11			
-----				-----			
	0	98.3973	150.	0			
	1	1.5940	100.	1			
	2	.0086	50.	2			
	3	.0000	0.	3			
-----				-----			
32 NSURGE				12			
-----				-----			
	0	99.7999	133.	0			
	1	.1999	100.	1			
	2	.0002	67.	2			
	3	.0000	33.	3			
	4	.0000	0.	4			
-----				-----			

33 NBAG			
0	99.4543	200.	
1	.5450	100.	
2	.0007	0.	
34 NVAC			
0	98.9550	200.	
1	1.0423	100.	
2	.0027	0.	
35 NAER			
0	99.6832	200.	
1	.3165	100.	
2	.0003	0.	
36 NCONDIT			
0	99.8402	300.	
1	.1586	200.	
	.0006	200.	
	.0006	200.	
2	.0000	100.	
	.0000	100.	
	.0000	100.	
3	.0000	0.	
37 NASHSYSTEM			
0	99.7572	100.	
1	.1998	100.	
2	.0002	67.	
3	.0000	33.	
4	.0428	0.	
38 NMILL			
0	96.1904	133.	
1	3.7543	100.	
2	.0549	67.	
3	.0004	33.	
4	.0000	0.	
39 NFEEED			
0	99.0984	133.	
1	.8985	100.	
2	.0031	67.	
3	.0000	33.	
4	.0000	0.	
40 COALLIME			
0	95.0201	100.	
1	3.7422	100.	
	.8615	100.	
2	.0548	67.	
	.0030	67.	
3	.0004	33.	
	.0000	33.	
4	.3180	0.	
41 NPLANT			

13						

0						
1						
2						

14						

0						
1						
2						

16						

0						
1						
2						

17	17	18				

0	0	0				
0	0	1				
0	1	0				
1	0	0				
0	1	1				
1	0	1				
1	1	0				
*						

15	32	33	34	35	36	

0	0	-2	-2	-2	-3	
0	1	-2	-2	-2	-3	
0	2	-2	-2	-2	-3	
0	3	-2	-2	-2	-3	
*						

20						

0						
1						
2						
3						
4						

21						

0						
1						
2						
3						
4						

19	38	39				

0	0	0				
0	1	-2				
0	0	1				
0	2	-3				
0	-2	2				
0	3	-4				
0	-3	3				
*						

6	10	27	28	29	30	31 37

0	67.8818	100.	0	0	0	0	0	0	-2	-2
1	11.7432	89.	0	0	1	0	0	0	-2	-2
2	.0462	67.	0	0	-2	0	0	0	-2	-2
3	7.0271	67.	0	0	-2	0	1	0	-2	-3
	.0001	67.	0	0	-2	0	0	0	-2	2
4	.8293	59.	0	0	2	0	-2	0	-2	-3
5	.4610	50.	0	0	-3	1	-2	-2	-3	-3
	9.3990	50.	0	0	-3	0	-2	1	-3	-3
	.0075	50.	0	0	-3	0	-2	0	2	-3
6	.0004	33.	0	0	-3	-2	-2	-2	-3	-3
7	.2321	33.	0	0	-3	-2	2	-2	-3	-4
	.0000	33.	0	0	-3	-2	-2	-2	-3	3
8	.0267	30.	0	0	3	-2	-3	-2	-3	-4
9	2.3456	0.	*							

40

0	67.8818	100.	-2
1	11.7432	89.	-2
2	.0462	67.	2
3	7.0271	67.	-3
	.0001	67.	-3
4	.8293	59.	-3
5	.4610	50.	-3
	9.3990	50.	-3
	.0075	50.	-3
6	.0004	33.	3
7	.2321	33.	-4
	.0000	33.	-4
8	.0267	30.	-4
9	2.3456	0.	*

NOTE: "*" INDICATES THAT NO SINGLE DEFINITION CAN COMPLETELY DESCRIBE A GIVEN SUBSTATE.
SEE USER'S GUIDE FOR DETAILED EXPLANATION.

PLANT STATE	OUTPUT STATE PROBABILITY	OUTPUT CAPABILITY	DAYS/YEAR	POWER (MW) OUTPUT
1	67.88%	99.99%	236.12	399.96
2	11.74%	89.10%	40.85	356.40
3	.05%	66.67%	.1	266.66
4	7.03%	66.66%	24.4	266.64
5	.83%	59.40%	2.88	237.60
6	9.87%	50.00%	34.32	200.00
7	.00%	33.33%	.00	133.33
8	.23%	33.33%	.81	133.32
9	.03%	29.70%	.09	118.80
10	2.35%	.00%	8.16	.00

I	PLANT RAM MEASURES				I
I					I
I					I
I		EQUIVALENT			I
I	FORCED	FORCED	EQUIVALENT		I
I	OUTAGE	OUTAGE	AVAILABILITY	AVAILABILITY	I
I	RATE	RATE	FACTOR	FACTOR	I
I	2.35%	11.43%	84.40%	93.06%	I

UNIT-LEVEL R&M VALUES				
UNIT MEAN TIME	MEAN UNIT FULL			
BETWEEN FULL	FORCED OUTAGE			
FORCED OUTAGES	DURATION			
(HOURS)	(HOURS)			
1620.1	38.9			

WANSLEY IGCC STUDY 1I FOR D.O.E. (CASE 2)

>> SCHEDULED OUTAGE HOURS = 412.0
 >>> RESERVE SHUTDOWN HOURS = .0
 >>> SIZE = 100.0 MW

 BASIC SUBSYSTEM INFORMATION NUMBER OF BASIC SUBSYSTEMS = 20

ID#	NAME	THROUGHPUT CAPACITY	NUMBER OF GATES	GATE NUMBER	GATE TYPE	GATE ABOVE
1	BOOST_AC	100.0	1			
2	CATCHALL6	100.0	1	1	-1	0
3	SULFATOR	100.0	1	1	-1	0
4	CHLORIDEGD	100.0	1	1	-1	0
5	RGCOMPRESS	100.0	0			
6	STARTHEATR	100.0	1	1	-1	0
7	ZINCFERR	100.0	1	1	-1	0
8	DUMMY	100.0	0			
9	CTHRSG50	100.0	1	1	-1	0
10	BASICBFP	50.0	1	1	-1	0
11	DISCHARGE	100.0	1	1	-1	0
12	BAGHOUSE	100.0	0			
13	VACUUM	100.0	1	1	-1	0
14	SILO	100.0	0			
15	AERATION	100.0	1	1	-1	0
16	CONDITION	100.0	1	1	-1	0
17	DRYUNLOAD	100.0	1	1	-1	0
18	COAL100	100.0	2	1	-1	0
				2	2	1
19	MILLINGSYS	100.0	1	1	-1	0
20	LASTFEED	100.0	1	1	-1	0

 COMPONENT INFORMATION TOTAL NUMBER OF COMPONENTS = 96

SUBSYSTEM ID#/NAME	GATE ABOVE	MTBF (HOURS)	MDT (HOURS)
COMPONENT ID#/NAME			

1	BOOST AC			
1	BFWPREHEATER	1	10000000.0	1.0
2	TRIMCOOLER	1	99864.0	34.0
3	KNOCKOUTDRUM	1	91104.0	16.0
4	BOOSTERAC	1	26316.0	40.0
5	AIRRECUPERATOR	1	99864.0	34.0
2	CATCHALL6			
6	FEEDSURGEDRUM	1	10989.0	4.0
7	VALVECP1	1	24400.0	2.0
8	VALVECP2	1	66667.0	2.0
9	PRESSZN.AC	1	26313.0	40.0
10	PRSZN.AIRRECEIVER	1	99967.0	34.0
11	COALFEEDVENTFILTER	1	99967.0	8.0
12	PRSZN.FEEDHOPPER	1	4602.0	3.0
13	VALVECP3	1	66667.0	2.0
14	FEEDHOPPER	1	7513.0	4.0
15	COALFEEDER	1	1610.0	2.0
16	GASIFIER	1	40000.0	252.0
17	CYCLONE	1	40000.0	120.0
18	ASHFEEDER	1	1610.0	2.0
19	FINESCYCLONE	1	40000.0	120.0
20	CYC.FINE.LOCKHOPPER	1	10989.0	8.0
21	VALVEFR1	1	24400.0	2.0
22	VALVEFR2	1	66667.0	2.0
23	CFDEPRESSLOCKHOPPER	1	10989.0	8.0
24	CYCLOCKHOPPERFILTER	1	99978.0	8.0
25	VALVEFR3	1	66667.0	2.0
26	GASFILTER	1	33333.0	40.0
27	VALVEFR5	1	66667.0	2.0
28	G.F.DEPRSSLOCKHOPPER	1	10989.0	8.0
29	GFLOCKHOPPERFILTER	1	17520.0	8.0
30	VALVEFR6	1	66667.0	2.0
31	SOLIDSCONVEYOR	1	11236.0	18.0
32	RECYCLEGASCOOLER	1	91100.0	18.0
33	RECYCLEGASRECEIVER	1	91100.0	18.0
3	SULFATOR			
34	LASHRECEIVER	1	10989.0	80.0
35	SCREWFEEDER	1	17520.0	18.0
36	SULFN.AIRCOMPRESSOR	1	26316.0	40.0
37	SULFATOR	1	40000.0	252.0
38	SULFATORINTERNALS	1	8772.0	60.0
39	SOLIDSDISENGAGER	1	87600.0	34.0
40	SULFATORHEATRECOVERY	1	33288.0	34.0
41	SULFATORSTEAMDRUM	1	66667.0	39.0
42	FLUGASBAGHOUSEFILTER	1	26280.0	50.0
43	FLGASOLIDSCREWCOOLER	1	66667.0	39.0
44	SULFSOLIDSCREWCOOLER	1	11236.0	18.0
4	CHLORIDEGD			
45	VALVECG1	1	17520.0	6.0
46	CHLORIDEGUARDDRUM	1	5300.0	24.0
47	VALVECG2	1	17520.0	6.0
5	RGCOMPRESS			
48	RECYCLEGASCOMPRESSOR	0	13158.0	40.0
6	STARTHEATR			
49	STARTUPHEATER	1	99864.0	120.0
50	ST.UP.HEAT.INTERNAL	1	33333.0	32.0
7	ZINCERR			
51	VALVEZFDS1	1	21900.0	6.0
52	ZINCERRITEREACTOR	1	70000.0	250.0
53	ZNFERRITESORBENT	1	10000000.0	1.0
54	VALVEZFDS2	1	21900.0	6.0
55	VALVEREGENER3	1	17520.0	6.0
56	VALVEZFDS4	1	21900.0	6.0
57	ZINCERRITEREACTOR2	1	70000.0	250.0
58	ZNFERRITESORBENT2	1	10000000.0	1.0
59	VALVEZFDS5	1	21900.0	6.0

60	VALVEREGENER6	1	17520.0	6.0
61	GUARDCYCLONE	1	87600.0	34.0
62	GUARDDRUM	1	87600.0	34.0
63	EXITGASCOOLER	1	99864.0	34.0
8	DUMMY			
64	DUMMYCOMP	0	59475.6	18.0
9	CTHRSG50			
65	CTG	1	4300.0	146.7
66	HRSG	1	1321.0	25.0
10	BASICBFP			
67	BFP	1	10000.0	54.0
11	DISCHARGE			
68	SURGEBIN	1	10989.0	4.2
69	OUTCYLVALVE	1	33744.2	4.0
12	BAGHOUSE			
70	BAGHOUSES	0	730.0	2.0
13	VACUUM			
71	FILTER	1	16404.0	5.3
72	VACPUMP	1	5711.0	28.2
14	SIL0			
73	ASHSILO	0	116508.0	45.5
15	AERATION			
74	BLOWER	1	17786.0	28.2
16	CONDITION			
75	CVALVE	1	3900298.0	4.0
76	ASHCONDITIONER	1	869461.5	4.3
17	DRYUNLOAD			
77	DVALVE	1	3900298.0	4.0
78	UNLOADERBLOWER	1	17786.0	28.2
18	COAL100			
79	RECLAIMCONVEYOR	1	37651.4	18.3
80	TRANSFERCONVEYOR1	1	37651.4	18.3
81	TRANSFERCONVEYOR2	1	37651.4	18.3
82	TRANSFERCONVEYOR3	1	37651.4	18.3
83	TRANSFERCONVEYOR4	1	37651.4	18.3
84	CASCADECONVEYOR	1	37651.4	18.3
85	PNEULIMECONVEYORS	2	27561.7	42.0
86	HOLDINGBIN	1	15705.9	4.2
19	MILLINGSYS			
87	COALSILO	1	109090.9	11.1
88	COALFEEDER1	1	22716.3	7.7
89	COALMILL	1	31856.4	15.5
90	EXHAUSTFAN	1	9408.6	28.0
91	BLOWFAN	1	9408.6	28.0
92	BAGFILTER	1	9521.3	24.0
93	CYCLONE	1	49631.9	16.0
20	LASTFEED			
94	LIMESTONEFEEDER	1	21288.1	7.7
95	COALFEEDER2	1	21288.1	7.7
96	PNEUMATICELEVATOR	1	27235.5	42.0

 NESTED SUBSYSTEM INFORMATION
 NUMBER OF NESTED SUBSYSTEMS = 13

ID#	NAME	STRUCTURE	CODE	SUBSYSTEMS DIRECTLY CONTAINED		
-----	-----	-----	-----	ID#	NAME	STRUCTURE CODE QUANTITY
-----	-----	-----	-----	-----	-----	-----

21	NCHLRGD100	PARALLEL-IDENT	2
22	NRGCOMPR	PARALLEL-IDENT	2

4	CHLORIDEGD BASIC	1	2
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23	N3BFP50	PARALLEL-IDENT	2	5	RGCOMPRESS	BASIC	1	2
24	NSURGE	PARALLEL-IDENT	2	10	BASICBFP	BASIC	1	3
25	NBAG	PARALLEL-IDENT	2	11	DISCHARGE	BASIC	1	2
26	NVAC	PARALLEL-IDENT	2	12	BAGHOUSE	BASIC	1	3
27	NAER	PARALLEL-IDENT	2	13	VACUUM	BASIC	1	2
28	NCONDIT	PARALLEL	3	15	AERATION	BASIC	1	2
				16	CONDITION	BASIC	1	1
				16	CONDITION	BASIC	1	1
				17	DRYUNLOAD	BASIC	1	1
29	NASHSYSTEM	SERIAL	4	14	SILO	BASIC	1	1
				24	NSURGE	PARLL-IDNT	2	1
				25	NBAG	PARLL-IDNT	2	1
				26	NVAC	PARLL-IDNT	2	1
				27	NAER	PARLL-IDNT	2	1
				28	NCONDIT	PARALLEL	3	1
30	NMILL	PARALLEL-IDENT	2	19	MILLINGSYS	BASIC	1	2
31	NFEED	PARALLEL-IDENT	2	20	LASTFEED	BASIC	1	2
32	NCOALLIME	SERIAL	4	18	COAL100	BASIC	1	1
				30	NMILL	PARLL-IDNT	2	1
				31	NFEED	PARLL-IDNT	2	1
33	PLANT	SERIAL	4	1	BOOST_AC	BASIC	1	1
				2	CATCHALL6	BASIC	1	1
				3	SULFATOR	BASIC	1	1
				6	STARTHEATR	BASIC	1	1
				7	ZINCFERR	BASIC	1	1
				9	CTHRSG50	BASIC	1	1
				21	NCHLRGD100	PARLL-IDNT	2	1
				22	NRGCOMPR	PARLL-IDNT	2	1
				23	N3BFP50	PARLL-IDNT	2	1
				29	NASHSYSTEM	SERIAL	4	1
				32	NCOALLIME	SERIAL	4	1

 ***** RUN 1 *****

BASELINE RUN

FAULT TREE EVALUATION

SUBSYSTEM	MEAN DOWNTIME (HOURS)	MEAN TIME BETWEEN FAILURES (HOURS)	AVAILABILITY
1 BOOST_AC	34.41	14471.07	.9976
2 CATCHALL6	10.04	404.92	.9758
3 SULFATOR	56.15	1909.91	.9714
4 CHLORIDEGD	17.23	3302.13	.9948
5 RGCOMPRESS	40.00	13158.00	.9970
6 STARTHEATR	54.05	24991.30	.9978

7	ZINCFERR	28.12	2790.02	.9900
8	DUMMY	18.00	59475.55	.9997
9	CTHRSG50	54.25	1010.55	.9490
10	BASICBFP	54.00	10000.00	.9946
11	DISCHARGE	4.15	8289.48	.9995
12	BAGHOUSE	2.00	730.00	.9973
13	VACUUM	22.31	4236.19	.9948
14	SILO	45.50	116508.00	.9996
15	AERATION	28.24	17785.99	.9984
16	CONDITION	4.28	710984.00	1.0000
17	DRYUNLOAD	28.13	17705.25	.9984
18	COAL100	14.30	4481.55	.9968
19	MILLINGSYS	23.10	2367.84	.9903
20	LASTFEED	17.35	7653.10	.9977

STATE DEFINITIONS WITH CORRESPONDING
OUTPUT CAPABILITIES

SUBSYSTEM	CURTAIL- MENT LEVEL	STATE AVAILABILITY (PERCENTAGE)	STATE OUTPUT CAPACITY (PERCENTAGE)	STATE DEFINITIONS (SUBSYSTEM IDENTIFICATION NUMBERS AND CURTAILMENT LEVELS)

21 NCHLRGD100				4

	0	98.9648	200.	0
	1	1.0325	100.	1
	2	.0027	0.	2

22 NRGCOMPR				5

	0	99.3948	200.	0
	1	.6043	100.	1
	2	.0009	0.	2

23 N3BFP50				10

	0	98.3973	150.	0
	1	1.5940	100.	1
	2	.0086	50.	2
	3	.0000	0.	3

24 NSURGE				11

	0	99.8999	200.	0
	1	.1001	100.	1
	2	.0000	0.	2

25 NBAG				12

	0	99.4543	200.	0
	1	.5450	100.	1
	2	.0007	0.	2

26 NVAC				13

	0	98.9550	200.	0
	1	1.0423	100.	1
	2	.0027	0.	2

27 NAER				15

	0	99.6832	200.	0
	1	.3165	100.	1

PLANT RAM MEASURES				
EQUIVALENT				
FORCED	FORCED	EQUIVALENT		
OUTAGE	OUTAGE	AVAILABILITY	AVAILABILITY	
RATE	RATE	FACTOR	FACTOR	
11.67%	11.67%	84.17%	84.18%	

UNIT-LEVEL R&M VALUES			
UNIT MEAN TIME	MEAN UNIT FULL		
BETWEEN FULL	FORCED OUTAGE		
FORCED OUTAGES	DURATION		
(HOURS)	(HOURS)		
212.6	28.1		

WANSLEY IGCC STUDY II FOR D.O.E. (CASE 3)

>>> SCHEDULED OUTAGE HOURS = 412.0
 >>> RESERVE SHUTDOWN HOURS = .0
 >>> SIZE = 150.0 MW

 BASIC SUBSYSTEM INFORMATION NUMBER OF BASIC SUBSYSTEMS = 21

ID#	NAME	THROUGHPUT CAPACITY	NUMBER OF GATES	GATE NUMBER	GATE TYPE	GATE ABOVE
1	BOOST_AC	100.0	1			
2	CATCHALL6	59.5	1	1	-1	0
3	SULFATOR	50.0	1	1	-1	0
4	CHLORIDEGD	59.5	1	1	-1	0
5	RGCOMPRESS	59.5	0			
6	STARHEATR	100.0	1	1	-1	0
7	ZINCFERR	59.5	1	1	-1	0
8	DUMMY	100.0	0			
9	CTHRSG50	100.0	1	1	-1	0
10	BOPSTG	100.0	3	1	-1	0
				2	2	1
				3	2	1
11	BASICBFP	50.0	1	1	-1	0
12	DISCHARGE	50.0	1	1	-1	0
13	BAGHOUSE	100.0	0			
14	VACUUM	100.0	1	1	-1	0
15	SILO	100.0	0			
16	AERATION	100.0	1	1	-1	0
17	CONDITION	100.0	1	1	-1	0
18	DRYUNLOAD	100.0	1	1	-1	0
19	COAL100	100.0	2	1	-1	0
				2	2	1
20	MILLINGSYS	50.0	1	1	-1	0
21	LASTFEED	50.0	1	1	-1	0

 COMPONENT INFORMATION

TOTAL NUMBER OF COMPONENTS = 102

 SUBSYSTEM ID#/NAME

COMPONENT ID#/NAME	GATE ABOVE	MTBF (HOURS)	MDT (HOURS)
1 BOOST AC			
1 BFWPREHEATER	1	10000000.0	1.0
2 TRIMCOOLER	1	99864.0	34.0
3 KNOCKOUTDRUM	1	91104.0	16.0
4 BOOSTERAC	1	26316.0	40.0
5 AIRRECUPERATOR	1	99864.0	34.0
2 CATCHALL6			
6 FEEDSURGEDRUM	1	10989.0	4.0
7 VALVECP1	1	24400.0	2.0
8 VALVECP2	1	66667.0	2.0
9 PRESSZN.AC	1	26313.0	40.0
10 PRSZN.AIRRECEIVER	1	99967.0	34.0
11 COALFEEDVENTFILTER	1	99967.0	8.0
12 PRSZN.FEEDHOPPER	1	4602.0	3.0
13 VALVECP3	1	66667.0	2.0
14 FEEDHOPPER	1	7513.0	4.0
15 COALFEEDER	1	1610.0	2.0
16 GASIFIER	1	40000.0	252.0
17 CYCLONE	1	40000.0	120.0
18 ASHFEEDER	1	1610.0	2.0
19 PRODUCTGASCOOLER	1	8772.0	60.0
20 STEAMDRUM	1	52356.0	50.0
21 FINESCYCLONE	1	40000.0	120.0
22 CYC.FINE.LOCKHOPPER	1	10989.0	8.0
23 VALVEFR1	1	24400.0	2.0
24 VALVEFR2	1	66667.0	2.0
25 CFDEPRESSLOCKHOPPER	1	10989.0	8.0
26 CYCLOCKHOPPERFILTER	1	99978.0	8.0
27 VALVEFR3	1	66667.0	2.0
28 GASFILTER	1	33333.0	40.0
29 VALVEFR5	1	66667.0	2.0
30 G.F.DEPRSSLOCKHOPPER	1	10989.0	8.0
31 GFLOCKHOPPERFILTER	1	17520.0	8.0
32 VALVEFR6	1	66667.0	2.0
33 SOLIDSCONVEYOR	1	11236.0	18.0
34 RECYCLEGASCOOLER	1	91100.0	18.0
35 RECYCLEGASRECEIVER	1	91100.0	18.0
3 SULFATOR			
36 LASHRECEIVER	1	10989.0	80.0
37 SCREWFEEDER	1	17520.0	18.0
38 SULFN.AIRCOMPRESSOR	1	26316.0	40.0
39 SULFATOR	1	40000.0	252.0
40 SULFATORINTERNALS	1	8772.0	60.0
41 SOLIDSDISENGAGER	1	87600.0	34.0
42 SULFATORHEATRECOVERY	1	33288.0	34.0
43 SULFATORSTEAMDRUM	1	66667.0	39.0
44 FLUGASBAGHOUSEFILTER	1	26280.0	50.0
45 FLGASOLIDSCREWCOOLER	1	66667.0	39.0
46 SULFSOLIDSCREWCOOLER	1	11236.0	18.0
4 CHLORIDE GD			
47 VALVECG1	1	17520.0	6.0
48 CHLORIDE GUARDDRUM	1	5300.0	24.0
49 VALVECG2	1	17520.0	6.0
5 RGCOMPRESS			
50 RECYCLEGASCOMPRESSOR	0	13158.0	40.0
6 STARTHEATR			
51 STARTUPHEATER	1	99864.0	120.0
52 ST.UP.HEAT.INTERNAL	1	33333.0	32.0
7 ZINC FERR			
53 VALVEZFDS1	1	21900.0	6.0
54 ZINC FERRITEREACTOR	1	70000.0	250.0
55 ZNFERRITESORBENT	1	10000000.0	1.0

56	VALVEZFDS2	1	21900.0	6.0
57	VALVEREGENER3	1	17520.0	6.0
58	VALVEZFDS4	1	21900.0	6.0
59	ZINCFERRITEREACTOR2	1	70000.0	250.0
60	ZNFERRITESORBENT2	1	10000000.0	1.0
61	VALVEZFDS5	1	21900.0	6.0
62	VALVEREGENER6	1	17520.0	6.0
63	GUARDCYCLONE	1	87600.0	34.0
64	GUARDDRUM	1	87600.0	34.0
65	EXITGASCOOLER	1	99864.0	34.0
8	DUMMY			
66	DUMMYCOMP	0	59475.6	18.0
9	CTHRSG50			
67	CTG	1	4300.0	146.7
68	HRSG	1	1321.0	25.0
10	BOPSTG			
69	CONDPUMP	2	33333.0	31.0
70	CIRCWPUMP	3	100000.0	29.0
71	CONDENSER	1	25000.0	36.0
72	STG	1	5076.0	71.0
11	BASICBFP			
73	BFP	1	10000.0	54.0
12	DISCHARGE			
74	SURGEBIN	1	10989.0	4.2
75	OUTCYLVALVE	1	33744.2	4.0
13	BAGHOUSE			
76	BAGHOUSES	0	730.0	2.0
14	VACUUM			
77	FILTER	1	16404.0	5.3
78	VACPUMP	1	5711.0	28.2
15	SIL0			
79	ASHSILO	0	116508.0	45.5
16	AERATION			
80	BLOWER	1	17786.0	28.2
17	CONDITION			
81	CVALVE	1	3900298.0	4.0
82	ASHCONDITIONER	1	869481.5	4.3
18	DRYUNLOAD			
83	DVALVE	1	3900298.0	4.0
84	UNLOADERBLOWER	1	17786.0	28.2
19	COAL100			
85	RECLAIMCONVEYOR	1	37651.4	18.3
86	TRANSFERCONVEYOR1	1	37651.4	18.3
87	TRANSFERCONVEYOR2	1	37651.4	18.3
88	TRANSFERCONVEYOR3	1	37651.4	18.3
89	TRANSFERCONVEYOR4	1	37651.4	18.3
90	CASCADECONVEYOR	1	37651.4	18.3
91	PNEULIMECONVEYORS	2	27561.7	42.0
92	HOLDINGBIN	1	15705.9	4.2
20	MILLINGSYS			
93	COALSILO	1	109090.9	11.1
94	COALFEEDER1	1	22716.3	7.7
95	COALMILL	1	31856.4	15.5
96	BLOWFAN	1	9408.6	28.0
97	BAGFILTER	1	9521.3	24.0
98	EXHAUSTFAN	1	9408.6	28.0
99	CYCLONE	1	49631.9	16.0
21	LASTFEED			
100	LIMESTONEFEEDER	1	21288.1	7.7
101	COALFEEDER2	1	21288.1	7.7
102	PNEUMATICELEVATOR	1	27235.5	42.0

NESTED SUBSYSTEM INFORMATION

NUMBER OF NESTED SUBSYSTEMS = 17

ID#	NAME	STRUCTURE	CODE	SUBSYSTEMS DIRECTLY CONTAINED			
ID#	NAME	STRUCTURE	CODE	ID#	NAME	STRUCTURE	CODE QUANTITY
22	NCHLRGD100	PARALLEL-IDENT	2	4	CHLORIDEGD	BASIC	1 2
23	NRGCOMPR	PARALLEL-IDENT	2	5	RGCOMPRESS	BASIC	1 3
24	NGAS	SERIAL	4	2	CATCHALL6	BASIC	1 1
				7	ZINCFERR	BASIC	1 1
25	NGASTRAIN2	PARALLEL	3	22	NCHLRGD100	PARLL-IDNT	2 1
26	NGASTRAIN	SERIAL	4	24	NGAS	SERIAL	4 1
				24	NGAS	SERIAL	4 1
27	NSULFATOR	PARALLEL-IDENT	2	25	NGASTRAIN2	PARALLEL	3 1
28	N3BFP50	PARALLEL-IDENT	2	23	NRGCOMPR	PARLL-IDNT	2 1
29	NSURGE	PARALLEL-IDENT	2	3	SULFATOR	BASIC	1 2
30	NBAG	PARALLEL-IDENT	2	11	BASICBFP	BASIC	1 3
31	NVAC	PARALLEL-IDENT	2	12	DISCHARGE	BASIC	1 3
32	NAER	PARALLEL-IDENT	2	13	BAGHOUSE	BASIC	1 2
33	NCONDIT	PARALLEL	3	14	VACUUM	BASIC	1 2
				16	AERATION	BASIC	1 2
34	NASHSYSTEM	SERIAL	4	17	CONDITION	BASIC	1 1
				17	CONDITION	BASIC	1 1
				18	DRYUNLOAD	BASIC	1 1
				15	SILO	BASIC	1 1
				29	NSURGE	PARLL-IDNT	2 1
				30	NBAG	PARLL-IDNT	2 1
				31	NVAC	PARLL-IDNT	2 1
				32	NAER	PARLL-IDNT	2 1
				33	NCONDIT	PARALLEL	3 1
35	NMILL	PARALLEL-IDENT	2	20	MILLINGSYS	BASIC	1 3
36	NFEED	PARALLEL-IDENT	2	21	LASTFEED	BASIC	1 3
37	NCOALLIME	SERIAL	4	19	COAL100	BASIC	1 1
				35	NMILL	PARLL-IDNT	2 1
				36	NFEED	PARLL-IDNT	2 1
38	NPLANT	SERIAL	4	1	BOOST AC	BASIC	1 1
				6	STARTHEATR	BASIC	1 1
				9	CTHRSG50	BASIC	1 1
				10	BOPSTG	BASIC	1 1
				26	NGASTRAIN	SERIAL	4 1
				27	NSULFATOR	PARLL-IDNT	2 1
				28	N3BFP50	PARLL-IDNT	2 1
				34	NASHSYSTEM	SERIAL	4 1
				37	NCOALLIME	SERIAL	4 1

***** RUN 1 *****

BASELINE RUN

FAULT TREE EVALUATION

SUBSYSTEM	MEAN DOWNTIME (HOURS)	MEAN TIME BETWEEN FAILURES (HOURS)	AVAILABILITY
1 BOOST AC	34.41	14471.07	.9976
2 CATCHALL6	12.60	384.21	.9682
3 SULFATOR	56.15	1909.91	.9714
4 CHLORIDEGD	17.23	3302.13	.9948
5 RGCOMPRESS	40.00	13158.00	.9970
6 STARTHEATR	54.05	24991.30	.9978
7 ZINCFERR	28.12	2790.02	.9900
8 DUMMY	18.00	59475.55	.9997
9 CTHRSG50	54.25	1010.55	.9490
10 BOPSTG	65.16	4218.22	.9848
11 BASICBFP	54.00	10000.00	.9946
12 DISCHARGE	4.15	8289.48	.9995
13 BAGHOUSE	2.00	730.00	.9973
14 VACUUM	22.31	4236.19	.9948
15 SILO	45.50	116508.00	.9996
16 AERATION	28.24	17785.99	.9984
17 CONDITION	4.28	710984.00	1.0000
18 DI UNLOAD	28.13	17705.25	.9984
19 COAL100	14.30	4481.55	.9968
20 MILLINGSYS	23.10	2367.84	.9903
21 LASTFEED	17.35	7653.10	.9977

STATE DEFINITIONS WITH CORRESPONDING
 OUTPUT CAPABILITIES

SUBSYSTEM	CURTAIL- MENT LEVEL	STATE AVAILABILITY (PERCENTAGE)	STATE OUTPUT CAPACITY (PERCENTAGE)	STATE DEFINITIONS (SUBSYSTEM IDENTIFICATION NUMBERS AND CURTAILMENT LEVELS)

22 NCHLRGD100				

	0	98.9648	119.	4
	1	1.0325	60.	0
	2	.0027	0.	1

23 NRGCOMPR				

	0	99.0935	179.	5
	1	.9037	119.	0
	2	.0027	60.	1
	3	.0000	0.	2

24 NGAS				

	0	95.8558	60.	2 7 22
	1	4.1442	0.	0 0 -2

25 NGASTRIN2				

				24 24

	0	91.8833	119.	0	0				
	1	3.9725	60.	0	1				
		3.9725	60.	1	0				
	2	.1717	0.	*					

26 NGASTRAIN				25	23				

	0	91.8808	119.	0	-2				
	1	7.9449	60.	1	-3				
		.0025	60.	0	2				
	2	.1717	0.	*					

27 NSULFATOR				3					

	0	94.3695	100.	0					
	1	5.5490	50.	1					
	2	.0816	0.	2					

28 N3BFP50				11					

	0	98.3973	150.	0					
	1	1.5940	100.	1					
	2	.0086	50.	2					
	3	.0000	0.	3					

29 NSURGE				12					

	0	99.8499	150.	0					
	1	.1500	100.	1					
	2	.0001	50.	2					
	3	.0000	0.	3					

30 NBAG				13					

	0	99.4543	200.	0					
	1	.5450	100.	1					
	2	.0007	0.	2					

31 NVAC				14					

	0	98.9550	200.	0					
	1	1.0423	100.	1					
	2	.0027	0.	2					

32 NAER				16					

	0	99.6832	200.	0					
	1	.3165	100.	1					
	2	.0003	0.	2					

33 NCONDIT				17	17	18			

	0	99.8402	300.	0	0	0			
	1	.1586	200.	0	0	1			
		.0006	200.	0	1	0			
		.0006	200.	1	0	0			
	2	.0000	100.	0	1	1			
		.0000	100.	1	0	1			
		.0000	100.	1	1	0			
	3	.0000	0.	*					

34 NASHSYSTEM				15	29	30	31	32	33

	0	99.9571	100.	0	-2	-2	-2	-2	-3
	1	.0001	50.	0	2	-2	-2	-2	-3

-----	2	.0428	0.	*	-----
35 NMILL				20	-----
-----	0	97.1290	150.	0	-----
	1	2.8432	100.	1	-----
	2	.0277	50.	2	-----
	3	.0001	0.	3	-----
36 NFEED				21	-----
-----	0	99.3231	150.	0	-----
	1	.6754	100.	1	-----
	2	.0015	50.	2	-----
	3	.0000	0.	3	-----
37 NCOALLIME				19 35 36	-----
-----	0	99.6527	100.	0 -2 -2	-----
	1	.0277	50.	0 2 -3	-----
		.0015	50.	0 -2 2	-----
	2	.3181	0.	*	-----
38 NPLANT				1 6 9 10 26 27 28 34	-----
-----	0	80.3494	100.	0 0 0 0 0 0 -2 0	-----
	1	6.9500	60.	0 0 0 0 1 0 -2 0	-----
	2	5.1352	50.	0 0 0 0 -2 1 -3 -2	-----
		.0075	50.	0 0 0 0 -2 0 2 -2	-----
		.0001	50.	0 0 0 0 -2 0 -2 1	-----
		.0256	50.	0 0 0 0 -2 0 -2 0	-----
	3	7.5323	0.	*	-----
				37	-----
	0	80.3494	100.	0	-----
	1	6.9500	60.	0	-----
	2	5.1352	50.	-2	-----
		.0075	50.	-2	-----
		.0001	50.	-2	-----
		.0256	50.	1	-----
	3	7.5323	0.	*	-----

NOTE: "*" INDICATES THAT NO SINGLE DEFINITION CAN COMPLETELY DESCRIBE A GIVEN SUBSTATE.
SEE USER'S GUIDE FOR DETAILED EXPLANATION.

PLANT STATE	OUTPUT PROBABILITY	OUTPUT CAPABILITY	DAYS/ YEAR	POWER (MW) OUTPUT
1	80.35%	100.00%	279.48	150.00
2	6.95%	59.50%	24.17	89.25
3	5.17%	50.00%	17.98	75.00
4	7.53%	.00%	26.20	.00

I	PLANT RAM MEASURES	I
I		I
I	EQUIVALENT	I

I	FORCED	FORCED	EQUIVALENT		I
I	OUTAGE	OUTAGE	AVAILABILITY	AVAILABILITY	I
I	RATE	RATE	FACTOR	FACTOR	I
I	7.53%	12.93%	82.97%	88.12%	I

I	UNIT-LEVEL R&M VALUES				I
I	UNIT MEAN TIME	I	MEAN UNIT FULL		I
I	BETWEEN FULL	I	FORCED OUTAGE		I
I	FORCED OUTAGES	I	DURATION		I
I	(HOURS)	I	(HOURS)		I
I	542.7	I	44.2		I

>>> SCHEDULED OUTAGE HOURS - 412.0
 >>> RESERVE SHUTDOWN HOURS - .0
 >>> SIZE - 400.0 MW

 BASIC SUBSYSTEM INFORMATION NUMBER OF BASIC SUBSYSTEMS - 20

ID#	NAME	THROUGHPUT CAPACITY	NUMBER OF GATES	GATE NUMBER	GATE TYPE	GATE ABOVE
1	BOOST_AC	50.0	1			
				1	-1	0
2	CATCHALL6	29.7	1			
				1	-1	0
3	SULFATOR	50.0	1			
				1	-1	0
4	CHLORIDEGD	29.7	1			
				1	-1	0
5	RGCOMPRESS	29.7	0			
6	STARTHEATR	100.0	1			
				1	-1	0
7	ZINCFERR	29.7	1			
				1	-1	0
8	DUMMY	100.0	0			
9	CTHRSG50	25.0	1			
				1	-1	0
10	BASICBFP	12.5	1			
				1	-1	0
11	DISCHARGE	50.0	1			
				1	-1	0
12	BAGHOUSE	100.0	0			
13	VACUUM	100.0	1			
				1	-1	0
14	SILO	100.0	0			
15	AERATION	100.0	1			
				1	-1	0
16	CONDITION	100.0	1			
				1	-1	0
17	DRYUNLOAD	100.0	1			
				1	-1	0
18	COAL100	100.0	2			
				1	-1	0
				2	2	1
19	MILLINGSYS	33.3	1			
				1	-1	0
20	LASTFEED	33.3	1			
				1	-1	0

 COMPONENT INFORMATION

TOTAL NUMBER OF COMPONENTS - 96

SUBSYSTEM ID#/NAME

COMPONENT ID#/NAME	GATE ABOVE	MTBF (HOURS)	MDT (HOURS)
1 BOOST AC			
1 BFWPREHEATER	1	10000000.0	1.0
2 TRIMCOOLER	1	99864.0	34.0
3 KNOCKOUTDRUM	1	91104.0	16.0
4 BOOSTERAC	1	26316.0	40.0
5 AIRRECUPERATOR	1	99864.0	34.0
2 CATCHALL6			
6 FEEDSURGEDRUM	1	10989.0	4.0
7 VALVECP1	1	24400.0	2.0
8 VALVECP2	1	66667.0	2.0
9 PRESSZN.AC	1	26313.0	40.0
10 PRSZN.AIRRECEIVER	1	99967.0	34.0
11 COALFEEDVENTFILTER	1	99967.0	8.0
12 PRSZN.FEEDHOPPER	1	4602.0	3.0
13 VALVECP3	1	66667.0	2.0
14 FEEDHOPPER	1	7513.0	4.0
15 COALFEEDER	1	1610.0	2.0
16 GASIFIER	1	40000.0	252.0
17 CYCLONE	1	40000.0	120.0
18 ASHFEEDER	1	1610.0	2.0
19 FINESCYCLONE	1	40000.0	120.0
20 CYC.FINE.LOCKHOPPER	1	10989.0	8.0
21 VALVEFR1	1	24400.0	2.0
22 VALVEFR2	1	66667.0	2.0
23 CFDEPRESSLOCKHOPPER	1	10989.0	8.0
24 CYCLOCKHOPPERFILTER	1	99978.0	8.0
25 VALVEFR3	1	66667.0	2.0
26 GASFILTER	1	33333.0	40.0
27 VALVEFR5	1	66667.0	2.0
28 G.F.DEPRSSLOCKHOPPER	1	10989.0	8.0
29 GFLOCKHOPPERFILTER	1	17520.0	8.0
30 VALVEFR6	1	66667.0	2.0
31 SOLIDSCONVEYOR	1	11236.0	18.0
32 RECYCLEGASCOOLER	1	91100.0	18.0
33 RECYCLEGASRECEIVER	1	91100.0	18.0
3 SULFATOR			
34 LASHRECEIVER	1	10989.0	80.0
35 SCREWFEEDER	1	17520.0	18.0
36 SULFN.AIRCOMPRESSOR	1	26316.0	40.0
37 SULFATOR	1	40000.0	252.0
38 SULFATORINTERNALS	1	8772.0	60.0
39 SOLIDSDISENGAGER	1	87600.0	34.0
40 SULFATORHEATRECOVERY	1	33288.0	34.0
41 SULFATORSTEAMDRUM	1	66667.0	39.0
42 FLUGASBAGHOUSEFILTER	1	26280.0	50.0
43 FLGASOLIDSCREWCOOLER	1	66667.0	39.0
44 SULFSOLIDSCREWCOOLER	1	11236.0	18.0
4 CHLORIDEGD			
45 VALVECG1	1	17520.0	6.0
46 CHLORIDEGUARDDRUM	1	5300.0	24.0
47 VALVECG2	1	17520.0	6.0
5 RGCOMPRESS			
48 RECYCLEGASCOMPRESSOR	0	13158.0	40.0
6 STARTHEATR			
49 STARTUPHEATER	1	99864.0	120.0

50 ST.UP.HEAT.INTERNAL	1	33333.0	32.0
7 ZINCERR			
51 VALVEZFDS1	1	21900.0	6.0
52 ZINCERRITEREACTOR	1	70000.0	250.0
53 ZNFERRITESORBENT	1	10000000.0	1.0
54 VALVEZFDS2	1	21900.0	6.0
55 VALVEREGENER3	1	17520.0	6.0
56 VALVEZFDS4	1	21900.0	6.0
57 ZINCERRITEREACTOR2	1	70000.0	250.0
58 ZNFERRITESORBENT2	1	10000000.0	1.0
59 VALVEZFDS5	1	21900.0	6.0
60 VALVEREGENER6	1	17520.0	6.0
61 GUARDCYCLONE	1	87600.0	34.0
62 GUARDDRUM	1	87600.0	34.0
63 EXITGASCOOLER	1	99864.0	34.0
8 DUMMY			
64 RECLAIM	0	59475.6	18.0
9 CTHRSG50			
65 CTG	1	4300.0	146.7
66 HRSG	1	1321.0	25.0
10 BASICBFP			
67 BFP	1	10000.0	54.0
11 DISCHARGE			
68 SURGEBIN	1	10989.0	4.2
69 OUTCYLVALVE	1	33744.2	4.0
12 BAGHOUSE			
70 BAGHOUSES	0	730.0	2.0
13 VACUUM			
71 FILTER	1	16404.0	5.3
72 VACPUMP	1	5711.0	28.2
14 SILO			
73 ASHSILO	0	116508.0	45.5
15 AERATION			
74 BLOWER	1	17786.0	28.2
16 CONDITION			
75 CVALVE	1	3900298.0	4.0
76 ASHCONDITIONER	1	869481.5	4.3
17 DRYUNLOAD			
77 DVALVE	1	3900298.0	4.0
78 UNLOADERBLOWER	1	17786.0	28.2
18 COAL100			
79 RECLAIMCONVEYOR	1	37651.4	18.3
80 TRANSFERCONVEYOR1	1	37651.4	18.3
81 TRANSFERCONVEYOR2	1	37651.4	18.3
82 TRANSFERCONVEYOR3	1	37651.4	18.3
83 TRANSFERCONVEYOR4	1	37651.4	18.3
84 CASCADECONVEYOR	1	37651.4	18.3
85 PNEULIMECONVEYORS	2	27561.7	42.0
86 HOLDINGBIN	1	15705.9	4.2
19 MILLINGSYS			
87 COALSILO	1	109090.9	11.1
88 COALFEEDER1	1	22716.3	7.7
89 COALMILL	1	31856.4	15.5
90 BLOWFAN	1	9408.6	28.0
91 BAGFILTER	1	9521.3	24.0
92 EXHAUSTFAN	1	9408.6	28.0
93 CYCLONE	1	49631.9	16.0
20 LASTFEED			
94 LIMESTONEFEEDER	1	21288.1	7.7
95 COALFEEDER2	1	21288.1	7.7

NESTED SUBSYSTEM INFORMATIONNUMBER OF NESTED SUBSYSTEMS - 22

ID#	NAME	STRUCTURE	CODE	SUBSYSTEMS DIRECTLY CONTAINED				
ID#	NAME	STRUCTURE	CODE	QUANTITY				
21	NCHLRGD100	PARALLEL-IDENT	2	4	CHLORIDEGD	BASIC	1	2
22	NRGCOMPR	PARALLEL-IDENT	2	5	RGCOMPRESS	BASIC	1	3
23	NGAS	SERIAL	4	2	CATCHALL6	BASIC	1	1
				7	ZINCFERR	BASIC	1	1
				21	NCHLRGD100	PARLL-IDNT	2	1
24	NGASTRAIN2	PARALLEL	3	23	NGAS	SERIAL	4	1
				23	NGAS	SERIAL	4	1
25	NGASTRAIN	SERIAL	4	22	NRGCOMPR	PARLL-IDNT	2	1
				24	NGASTRAIN2	PARALLEL	3	1
26	NGASTRAINS	PARALLEL	3	25	NGASTRAIN	SERIAL	4	1
				25	NGASTRAIN	SERIAL	4	1
27	NBOOSTAC	PARALLEL-IDENT	2	1	BOOST_AC	BASIC	1	2
28	NSULFATOR	PARALLEL-IDENT	2	3	SULFATOR	BASIC	1	2
29	N3BFP50	PARALLEL-IDENT	2	10	BASICBFP	BASIC	1	3
30	NBFPCTHRSG	SERIAL	4	29	N3BFP50	PARLL-IDNT	2	1
				9	CTHRSG50	BASIC	1	1
31	NPOWER2	PARALLEL	3	30	NBFPCTHRSG	SERIAL	4	1
				30	NBFPCTHRSG	SERIAL	4	1
32	NPOWER4	PARALLEL	3	31	NPOWER2	PARALLEL	3	1
				31	NPOWER2	PARALLEL	3	1
33	NSURGE	PARALLEL-IDENT	2	11	DISCHARGE	BASIC	1	3
34	NBAG	PARALLEL-IDENT	2	12	BAGHOUSE	BASIC	1	2
35	NVAC	PARALLEL-IDENT	2	13	VACUUM	BASIC	1	2
36	NAER	PARALLEL-IDENT	2	15	AERATION	BASIC	1	2
37	NCONDIT	PARALLEL	3	16	CONDITION	BASIC	1	1
				16	CONDITION	BASIC	1	1
				17	DRYUNLOAD	BASIC	1	1
38	NASHSYSTEM	SERIAL	4	14	SILO	BASIC	1	1

			33 NSURGE	PARLL-IDNT	2	1
			34 NBAG	PARLL-IDNT	2	1
			35 NVAC	PARLL-IDNT	2	1
			36 NAER	PARLL-IDNT	2	1
			37 NCONDIT	PARALLEL	3	1
39 NMILL	PARALLEL-IDENT	2				
			19 MILLINGSYS	BASIC	1	4
40 NFEED	PARALLEL-IDENT	2				
			20 LASTFEED	BASIC	1	4
41 NCOALLIME	SERIAL	4				
			18 COAL100	BASIC	1	1
			39 NMILL	PARLL-IDNT	2	1
			40 NFEED	PARLL-IDNT	2	1
42 NPLANT	SERIAL	4				
			6 STARTHEATR	BASIC	1	1
			26 NGASTRANS	PARALLEL	3	1
			27 NBOOSTAC	PARLL-IDNT	2	1
			28 NSULFATOR	PARLL-IDNT	2	1
			32 NPOWER4	PARALLEL	3	1
			38 NASHSYSTEM	SERIAL	4	1
			41 NCOALLIME	SERIAL	4	1

 ***** RUN 1 *****

BASELINE RUN

FAULT TREE EVALUATION

SUBSYSTEM	MEAN DOWNTIME (HOURS)	MEAN TIME BETWEEN FAILURES (HOURS)	AVAILABILITY
1 BOOST_AC	34.41	14471.07	.9976
2 CATCHALL6	10.04	404.92	.9758
3 SULFATOR	56.15	1909.91	.9714
4 CHLORIDEGD	17.23	3302.13	.9948
5 RGCOMPRESS	40.00	13158.00	.9970
6 STARTHEATR	54.05	24991.30	.9978
7 ZINCFERR	28.12	2790.02	.9900
8 DUMMY	18.00	59475.55	.9997
9 CTHRSG50	54.25	1010.55	.9490
10 BASICBFP	54.00	10000.00	.9946
11 DISCHARGE	4.15	8289.48	.9995
12 BAGHOUSE	2.00	730.00	.9973
13 VACUUM	22.31	4236.19	.9948
14 SILO	45.50	116508.00	.9996
15 AERATION	28.24	17785.99	.9984
16 CONDITION	4.28	710984.00	1.0000
17 DRYUNLOAD	28.13	17705.25	.9984
18 COAL100	14.30	4481.55	.9968
19 MILLINGSYS	23.10	2367.84	.9903
20 LASTFEED	17.35	7653.10	.9977

STATE DEFINITIONS WITH CORRESPONDING
OUTPUT CAPABILITIES

SUBSYSTEM	CURTAIL- MENT LEVEL	STATE AVAILABILITY (PERCENTAGE)	STATE OUTPUT CAPACITY (PERCENTAGE)	STATE DEFINITIONS (SUBSYSTEM IDENTIFICATION NUMBERS AND CURTAILMENT LEVELS)	
21 NCHLRGD100				4	
	0	98.9648	59.	0	
	1	1.0325	30.	1	
	2	.0027	0.	2	
22 NRGCOMPR				5	
	0	99.0935	89.	0	
	1	.9037	59.	1	
	2	.0027	30.	2	
	3	.0000	0.	3	
23 NGAS				2 7 21	
	0	96.6036	30.	0 0 -2	
	1	3.3964	0.	*	
24 NGASTRIN2				23 23	
	0	93.3226	59.	0 0	
	1	3.2810	30.	0 1	
		3.2810	30.	1 0	
	2	.1154	0.	*	
25 NGASTRIN				22 24	
	0	93.3200	59.	-2 0	
	1	.0027	30.	2 -2	
		6.5619	30.	-2 1	
	2	.1154	0.	*	
26 NGASTRAINS				25 25	
	0	87.0862	119.	0 0	
	1	6.1261	89.	0 1	
		6.1261	89.	1 0	
	2	.1077	59.	0 2	
		.4309	59.	1 1	
		.1077	59.	2 0	
	3	.0076	30.	1 2	
		.0076	30.	2 1	
	4	.0001	0.	*	
27 NBOOSTAC				1	
	0	99.5261	100.	0	
	1	.4734	50.	1	
	2	.0006	0.	2	

28 NSULFATOR

3

0	94.3695	100.
1	5.5490	50.
2	.0816	0.

0
1
2

29 N3BFP50

10

0	98.3973	38.
1	1.5940	25.
2	.0086	13.
3	.0000	0.

0
1
2
3

30 NBFPCTHRSG

29 9

0	94.8967	25.
1	.0082	13.
2	5.0952	0.

-2	0
2	0
*	

31 NPOWER2

30 30

0	90.0538	50.
1	.0078	38.
	.0078	38.
2	4.8351	25.
	.0000	25.
	4.8351	25.
3	.0004	13.
	.0004	13.
4	.2596	0.

0	0
0	1
1	0
0	2
1	1
2	0
1	2
2	1
*	

32 NPOWER4

31 31

0	81.0968	100.
1	.0140	88.
	.0140	88.
2	8.7085	75.
	.0000	75.
	8.7085	75.
3	.0007	63.
	.0015	63.
	.0015	63.
	.0007	63.
4	.2338	50.
	.0000	50.
	.9351	50.
	.0000	50.
	.2338	50.
5	.0000	38.
	.0001	38.
	.0001	38.
	.0000	38.
6	.0251	25.
	.0000	25.
	.0251	25.
7	.0000	13.
	.0000	13.
8	.0007	0.

0	0
0	1
1	0
0	2
1	1
2	0
0	3
1	2
2	1
3	0
0	4
1	3
2	2
3	1
4	0
1	4
2	3
3	2
4	1
2	4
3	3
4	2
3	4
4	3
*	

33 NSURGE

-----				-----			
	0	99.8499	150.		0		
	1	.1500	100.		1		
	2	.0001	50.		2		
	3	.0000	0.		3		
-----				-----			
34 NBAG				12			
-----				-----			
	0	99.4543	200.		0		
	1	.5450	100.		1		
	2	.0007	0.		2		
-----				-----			
35 NVAC				13			
-----				-----			
	0	98.9550	200.		0		
	1	1.0423	100.		1		
	2	.0027	0.		2		
-----				-----			
36 NAER				15			
-----				-----			
	0	99.6832	200.		0		
	1	.3165	100.		1		
	2	.0003	0.		2		
-----				-----			
37 NCONDIT				16 16 17			
-----				-----			
	0	99.8402	300.		0	0	0
	1	.1586	200.		0	0	1
		.0006	200.		0	1	0
		.0006	200.		1	0	0
	2	.0000	100.		0	1	1
		.0000	100.		1	0	1
		.0000	100.		1	1	0
	3	.0000	0.		*		
-----				-----			
38 NASHSYSTEM				14 33 34 35 36 37			
-----				-----			
	0	99.9571	100.		0	-2	-2
	1	.0001	50.		0	2	-2
	2	.0428	0.		*		
-----				-----			
39 NMILL				19			
-----				-----			
	0	96.1904	133.		0		
	1	3.7543	100.		1		
	2	.0549	67.		2		
	3	.0004	33.		3		
	4	.0000	0.		4		
-----				-----			
40 NFEED				20			
-----				-----			
	0	99.0984	133.		0		
	1	.8985	100.		1		
	2	.0031	67.		2		
	3	.0000	33.		3		
	4	.0000	0.		4		
-----				-----			
41 NCOALLIME				18 39 40			
-----				-----			
	0	95.0201	100.		0	0	0

1	3.7422	100.	0	1	-2
	.8615	100.	0	0	1
2	.0548	67.	0	2	-3
	.0030	67.	0	-2	2
3	.0004	33.	0	3	-4
	.0000	33.	0	-3	3
4	.3180	0.	*		

42 NPLANT

			6	26	27	28	32	38	41
0	62.8655	100.	0	0	0	0	0	0	0
1	3.0459	100.	0	0	0	0	0	0	1
2	9.2731	89.	0	1	0	0	0	0	-2
3	.0259	88.	0	-2	0	0	1	0	-2
4	16.1471	75.	0	-2	0	0	2	0	-2
5	.0530	67.	0	-2	0	0	-3	0	2
6	.0042	63.	0	-2	0	0	3	0	-3
7	.5947	59.	0	2	0	0	-4	0	-3
8	.4700	50.	0	-3	1	-2	-5	-2	-3
	5.4872	50.	0	-3	0	1	-5	-2	-3
	1.3097	50.	0	-3	0	0	4	-2	-3
	.0001	50.	0	-3	0	0	-4	1	-3
9	.0002	38.	0	-3	-2	-2	5	-2	-3
10	.0004	33.	0	-3	-2	-2	-6	-2	3
11	.0150	30.	0	3	-2	-2	-6	-2	-4
12	.0499	25.	0	-4	-2	-2	6	-2	-4
13	.0000	13.	0	-4	-2	-2	7	-2	-4
14	.6581	0.	*						

NOTE: "*" INDICATES THAT NO SINGLE DEFINITION CAN COMPLETELY DESCRIBE A GIVEN SUBSTATE.
SEE USER'S GUIDE FOR DETAILED EXPLANATION.

PLANT STATE	OUTPUT STATE PROBABILITY	OUTPUT CAPABILITY	DAYS/ YEAR	POWER (MW) OUTPUT
----------------	-----------------------------	----------------------	---------------	----------------------

1	62.87%	100.00%	218.67	400.00
2	3.05%	100.00%	10.59	400.00
3	9.27%	89.10%	32.26	356.00
4	.03%	87.50%	.09	350.00
5	16.15%	75.00%	56.17	300.00
6	.05%	66.67%	.18	266.66
7	.00%	62.50%	.01	250.00
8	.59%	59.40%	2.07	237.60
9	7.27%	50.00%	25.28	200.00
10	.00%	37.50%	.00	150.00
11	.00%	33.33%	.00	133.33
12	.02%	29.70%	.05	118.80
13	.05%	25.00%	.17	100.00
14	.00%	12.50%	.00	50.00
15	.66%	.00%	2.29	.00

I	PLANT RAM MEASURES				I
I					I
I	EQUIVALENT				I
I	FORCED	FORCED	EQUIVALENT		I
I	OUTAGE	OUTAGE	AVAILABILITY	AVAILABILITY	I
I	RATE	RATE	FACTOR	FACTOR	I

I	.66%	9.65%	86.10%	94.67%	I

I	UNIT-LEVEL R&M VALUES				I

I	UNIT MEAN TIME	I	MEAN UNIT FULL		I
I	BETWEEN FULL	I	FORCED OUTAGE		I
I	FORCED OUTAGES	I	DURATION		I
I	(HOURS)	I	(HOURS)		I

I	3203.4	I	21.2		I

WANSLEY IGCC STUDY II FOR D.O.E. (CASE 5)

>>> SCHEDULED OUTAGE HOURS = 412.0
 >>> RESERVE SHUTDOWN HOURS = .0
 >>> SIZE = 400.0 MW

 BASIC SUBSYSTEM INFORMATION NUMBER OF BASIC SUBSYSTEMS = 22

ID#	NAME	THROUGHPUT CAPACITY	NUMBER OF GATES	GATE NUMBER	GATE TYPE	GATE ABOVE
1	BOOST_AC	50.0	1			
				1	-1	0
2	CATCHALL6	17.0	1			
				1	-1	0
3	CHLORIDEGD	17.0	1			
				1	-1	0
4	RGCOMPRESS	17.0	0			
5	STARTHEATR	100.0	1			
				1	-1	0
6	ZINCFERR	17.0	1			
				1	-1	0
7	DUMMY	100.0	0			
8	CTHRSG50	50.0	1			
				1	-1	0
9	BOPSTG	100.0	3			
				1	-1	0
				2	2	1
				3	2	1
10	BASICBFP	50.0	1			
				1	-1	0
11	SACIDPLANT	100.0	0			
12	O2PLANT	100.0	0			
13	DISCHARGE	16.7	1			
				1	-1	0
14	BAGHOUSE	100.0	0			
15	VACUUM	100.0	1			
				1	-1	0
16	SILO	100.0	0			
17	AERATION	100.0	1			
				1	-1	0
18	CONDITION	100.0	1			
				1	-1	0
19	DRYUNLOAD	100.0	1			
				1	-1	0
20	COAL100	100.0	1			
				1	-1	0
21	MILLINGSYS	33.3	1			
				1	-1	0
22	LASTFEED	33.3	1			
				1	-1	0

 COMPONENT INFORMATION

TOTAL NUMBER OF COMPONENTS = 91

SUBSYSTEM ID#/NAME

COMPONENT ID#/NAME	GATE ABOVE	MTBF (HOURS)	MDT (HOURS)
1 BOOST AC			
1 BFWPREHEATER	1	10000000.0	1.0
2 TRIMCOOLER	1	99864.0	34.0
3 KNOCKOUTDRUM	1	91104.0	16.0
4 BOOSTERAC	1	26316.0	40.0
5 AIRRECUPERATOR	1	99864.0	34.0
2 CATCHALL6			
6 FEEDSURGEDRUM	1	10989.0	4.0
7 VALVECP1	1	24400.0	2.0
8 VALVECP2	1	66667.0	2.0
9 PRESSZN.AC	1	26313.0	40.0
10 PRSZN.AIRRECEIVER	1	99967.0	34.0
11 COALFEEDVENTFILTER	1	99967.0	8.0
12 PRSZN.FEEDHOPPER	1	4602.0	3.0
13 VALVECP3	1	66667.0	2.0
14 FEEDHOPPER	1	7513.0	4.0
15 COALFEEDER	1	1610.0	2.0
16 GASIFIER	1	40000.0	252.0
17 CYCLONE	1	40000.0	120.0
18 ASHFEEDER	1	1610.0	2.0
19 PRODUCTGASCOOLER	1	8772.0	193.0
20 STEAMDRUM	1	52356.0	50.0
21 FINESCYCLONE	1	40000.0	120.0
22 CYC.FINE.LOCKHOPPER	1	10989.0	8.0
23 VALVEFR1	1	24400.0	2.0
24 VALVEFR2	1	66667.0	2.0
25 CFDEPRESSLOCKHOPPER	1	10989.0	8.0
26 CYCLOCKHOPPERFILTER	1	99978.0	8.0
27 VALVEFR3	1	66667.0	2.0
28 GASFILTER	1	33333.0	40.0
29 VALVEFR5	1	66667.0	2.0
30 G.F.DEPRSSLOCKHOPPER	1	10989.0	8.0
31 GFLOCKHOPPERFILTER	1	17520.0	8.0
32 VALVEFR6	1	66667.0	2.0
33 SOLIDSCONVEYOR	1	11236.0	18.0
34 RECYCLEGASCOOLER	1	91100.0	18.0
35 RECYCLEGASRECEIVER	1	91100.0	18.0
3 CHLORIDEGD			
36 VALVECG1	1	17520.0	6.0
37 CHLORIDEGUARDDRUM	1	5300.0	24.0
38 VALVECG2	1	17520.0	6.0
4 RGCOMPRESS			
39 RECYCLEGASCOMPRESSOR	0	13158.0	40.0
5 STARTHEATR			
40 STARTUPHEATER	1	99864.0	120.0
41 ST.UP.HEAT.INTERNAL	1	33333.0	32.0
6 ZINCERR			
42 VALVEZFDS1	1	21900.0	6.0
43 ZINCERRITEREACTOR	1	70000.0	250.0
44 ZNFERRITESORBENT	1	1000000.0	1.0
45 VALVEZFDS2	1	21900.0	6.0
46 VALVEREGENER3	1	17520.0	6.0

47 VALVEZFDS4	1	21900.0	6.0
48 ZINCFERRITEREACTOR2	1	70000.0	250.0
49 ZNFERRITESORBENT2	1	10000000.0	1.0
50 VALVEZFDS5	1	21900.0	6.0
51 VALVEREGENER6	1	17520.0	6.0
52 GUARDCYCLONE	1	87600.0	34.0
53 GUARDDRUM	1	87600.0	34.0
54 EXITGASCOOLER	1	99864.0	34.0
7 DUMMY			
55 DUMMYCOMP	0	59475.6	18.0
8 CTHRSG50			
56 CTG	1	4300.0	146.7
57 HRSG	1	1321.0	25.0
9 BOPSTG			
58 CONDPUMP	2	33333.0	31.0
59 CIRCWPUMP	3	100000.0	29.0
60 CONDENSER	1	25000.0	36.0
61 STG	1	5076.0	71.0
10 BASICBFP			
62 BFP	1	10000.0	54.0
11 SACIDPLANT			
63 SULFURICACIDPLANT	0	10000000.0	1.0
12 O2PLANT			
64 OXYGENPLANT	0	11042.1	10.5
13 DISCHARGE			
65 SURGEBIN	1	10989.0	4.2
66 OUTCYLVALVE	1	33744.2	4.0
14 BAGHOUSE			
67 BAGHOUSES	0	730.0	2.0
15 VACUUM			
68 FILTER	1	16404.0	5.3
69 VACPUMP	1	5711.0	28.2
16 SILO			
70 ASHSILO	0	116508.0	45.5
17 AERATION			
71 BLOWER	1	17786.0	28.2
18 CONDITION			
72 CVALVE	1	3900298.0	4.0
73 ASHCONDITIONER	1	869481.5	4.3
19 DRYUNLOAD			
74 DVALVE	1	3900298.0	4.0
75 UNLOADERBLOWER	1	17786.0	28.2
20 COAL100			
76 RECLAIMCONVEYOR	1	37651.4	18.3
77 TRANSFERCONVEYOR1	1	37651.4	18.3
78 TRANSFERCONVEYOR2	1	37651.4	18.3
79 TRANSFERCONVEYOR3	1	37651.4	18.3
80 TRANSFERCONVEYOR4	1	37651.4	18.3
81 CASCADECONVEYOR	1	37651.4	18.3
82 HOLDINGBIN	1	15705.9	4.2
21 MILLINGSYS			
83 COALSILO	1	109090.9	11.1
84 COALFEEDER1	1	22716.3	7.7
85 COALMILL	1	31856.4	15.5
86 BLOWFAN	1	9408.6	28.0
87 BAGFILTER	1	9521.3	24.0
88 EXHAUSTFAN	1	9408.6	28.0
89 CYCLONE	1	49631.9	16.0
22 LASTFEED			
90 COALFEEDER2	1	21288.1	7.7

NESTED SUBSYSTEM INFORMATIONNUMBER OF NESTED SUBSYSTEMS = 19

ID#	NAME	STRUCTURE	CODE	SUBSYSTEMS DIRECTLY CONTAINED		
ID#	NAME	STRUCTURE	CODE	QUANTITY		
23	NCHLRGD100	PARALLEL-IDENT	2	3	CHLORIDEGD	BASIC 1 2
24	NRGCOMPR	PARALLEL-IDENT	2	4	RGCOMPRESS	BASIC 1 4
25	NGAS	SERIAL	4	2	CATCHALL6	BASIC 1 1
				6	ZINCFERR	BASIC 1 1
26	NGASTRAIN3	PARALLEL	3	23	NCHLRGD100	PARLL-IDNT 2 1
				25	NGAS	SERIAL 4 1
				25	NGAS	SERIAL 4 1
				25	NGAS	SERIAL 4 1
27	NGASTRAIN	SERIAL	4	26	NGASTRAIN3	PARALLEL 3 1
				24	NRGCOMPR	PARLL-IDNT 2 1
28	NGASTRAINS	PARALLEL	3	27	NGASTRAIN	SERIAL 4 1
				27	NGASTRAIN	SERIAL 4 1
29	NBOOSTAC	PARALLEL-IDENT	2	1	BOOST_AC	BASIC 1 2
30	NCTHRSG	PARALLEL-IDENT	2	8	CTHRSG50	BASIC 1 2
31	N3BFP50	PARALLEL-IDENT	2	10	BASICBFP	BASIC 1 3
32	NSURGE	PARALLEL-IDENT	2	13	DISCHARGE	BASIC 1 6
33	NBAG	PARALLEL-IDENT	2	14	BAGHOUSE	BASIC 1 2
34	NVAC	PARALLEL-IDENT	2	15	VACUUM	BASIC 1 2
35	NAER	PARALLEL-IDENT	2	17	AERATION	BASIC 1 2
36	NCONDIT	PARALLEL	3	18	CONDITION	BASIC 1 1
				18	CONDITION	BASIC 1 1
				19	DRYUNLOAD	BASIC 1 1
37	NASHSYSTEM	SERIAL	4	16	SILO	BASIC 1 1
				32	NSURGE	PARLL-IDNT 2 1
				33	NBAG	PARLL-IDNT 2 1
				34	NVAC	PARLL-IDNT 2 1
				35	NAER	PARLL-IDNT 2 1
				36	NCONDIT	PARALLEL 3 1
38	NMILL	PARALLEL-IDENT	2	21	MILLINGSYS	BASIC 1 4
39	NFEED	PARALLEL-IDENT	2			

40 NCOAL	SERIAL	4	22 LASTFEED	BASIC	1	4
			20 COAL100	BASIC	1	1
			38 NMILL	PARLL-IDNT	2	1
			39 NFEED	PARLL-IDNT	2	1
41 NPLANT	SERIAL	4	5 STARTHEATR	BASIC	1	1
			9 BOPSTG	BASIC	1	1
			11 SACIDPLANT	BASIC	1	1
			12 O2PLANT	BASIC	1	1
			28 NGASTRANS	PARALLEL	3	1
			29 NBOOSTAC	PARLL-IDNT	2	1
			30 NCTHRSG	PARLL-IDNT	2	1
			31 N3BFP50	PARLL-IDNT	2	1
			37 NASHSYSTEM	SERIAL	4	1
			40 NCOAL	SERIAL	4	1

 ***** RUN 1 *****

BASELINE RUN

FAULT TREE EVALUATION

SUBSYSTEM	MEAN DOWNTIME (HOURS)	MEAN TIME BETWEEN FAILURES (HOURS)	AVAILABILITY
1 BOOST AC	34.41	14471.07	.9976
2 CATCHALL6	18.58	384.21	.9539
3 CHLORIDEGD	17.23	3302.13	.9948
4 RGCOMPRESS	40.00	13158.00	.9970
5 STARTHEATR	54.05	24991.30	.9978
6 ZINCFERR	28.12	2790.02	.9900
7 DUMMY	18.00	59475.55	.9997
8 CTHRSG50	54.25	1010.55	.9490
9 BOPSTG	65.16	4218.22	.9848
10 BASICBFP	54.00	10000.00	.9946
11 SACIDPLANT	1.00	1000000.00	.9900
12 O2PLANT	10.50	11042.13	.9991
13 DISCHARGE	4.15	8289.48	.9995
14 BAGHOUSE	2.00	730.00	.9973
15 VACUUM	22.31	4236.19	.9948
16 SILO	45.50	116508.00	.9996
17 AERATION	28.24	17785.99	.9984
18 CONDITION	4.28	710984.00	1.0000
19 DRYUNLOAD	28.13	17705.25	.9984
20 COAL100	14.29	4483.77	.9968
21 MILLINGSYS	23.10	2367.84	.9903
22 LASTFEED	22.75	11948.67	.9981

 STATE DEFINITIONS WITH CORRESPONDING
 OUTPUT CAPABILITIES

SUBSYSTEM	CURTAIL- MENT LEVEL	STATE AVAILABILITY (PERCENTAGE)	STATE OUTPUT CAPACITY (PERCENTAGE)	STATE DEFINITIONS (SUBSYSTEM IDENTIFICATION NUMBERS AND CURTAILMENT LEVELS)		
----- 23 NCHLRGD100 -----				----- 3 -----		
	0	98.9648	34.	0		
	1	1.0325	17.	1		
	2	.0027	0.	2		
----- 24 NRGCOMPR -----				----- 4 -----		
	0	98.7932	68.	0		
	1	1.2013	51.	1		
	2	.0055	34.	2		
	3	.0000	17.	3		
	4	.0000	0.	4		
----- 25 NGAS -----				----- 2 6 23 -----		
	0	94.4337	17.	0	0	-2
	1	5.5663	0.	*		
----- 26 NGASTRIN3 -----				----- 25 25 25 -----		
	0	84.2134	51.	0	0	0
	1	4.9639	34.	0	0	1
		4.9639	34.	0	1	0
		4.9639	34.	1	0	0
	2	.2926	17.	0	1	1
		.2926	17.	1	0	1
		.2926	17.	1	1	0
	3	.0172	0.	*		
----- 27 NGASTRIN -----				----- 26 24 -----		
	0	84.2088	51.	0	-2	
	1	14.8916	34.	1	-3	
		.0046	34.	0	2	
	2	.8778	17.	2	-4	
		.0000	17.	-2	3	
	3	.0172	0.	*		
----- 28 NGASTRINS -----				----- 27 27 -----		
	0	70.9112	102.	0	0	
	1	12.5439	85.	0	1	
		12.5439	85.	1	0	
	2	.7392	68.	0	2	
		2.2190	68.	1	1	
		.7392	68.	2	0	
	3	.0145	51.	0	3	
		.1308	51.	1	2	
		.1308	51.	2	1	
		.0145	51.	3	0	
	4	.0026	34.	1	3	
		.0077	34.	2	2	

		.0026	34.	3	1	
	5	.0002	17.	2	3	
		.0002	17.	3	2	
	6	.0000	0.	*		
-----				-----		
29 NBOOSTAC				1		
-----				-----		
	0	99.5261	100.	0		
	1	.4734	50.	1		
	2	.0006	0.	2		
-----				-----		
30 NCTHRSG				8		
-----				-----		
	0	90.0693	100.	0		
	1	9.6711	50.	1		
	2	.2596	0.	2		
-----				-----		
31 N3BFP50				10		
-----				-----		
	0	98.3973	150.	0		
	1	1.5940	100.	1		
	2	.0086	50.	2		
	3	.0000	0.	3		
-----				-----		
32 NSURGE				13		
-----				-----		
	0	99.7000	100.	0		
	1	.2996	83.	1		
	2	.0004	67.	2		
	3	.0000	50.	3		
	4	.0000	33.	4		
	5	.0000	17.	5		
	6	.0000	0.	6		
-----				-----		
33 NBAG				14		
-----				-----		
	0	99.4543	200.	0		
	1	.5450	100.	1		
	2	.0007	0.	2		
-----				-----		
34 NVAC				15		
-----				-----		
	0	98.9550	200.	0		
	1	1.0423	100.	1		
	2	.0027	0.	2		
-----				-----		
35 NAER				17		
-----				-----		
	0	99.6832	200.	0		
	1	.3165	100.	1		
	2	.0003	0.	2		
-----				-----		
36 NCONDIT				18	18	19
-----				-----		
	0	99.8402	300.	0	0	0
	1	.1586	200.	0	0	1
		.0006	200.	0	1	0
		.0006	200.	1	0	0
	2	.0000	100.	0	1	1
		.0000	100.	1	0	1

	.0000	100.	1	1	0
3	.0000	0.	*		

37 NASHSYSTEM

			16	32	33	34	35	36
0	99.6574	100.	0	0	-2	-2	-2	-3
1	.2995	83.	0	1	-2	-2	-2	-3
2	.0004	67.	0	2	-2	-2	-2	-3
3	.0000	50.	0	3	-2	-2	-2	-3
4	.0000	33.	0	4	-2	-2	-2	-3
5	.0000	17.	0	5	-2	-2	-2	-3
6	.0428	0.	*					

38 NMILL

			21
0	96.1904	133.	0
1	3.7543	100.	1
2	.0549	67.	2
3	.0004	33.	3
4	.0000	0.	4

39 NFEED

			22
0	99.2419	133.	0
1	.7560	100.	1
2	.0022	67.	2
3	.0000	33.	3
4	.0000	0.	4

40 NCOAL

			20	38	39
0	95.1578	100.	0	0	0
1	3.7423	100.	0	1	-2
	.7249	100.	0	0	1
2	.0548	67.	0	2	-3
	.0022	67.	0	-2	2
3	.0004	33.	0	3	-4
	.0000	33.	0	-3	3
4	.3178	0.	*		

41 NPLANT

			5	9	11	12	28	29	30	31
0	59.1747	100.	0	0	0	0	0	0	0	-2
1	2.7779	100.	0	0	0	0	0	0	0	-2
2	21.9183	85.	0	0	0	0	1	0	0	-2
3	.2520	83.	0	0	0	0	-2	0	0	-2
4	3.2399	68.	0	0	0	0	2	0	0	-2
5	.0003	67.	0	0	0	0	-3	0	0	-2
6	.0499	67.	0	0	0	0	-3	0	0	-2
7	.2548	51.	0	0	0	0	3	0	0	-2
8	.0000	50.	0	0	0	0	-4	0	0	-2
9	.4618	50.	0	0	0	0	-4	1	-2	-3
	9.4141	50.	0	0	0	0	-4	0	1	-3
	.0075	50.	0	0	0	0	-4	0	0	2
10	.0125	34.	0	0	0	0	4	-2	-2	-3
11	.0000	33.	0	0	0	0	-5	-2	-2	-3
12	.0004	33.	0	0	0	0	-5	-2	-2	-3
13	.0003	17.	0	0	0	0	5	-2	-2	-3
14	.0000	17.	0	0	0	0	-6	-2	-2	-3
15	2.4356	0.	*							

WANSLEY IGCC STUDY II FOR D.O.E. (CASE 5)

>>> SCHEDULED OUTAGE HOURS = 412.0
 >>> RESERVE SHUTDOWN HOURS = .0
 >>> SIZE = 400.0 MW

 BASIC SUBSYSTEM INFORMATION

NUMBER OF BASIC SUBSYSTEMS = 22

ID#	NAME	THROUGHPUT CAPACITY	NUMBER OF GATES	GATE NUMBER	GATE TYPE	GATE ABOVE
1	BOOST_AC	50.0	1			
				1	-1	0
2	CATCHALL6	17.0	1			
				1	-1	0
3	CHLORIDEGD	17.0	1			
				1	-1	0
4	RGCOMPRESS	17.0	0			
5	STARTHEATR	100.0	1			
				1	-1	0
6	ZINCFERR	17.0	1			
				1	-1	0
7	DUMMY	100.0	0			
8	CTHRSG50	50.0	1			
				1	-1	0
9	BOPSTG	100.0	3			
				1	-1	0
				2	2	1
				3	2	1
10	BASICBFP	50.0	1			
				1	-1	0
11	SACIDPLANT	100.0	0			
12	O2PLANT	100.0	0			
13	DISCHARGE	16.7	1			
				1	-1	0
14	BAGHOUSE	100.0	0			
15	VACUUM	100.0	1			
				1	-1	0
16	SILO	100.0	0			
17	AERATION	100.0	1			
				1	-	0
18	CONDITION	100.0	1			
				1	-1	0
19	DRYUNLOAD	100.0	1			
				1	-1	0
20	COAL100	100.0	1			
				1	-1	0
21	MILLINGSYS	33.3	1			
				1	-1	0
22	LASTFEED	33.3	1			
				1	-1	0

 COMPONENT INFORMATION

TOTAL NUMBER OF COMPONENTS = 91

SUBSYSTEM ID#/NAME

COMPONENT ID#/NAME	GATE ABOVE	MTBF (HOURS)	MDT (HOURS)
1 BOOST AC			
1 BFWPREHEATER	1	10000000.0	1.0
2 TRIMCOOLER	1	99864.0	34.0
3 KNOCKOUTDRUM	1	91104.0	16.0
4 BOOSTERAC	1	26316.0	40.0
5 AIRRECUPERATOR	1	99864.0	34.0
2 CATCHALL6			
6 FEEDSURGEDRUM	1	10989.0	4.0
7 VALVECP1	1	24400.0	2.0
8 VALVECP2	1	66667.0	2.0
9 PRESSZN.AC	1	26313.0	40.0
10 PRSZN.AIRRECEIVER	1	99967.0	34.0
11 COALFEEDVENTFILTER	1	99967.0	8.0
12 PRSZN.FEEDHOPPER	1	4602.0	3.0
13 VALVECP3	1	66667.0	2.0
14 FEEDHOPPER	1	7513.0	4.0
15 COALFEEDER	1	1610.0	2.0
16 GASIFIER	1	40000.0	252.0
17 CYCLONE	1	40000.0	120.0
18 ASHFEEDER	1	1610.0	2.0
19 PRODUCTGASCOOLER	1	8772.0	193.0
20 STEAMDRUM	1	52356.0	50.0
21 FINESCYCLONE	1	40000.0	120.0
22 CYC.FINE.LOCKHOPPER	1	10989.0	8.0
23 VALVEFR1	1	24400.0	2.0
24 VALVEFR2	1	66667.0	2.0
25 CFDEPRESSLOCKHOPPER	1	10989.0	8.0
26 CYCLOCKHOPPERFILTER	1	99978.0	8.0
27 VALVEFR3	1	66667.0	2.0
28 GASFILTER	1	33333.0	40.0
29 VALVEFR5	1	66667.0	2.0
30 G.F.DEPRSSLOCKHOPPER	1	10989.0	8.0
31 GFLOCKHOPPERFILTER	1	17520.0	8.0
32 VALVEFR6	1	66667.0	2.0
33 SOLIDSCONVEYOR	1	11236.0	18.0
34 RECYCLEGASCOOLER	1	91104.0	18.0
35 RECYCLEGASRECEIVER	1	91104.0	18.0
3 CHLORIDEGD			
36 VALVECG1	1	17520.0	6.0
37 CHLORIDEGUARDDRUM	1	5300.0	24.0
38 VALVECG2	1	17520.0	6.0
4 RGCOMPRESS			
39 RECYCLEGASCOMPRESSOR	0	13158.0	40.0
5 STARTHEATR			
40 STARTUPHEATER	1	99864.0	120.0
41 ST.UP.HEAT.INTERNAL	1	33333.0	32.0
6 ZINC FERR			
42 VALVEZFDS1	1	21900.0	6.0
43 ZINC FERRITEREACTOR	1	70000.0	250.0
44 ZNFERRITESORBENT	1	10000000.0	1.0
45 VALVEZFDS2	1	21900.0	6.0
46 VALVEREGENER3	1	17520.0	6.0

47 VALVEZFDS4	1	21900.0	6.0
48 ZINC FERRITEREACTOR2	1	70000.0	250.0
49 ZNFERRITESORBENT2	1	10000000.0	1.0
50 VALVEZFDS5	1	21900.0	6.0
51 VALVEREGENER6	1	17520.0	6.0
52 GUARDCYCLONE	1	87600.0	34.0
53 GUARDDRUM	1	87600.0	34.0
54 EXITGASCOOLER	1	99864.0	34.0
7 DUMMY			
55 DUMMYCOMP	0	59475.6	18.0
8 CTHRSG50			
56 CTG	1	4300.0	146.7
57 HRSG	1	1321.0	25.0
9 BOPSTG			
58 CONDPUMP	2	33333.0	31.0
59 CIRCWPUMP	3	100000.0	29.0
60 CONDENSER	1	25000.0	36.0
61 STG	1	5076.0	71.0
10 BASICBFP			
62 BFP	1	10000.0	54.0
11 SACIDPLANT			
63 SULFURICACIDPLANT	0	10000000.0	1.0
12 O2PLANT			
64 OXYGENPLANT	0	11042.1	10.5
13 DISCHARGE			
65 SURGEBIN	1	10989.0	4.2
66 OUTCYLVALVE	1	33744.2	4.0
14 BAGHOUSE			
67 BAGHOUSES	0	730.0	2.0
15 VACUUM			
68 FILTER	1	16404.0	5.3
69 VACPUMP	1	5711.0	28.2
16 SILO			
70 ASHSILO	0	116508.0	45.5
17 AERATION			
71 BLOWER	1	17786.0	28.2
18 CONDITION			
72 CVALVE	1	3900298.0	4.0
73 ASHCONDITIONER	1	869481.5	4.3
19 DRYUNLOAD			
74 DVALVE	1	3900298.0	4.0
75 UNLOADERBLOWER	1	17786.0	28.2
20 COAL100			
76 RECLAIMCONVEYOR	1	37651.4	18.3
77 TRANSFERCONVEYOR1	1	37651.4	18.3
78 TRANSFERCONVEYOR2	1	37651.4	18.3
79 TRANSFERCONVEYOR3	1	37651.4	18.3
80 TRANSFERCONVEYOR4	1	37651.4	18.3
81 CASCADECONVEYOR	1	37651.4	18.3
82 HOLDINGBIN	1	15705.9	4.2
21 MILLINGSYS			
83 COALSILO	1	109090.9	11.1
84 COALFEEDER1	1	22716.3	7.7
85 COALMILL	1	31856.4	15.5
86 BLOWFAN	1	9408.6	28.0
87 BAGFILTER	1	9521.3	24.0
88 EXHAUSTFAN	1	9408.6	28.0
89 CYCLONE	1	49631.9	16.0
22 LASTFEED			
90 COALFEEDER2	1	21288.1	7.7

91 PNEUMATICELEVATOR

1

27235.5

42.0

NESTED SUBSYSTEM INFORMATIONNUMBER OF NESTED SUBSYSTEMS = 19

ID# NAME STRUCTURE CODE

SUBSYSTEMS DIRECTLY CONTAINED

ID# NAME STRUCTURE CODE QUANTITY

23 NCHLRGD100 PARALLEL-IDENT 2

24 NRGCOMPR PARALLEL-IDENT 2

25 NGAS SERIAL 4

26 NGASTRIN3 PARALLEL 3

27 NGASTRIN SERIAL 4

28 NGASTRINS PARALLEL 3

29 NBOOSTAC PARALLEL-IDENT 2

30 NCTHRSG PARALLEL IDENT 2

31 N3BFP50 PARALLEL-IDENT 2

32 NSURGE PARALLEL-IDENT 2

33 NBAG PARALLEL-IDENT 2

34 NVAC PARALLEL-IDENT 2

35 NAER PARALLEL-IDENT 2

36 NCONDIT PARALLEL 3

37 NASHSYSTEM SERIAL 4

38 NMILL PARALLEL-IDENT 2

39 NFEED PARALLEL-IDENT 2

3 CHLORIDEGD BASIC 1 2

4 RGCOMPRESS BASIC 1 4

2 CATCHALL6 BASIC 1 1

6 ZINCFERR BASIC 1 1

23 NCHLRGD100 PARLL-IDNT 2 1

25 NGAS SERIAL 4 1

25 NGAS SERIAL 4 1

25 NGAS SERIAL 4 1

26 NGASTRIN3 PARALLEL 3 1

24 NRGCOMPR PARLL-IDNT 2 1

27 NGASTRIN SERIAL 4 1

27 NGASTRIN SERIAL 4 1

1 BOOST_AC BASIC 1 2

8 CTHRSG50 BASIC 1 2

10 BASICBFP BASIC 1 3

13 DISCHARGE BASIC 1 6

14 BAGHOUSE BASIC 1 2

15 VACUUM BASIC 1 2

17 AERATION BASIC 1 2

18 CONDITION BASIC 1 1

18 CONDITION BASIC 1 1

19 DRYUNLOAD BASIC 1 1

16 SILO BASIC 1 1

32 NSURGE PARLL-IDNT 2 1

33 NBAG PARLL-IDNT 2 1

34 NVAC PARLL-IDNT 2 1

35 NAER PARLL-IDNT 2 1

36 NCONDIT PARALLEL 3 1

21 MILLINGSYS BASIC 1 4

40 NCOAL	SERIAL	4	22 LASTFEED	BASIC	1	4
			20 COAL100	BASIC	1	1
			38 NMILL	PARLL-IDNT	2	1
41 NPLANT	SERIAL	4	39 NFEED	PARLL-IDNT	2	1
			5 STARTHEATR	BASIC	1	1
			9 BOPSTG	BASIC	1	1
			11 SACIDPLANT	BASIC	1	1
			12 O2PLANT	BASIC	1	1
			28 NGASTRANS	PARALLEL	3	1
			29 NBOOSTAC	PARLL-IDNT	2	1
			30 NCTHRSG	PARLL-IDNT	2	1
			31 N3BFP50	PARLL-IDNT	2	1
			37 NASHSYSTEM	SERIAL	4	1
			40 NCOAL	SERIAL	4	1

 ***** RUN 1 *****

BASELINE RUN

FAULT TREE EVALUATION

SUBSYSTEM	MEAN DOWNTIME (HOURS)	MEAN TIME BETWEEN FAILURES (HOURS)	AVAILABILITY
1 BOOST AC	34.41	14471.07	.9976
2 CATCHALL6	18.58	384.21	.9539
3 CHLORIDEGD	17.23	3302.13	.9948
4 RGCOMPRESS	40.00	13158.00	.9970
5 STARTHEATR	54.05	24991.30	.9978
6 ZINCERR	28.12	2790.02	.9900
7 DUMMY	18.00	59475.55	.9997
8 CTHRSG50	54.25	1010.55	.9490
9 BOPSTG	65.16	4218.22	.9848
10 BASICBFP	54.00	10000.00	.9946
11 SACIDPLANT	1.00	10000000.00	1.0000
12 O2PLANT	10.50	11042.13	.9991
13 DISCHARGE	4.15	8289.48	.9995
14 BAGHOUSE	2.00	730.00	.9973
15 VACUUM	22.31	4236.19	.9948
16 SILO	45.50	116508.00	.9996
17 AERATION	28.24	17785.99	.9984
18 CONDITION	4.28	710984.00	1.0000
19 DRYUNLOAD	28.13	17705.25	.9984
20 COAL100	14.29	4483.77	.9968
21 MILLINGSYS	23.10	2367.84	.9903
22 LASTFEED	22.75	11948.67	.9981

STATE DEFINITIONS WITH CORRESPONDING
 OUTPUT CAPABILITIES

SUBSYSTEM	CURTAIL- MENT LEVEL	STATE AVAILABILITY (PERCENTAGE)	STATE OUTPUT CAPACITY (PERCENTAGE)	STATE DEFINITIONS (SUBSYSTEM IDENTIFICATION NUMBERS AND CURTAILMENT LEVELS)		
----- 23 NCHLRGD100 -----				----- 3 -----		
	0	98.9648	34.	0		
	1	1.0325	17.	1		
	2	.0027	0.	2		
----- 24 NRGCOMPR -----				----- 4 -----		
	0	98.7932	68.	0		
	1	1.2013	51.	1		
	2	.0055	34.	2		
	3	.0000	17.	3		
	4	.0000	0.	4		
----- 25 NGAS -----				----- 2 6 23 -----		
	0	94.4337	17.	0	0	-2
	1	5.5663	0.	*		
----- 26 NGASTRAIN3 -----				----- 25 25 25 -----		
	0	84.2134	51.	0	0	0
	1	4.9639	34.	0	0	1
		4.9639	34.	0	1	0
		4.9639	34.	1	0	0
	2	.2926	17.	0	1	1
		.2926	17.	1	0	1
		.2926	17.	1	1	0
	3	.0172	0.	*		
----- 27 NGASTRAIN -----				----- 26 24 -----		
	0	84.2088	51.	0	-2	
	1	14.8916	34.	1	-3	
		.0046	34.	0	2	
	2	.8778	17.	2	-4	
		.0000	17.		3	
	3	.0172	0.			
----- 28 NGASTRAINS -----				----- 27 27 -----		
	0	70.9112	102.	0	0	
	1	12.5439	85.	0	1	
		12.5439	85.	1	0	
	2	.7392	68.	0	2	
		2.2190	68.	1	1	
		.7392	68.	2	0	
	3	.0145	51.	0	3	
		.1308	51.	1	2	
		.1308	51.	2	1	
		.0145	51.	3	0	
	4	.0026	34.	1	3	
		.0077	34.	2	2	

		.0026	34.	3	1	
	5	.0002	17.	2	3	
		.0002	17.	3	2	
	6	.0000	0.	*		
-----				-----		
29 NBOOSTAC				1		
-----				-----		
	0	99.5261	100.	0		
	1	.4734	50.	1		
	2	.0006	0.	2		
-----				-----		
30 NCTHRSG				8		
-----				-----		
	0	90.0693	100.	0		
	1	9.6711	50.	1		
	2	.2596	0.	2		
-----				-----		
31 N3BFP50				10		
-----				-----		
	0	98.3973	150.	0		
	1	1.5940	100.	1		
	2	.0086	50.	2		
	3	.0000	0.	3		
-----				-----		
32 NSURGE				13		
-----				-----		
	0	99.7000	100.	0		
	1	.2996	83.	1		
	2	.0004	67.	2		
	3	.0000	50.	3		
	4	.0000	33.	4		
	5	.0000	.7.	5		
	6	.0000	0.	6		
-----				-----		
33 NBAG				14		
-----				-----		
	0	99.4543	200.	0		
	1	.5450	100.	1		
	2	.0007	0.	2		
-----				-----		
34 NVAC				15		
-----				-----		
	0	98.9550	200.	0		
	1	1.0423	100.	.		
	2	.0027	0.			
-----				-----		
35 NAER				17		
-----				-----		
	0	99.6832	200.	0		
	1	.3165	100.	1		
	2	.0003	0.	2		
-----				-----		
36 NCONDIT				18	18	19
-----				-----		
	0	99.8402	300.	0	0	0
	1	.1586	200.	0	0	1
		.0006	200.	0	1	0
		.0006	200.	1	0	0
	2	.0000	100.	0	1	1
		.0000	100.	1	0	1

	.0000	100.
3	.0000	0.

1	1	0
*		

37 NASHSYSTEM

0	99.6574	100.
1	.2995	83.
2	.0004	67.
3	.0000	50.
4	.0000	33.
5	.0000	17.
6	.0428	0.

16	32	33	34	35	36
----	----	----	----	----	----

0	0	-2	-2	-2	-3
0	1	-2	-2	-2	-3
0	2	-2	-2	-2	-3
0	3	-2	-2	-2	-3
0	4	-2	-2	-2	-3
0	5	-2	-2	-2	-3
*					

38 NMILL

0	96.1904	133.
1	3.7543	100.
2	.0549	67.
3	.0004	33.
4	.0000	0.

21

0
1
2
3
4

39 NFEED

0	99.2419	133.
1	.7560	100.
2	.0022	67.
3	.0000	33.
4	.0000	0.

22

0
1
2
3
4

40 NCOAL

0	95.1578	100.
1	3.7423	100.
	.7249	100.
2	.0548	67.
	.0022	67.
3	.0004	33.
	.0000	33.
4	.3178	0.

20	38	39
----	----	----

0	0	0
0	1	-2
0	0	1
0	2	-3
0	-2	2
0	3	-4
0	-3	3
*		

41 NPLANT

0	59.1747	100.
1	2.7779	100.
2	21.9183	85.
3	.2520	83.
4	3.2399	68.
5	.0003	67.
6	.0499	67.
7	.2548	51.
8	.0000	50.
9	.4618	50.
	9.4141	50.
	.0075	50.
10	.0125	34.
11	.0000	33.
12	.0004	33.
13	.0003	17.
14	.0000	17.
15	2.4356	0.

5	9	11	12	28	29	30	31
---	---	----	----	----	----	----	----

0	0	0	0	0	0	0	-2
0	0	0	0	0	0	0	-2
0	0	0	0	1	0	0	-2
0	0	0	0	-2	0	0	-2
0	0	0	0	2	0	0	-2
0	0	0	0	-3	0	0	-2
0	0	0	0	-3	0	0	-2
0	0	0	0	3	0	0	-2
0	0	0	0	-4	0	0	-2
0	0	0	0	-4	1	-2	-3
0	0	0	0	-4	0	1	-3
0	0	0	0	-4	0	0	2
0	0	0	0	4	-2	-2	-3
0	0	0	0	-5	-2	-2	-3
0	0	0	0	-5	-2	-2	-3
0	0	0	0	5	-2	-2	-3
0	0	0	0	-6	-2	-2	-3
*							

			37	40
0	59.1747	100.	0	0
1	2.7779	100.	0	1
2	21.9183	85.	0	-2
3	.2520	83.	1	-2
4	3.2399	68.	-2	-2
5	.0003	67.	2	-2
6	.0499	67.	-3	2
7	.2548	51.	-3	-3
8	.0000	50.	3	-3
9	.4618	50.	-4	-3
	9.4141	50.	-4	-3
	.0075	50.	-4	-3
10	.0125	34.	-4	-3
11	.0000	33.	4	-3
12	.0004	33.	-5	3
13	.0003	17.	-5	-4
14	.0000	17.	5	-4
15	2.4356	0.	*	

NOTE: "*" INDICATES THAT NO SINGLE DEFINITION CAN COMPLETELY DESCRIBE A GIVEN SUBSTATE.
SEE USER'S GUIDE FOR DETAILED EXPLANATION.

PLANT STATE	OUTPUT PROBABILITY	OUTPUT CAPABILITY	DAYS/ YEAR	POWER (MW) OUTPUT
1	59.17%	100.00%	205.83	400.00
2	2.78%	100.00%	9.66	400.00
3	21.92%	85.00%	76.24	340.00
4	.25%	83.35%	.88	333.40
5	3.24%	68.00%	11.27	272.00
6	.00%	66.68%	.00	266.72
7	.05%	66.67%	.17	266.66
8	.25%	51.00%	.89	204.00
9	.00%	50.01%	.00	200.04
10	9.88%	50.00%	34.38	200.00
11	.01%	34.00%	.04	136.00
12	.00%	33.34%	.00	133.36
13	.00%	33.33%	.00	133.33
14	.00%	17.00%	.00	68.00
15	.00%	16.67%	.00	66.68
16	2.44%	.00%	8.47	.00

PLANT RAM MEASURES				
I				I
I				I
I				I
I				I
I	FORCED	EQUIVALENT		I
I	OUTAGE	OUTAGE	EQUIVALENT	I
I	RATE	RATE	AVAILABILITY FACTOR	I
I			AVAILABILITY FACTOR	I

I	2.44%	11.89%	83.96%	92.98%	I
---	-------	--------	--------	--------	---

I	UNIT-LEVEL R&M VALUES				I
I	UNIT MEAN TIME	I	MEAN UNIT FULL	I	
I	BETWEEN FULL	I	FORCED OUTAGE	I	
I	FORCED OUTAGES	I	DURATION	I	
I	(HOURS)	I	(HOURS)	I	
I	1416.9	I	35.4	I	

>>> SCHEDULED OUTAGE HOURS = 412.0
 >>> RESERVE SHUTDOWN HOURS = .0
 >>> SIZE = 400.0 MW

 BASIC SUBSYSTEM INFORMATION NUMBER OF BASIC SUBSYSTEMS = 24

ID#	NAME	THROUGHPUT CAPACITY	NUMBER OF GATES	GATE NUMBER	GATE TYPE	GATE ABOVE
1	COAL&LIME	100.0	1			
2	BOOSTERAC	50.0	1	1	-1	0
3	TRANSPGAS	42.0	1	1	-1	0
4	TRNGASCYCL	42.0	1	1	-1	0
5	RECYCLEGAS	42.0	1	1	-1	0
6	SULFATOR	42.0	1	1	-1	0
7	SULFACYCL	42.0	1	1	-1	0
8	BAYONET	42.0	0			
9	ASHFILTERS	42.0	1	1	-1	0
10	BLOWBACK	42.0	0			
11	ASHHOPPER	42.0	1	1	-1	0
12	DISCHARGE	50.0	1	1	-1	0
13	BAGHOUSE	100.0	0			
14	VACUUM	100.0	1	1	-1	0
15	SILO	100.0	0			
16	AERATION	100.0	1	1	-1	0
17	CONDITION	100.0	1	1	-1	0
18	DRYUNLOAD	100.0	1	1	-1	0
19	COAL100	100.0	2	1	-1	0
				2	2	1
20	MILLINGSYS	25.0	1	1	-1	0
21	LASTFEED	25.0	1	1	-1	0
22	CTHRSG50	50.0	1	1	-1	0
23	BOPSTG	100.0	3	1	-1	0
				2	2	1

24 BASICBFP	50.0	1	3	2	1
			1	-1	0

COMPONENT INFORMATION

TOTAL NUMBER OF COMPONENTS = 82

SUBSYSTEM ID#/NAME

-----	GATE		
COMPONENT ID#/NAME	ABOVE	MTBF (HOURS)	MDT (HOURS)
-----	-----	-----	-----
1 COAL&LIME			
1 LIMESURGEBIN	1	6802.0	3.7
2 LIMEPRESSHOPPER	1	4305.0	3.8
3 LIMEFVENTFILTER	1	99967.0	8.0
4 LIMEFEEDHOPPER	1	7513.0	4.4
5 CFSURGEBIN	1	6802.0	3.7
6 COALPRESSHOPPER	1	4305.0	3.8
7 CFVENTFILTER	1	99967.0	8.0
8 COALFEEDHOPPER	1	7513.0	4.4
9 COALFEEDER	1	1610.0	2.0
10 LIMEFEEDER	1	1610.0	2.0
11 LOCKHOPPERAC	1	26313.0	38.0
12 AIRRECEIVER	1	99967.0	34.0
13 STARTUPHEATER	1	99864.0	120.0
14 SUHEATERINTERNALS	1	33333.0	32.0
2 BOOSTERAC			
15 BOOSTERAIRCOMP	1	26280.0	42.0
16 AIRRECUPERATOR	1	99864.0	34.0
17 BFWPREHEATER	1	1000000.0	1.0
18 TRIMCOOLER	1	99864.0	34.0
19 KNOCKOUTDRUM	1	91104.0	16.0
3 TRANSPGAS			
20 TRANSPORTGASIFIER	1	40000.0	252.0
21 STANDPIPE	1	40000.0	212.0
4 TRNGASCYCL			
22 STANDPIPECYCLONE	1	40000.0	100.0
23 PRIMARYCYCLONE	1	22000.0	150.0
24 SECONDARYCYCLONE	1	22000.0	150.0
5 RECYCLEGAS			
25 PRODUCTGASCOOLER	1	8.0	60.0
26 STEAMDRUM	1	5235.0	50.0
27 GASFILTER	1	33333.0	40.0
28 GFILTERVALVE	1	24400.0	2.0
29 RECYCLEGASCOOLER	1	91100.0	18.0
30 RECYCLEGASCOMPR	1	13200.0	40.0
31 RECYCLEGASRECEIVER	1	91100.0	18.0
6 SULFATOR			
32 FLUIDBED2NDCYCLONE	1	40000.0	100.0
33 FLUIDBED1STCYCLONE	1	40000.0	100.0
34 FBHEATXCHANGE	1	22000.0	225.0
35 HPASHCOOLER	1	40000.0	225.0
36 ASHCOOLERCYCLONE	1	40000.0	40.0
37 SULFATORSTEAMDRUM	1	40000.0	225.0
38 TRANSPORTSULFATOR	1	40000.0	250.0
7 SULFACYCL			

39	SULFATORCYCLONE1	1	22000.0	150.0
40	SULFATORCYCLONE2	1	22000.0	150.0
8	BAYONET			
41	BAYONETBOILER	0	22000.0	150.0
9	ASHFILTERS			
42	ASHVENTFILTER	1	43800.0	16.0
43	ASHSCREWCONVEYOR	1	66700.0	39.0
44	SULFBARRIERFILTER	1	10000000.0	1.0
45	AIRRECEIVER	1	99967.0	8.0
10	BLOWBACK			
46	BLOWBACKAC	0	26280.0	42.0
11	ASHHOPPER			
47	ASHDEPRESZNHOPPER	1	10989.0	8.0
12	DISCHARGE			
48	SURGEBIN	1	10989.0	4.2
49	OUTCYLVALVE	1	33744.2	4.0
13	BAGHOUSE			
50	BAGHOUSES	0	730.0	2.0
14	VACUUM			
51	FILTER	1	16404.0	5.3
52	VACPUMP	1	5711.0	28.2
15	SIL0			
53	ASHSILO	0	116508.0	45.5
16	AERATION			
54	BLOWER	1	17786.0	28.2
17	CONDITION			
55	CVALVE	1	3900298.0	4.0
56	ASHCONDITIONER	1	869481.5	4.3
18	DRYUNLOAD			
57	DVALVE	1	3900298.0	4.0
58	UNLOADERBLOWER	1	17786.0	28.2
19	COAL100			
59	RECLAIMCONVEYOR	1	37651.4	18.3
60	TRANSFERCONVEYOR1	1	37651.4	18.3
61	TRANSFERCONVEYOR2	1	37651.4	18.3
62	TRANSFERCONVEYOR3	1	37651.4	18.3
63	TRANSFERCONVEYOR4	1	37651.4	18.3
64	CASCADECONVEYOR	1	37651.4	18.3
65	PNEULIMECONVEYORS	2	27235.5	42.0
66	HOLDINGBIN	1	15705.9	4.2
20	MILLINGSYS			
67	COALSILO	1	109090.9	11.1
68	COALFEEDER1	1	22716.3	7.7
69	COALMILL	1	26799.4	30.0
70	BAGFILTER	1	9521.3	24.0
71	BLOWFAN	1	9408.6	28.0
72	EXHAUSTFAN	1	9408.6	28.0
73	CYCLONE	1	49631.9	16.0
21	LASTFEED			
74	COALFEEDER2	1	21288.1	7.7
75	PNEUMATICELEVATOR	1	27235.5	42.0
22	CTHRSG50			
76	CTG	1	4300.0	146.7
77	HRSRG	1	1321.0	25.0
23	BOPSTG			
78	CONDPUMP	2	33333.0	31.0
79	CIRCPUMP	3	100000.0	29.0
80	CONDENSER	1	25000.0	36.0
81	STG	1	5076.0	71.0
24	BASICBFP			

 NESTED SUBSYSTEM INFORMATION
 NUMBER OF NESTED SUBSYSTEMS = 25

ID#	NAME	STRUCTURE	CODE	SUBSYSTEMS DIRECTLY CONTAINED				
ID#	NAME	STRUCTURE	CODE	QUANTITY				
25	NTRANCYCL	PARALLEL-IDENT	2	4	TRNGASCYCL	BASIC	1	2
26	NTRANSPORT	SERIAL	4	25	NTRANCYCL	PARLL-IDNT	2	1
				3	TRANSPGAS	BASIC	1	1
27	NSULFACYCL	PARALLEL-IDENT	2	7	SULFACYCL	BASIC	1	2
28	NBAYONET	PARALLEL-IDENT	2	8	BAYONET	BASIC	1	2
29	NSULFATOR	SERIAL	4	6	SULFATOR	BASIC	1	1
				27	NSULFACYCL	PARLL-IDNT	2	1
				28	NBAYONET	PARLL-IDNT	2	1
30	NBLOWBACK	PARALLEL-IDENT	2	10	BLOWBACK	BASIC	1	2
31	NASHHOP	PARALLEL-IDENT	2	11	ASHHOPPER	BASIC	1	2
32	NASHFILTER	SERIAL	4	9	ASHFILTERS	BASIC	1	1
				30	NBLOWBACK	PARLL-IDNT	2	1
				31	NASHHOP	PARLL-IDNT	2	1
33	TRANSPORT2	PARALLEL	3	26	NTRANSPORT	SERIAL	4	1
				26	NTRANSPORT	SERIAL	4	1
				26	NTRANSPORT	SERIAL	4	1
34	NBOOSTERAC	PARALLEL-IDENT	2	2	BOOSTERAC	BASIC	1	2
35	SULFATOR2	PARALLEL	3	29	NSULFATOR	SERIAL	4	1
				29	NSULFATOR	SERIAL	4	1
				29	NSULFATOR	SERIAL	4	1
36	NRECYCLE	PARALLEL-IDENT	2	5	RECYCLEGAS	BASIC	1	3
37	ASHFILTER2	PARALLEL	3	32	NASHFILTER	SERIAL	4	1
				32	NASHFILTER	SERIAL	4	1
				32	NASHFILTER	SERIAL	4	1
38	NSURGE	PARALLEL-IDENT	2	12	DISCHARGE	BASIC	1	3
39	NBAG	PARALLEL-IDENT	2	13	BAGHOUSE	BASIC	1	2
40	NVAC	PARALLEL-IDENT	2	14	VACUUM	BASIC	1	2
41	NAER	PARALLEL-IDENT	2	16	AERATION	BASIC	1	2
42	NCONDIT	PARALLEL	3					

			17	CONDITION	BASIC	1	1
			17	CONDITION	BASIC	1	1
			18	DRYUNLOAD	BASIC	1	1
43	NASHSYSTEM	SERIAL	4				
			15	SILO	BASIC	1	1
			38	NSURGE	PARLL-IDNT	2	1
			39	NBAG	PARLL-IDNT	2	1
			40	NVAC	PARLL-IDNT	2	1
			41	NAER	PARLL-IDNT	2	1
			42	NCONDIT	PARALLEL	3	1
44	NMILL	PARALLEL-IDENT	2				
45	NFEED	PARALLEL-IDENT	2				
46	NCOALLIME	SERIAL	4				
			20	MILLINGSYS	BASIC	1	5
			21	LASTFEED	BASIC	1	5
			19	COAL100	BASIC	1	1
			44	NMILL	PARLL-IDNT	2	1
			45	NFEED	PARLL-IDNT	2	1
47	NCTHRSG	PARALLEL-IDENT	2				
48	N3BFP50	PARALLEL-IDENT	2				
49	NPLANT	SERIAL	4				
			22	CTHRSG50	BASIC	1	2
			24	BASICBFP	BASIC	1	3
			1	COAL&LIME	BASIC	1	1
			33	TRANSPORT2	PARALLEL	3	1
			37	ASHFILTER2	PARALLEL	3	1
			35	SULFATOR2	PARALLEL	3	1
			34	NBOOSTERAC	PARLL-IDNT	2	1
			36	NRECYCLE	PARLL-IDNT	2	1
			43	NASHSYSTEM	SERIAL	4	1
			46	NCOALLIME	SERIAL	4	1
			23	BOPSTG	BASIC	1	1
			47	NCTHRSG	PARLL-IDNT	2	1
			48	N3BFP50	PARLL-IDNT	2	1

 ***** RUN 1 *****

BASELINE RUN

----- FAULT TREE EVALUATION

SUBSYSTEM	MEAN DOWNTIME (HOURS)	MEAN TIME BETWEEN FAILURES (HOURS)	AVAILABILITY
1 COAL&LIME	4.49	421.04	.9894
2 BOOSTERAC	35.52	14460.18	.9975
3 TRANSPGAS	232.67	20000.00	.9885
4 TRNGASCYCL	139.91	8627.45	.9840
5 RECYCLEGAS	41.59	3313.51	.9876
6 SULFATOR	174.92	5116.28	.9669
7 SULFACYCL	150.51	11000.00	.9865
8 BAYONET	150.00	22000.00	.9932

9	ASHFILTERS	21.50	20865.13	.9990
10	BLOWBACK	42.00	26280.00	.9984
11	ASHHOPPER	8.00	10989.00	.9993
2	DISCHARGE	4.15	8289.48	.9995
3	BAGHOUSE	2.00	730.00	.9973
14	VACUUM	22.31	4236.19	.9948
15	SILO	45.50	116508.00	.9996
16	AERATION	28.24	17785.99	.9984
17	CONDITION	4.28	710984.00	1.0000
18	DRYUNLOAD	28.13	17705.25	.9984
19	COAL100	14.30	4481.50	.9968
20	MILLINGSYS	24.28	2335.09	.9897
21	LASTFEED	22.75	11948.67	.9981
22	CTHRSG50	54.25	1010.55	.9490
23	BOPSTG	65.16	4218.22	.9848
24	BASICBFP	54.00	10000.00	.9946

STATE DEFINITIONS WITH CORRESPONDING
OUTPUT CAPABILITIES

SUBSYSTEM	CURTAIL- MENT LEVEL	STATE AVAILABILITY (PERCENTAGE)	STATE OUTPUT CAPACITY (PERCENTAGE)	STATE DEFINITIONS (SUBSYSTEM IDENTIFICATION NUMBERS AND CURTAILMENT LEVELS)

25 NTRANCYCL				4

	0	96.8338	84.	0
	1	3.1407	42.	1
	2	.0255	0.	2

26 NTRANSPORT				25 3

	0	98.8249	42.	-2 0
	1	1.1751	0.	*

27 NSULFACYCL				7

	0	97.3186	84.	0
	1	2.6632	42.	1
	2	.0182	0.	2

28 NBAYONET				8

	0	98.6502	84.	0
	1	1.3452	42.	1
	2	.0046	0.	2

29 NSULFATOR				6 27 28

	0	96.6721	42.	0 -2 -2
	1	3.3279	0.	*

30 NBLOWBACK				10

	0	99.6811	84.	0
	1	.3186	42.	1
	2	.0003	0.	2

-----				-----			
31 NASHHOP				11			
-----				-----			
	0	99.8546	84.		0		
	1	.1454	42.		1		
	2	.0001	0.		2		
-----				-----			
32 NASHFILTER				9 30 31			
-----				-----			
	0	99.8968	42.		0	-2	-2
	1	.1032	0.		*		
-----				-----			
33 TRANSPORT2				26 26 26			
-----				-----			
	0	96.5159	126.		0	0	0
	1	1.1477	84.		0	0	1
		1.1477	84.		0	1	0
		1.1477	84.		1	0	0
	2	.0136	42.		0	1	1
		.0136	42.		1	0	1
		.0136	42.		1	1	0
	3	.0002	0.		*		
-----				-----			
34 NBOOSTERAC				2			
-----				-----			
	0	99.5105	100.		0		
	1	.4889	50.		1		
	2	.0006	0.		2		
-----				-----			
35 SULFATOR2				29 29 29			
-----				-----			
	0	90.3448	126.		0	0	0
	1	3.1101	84.		0	0	1
		3.1101	84.		0	1	0
		3.1101	84.		1	0	0
	2	.1071	42.		0	1	1
		.1071	42.		1	0	1
		.1071	42.		1	1	0
	3	.0037	0.		*		
-----				-----			
36 NRECYCLE				5			
-----				-----			
	0	96.3272	126.		0		
	1	3.6271	84.				
	2	.0455	42.				
	3	.0002	0.				
-----				-----			
37 ASHFILTER2				32 32 32			
-----				-----			
	0	99.6906	126.		0	0	0
	1	.1030	84.		0	0	1
		.1030	84.		0	1	0
		.1030	84.		1	0	0
	2	.0001	42.		0	1	1
		.0001	42.		1	0	1
		.0001	42.		1	1	0
	3	.0000	0.		*		
-----				-----			
38 NSURGE				12			
-----				-----			

0
1
2
3

13

012

14

012

16

0
1
2

17 17 18

0	0	0
0	0	1
0	1	0
1	0	0
0	1	1
1	0	1
1	1	0
*		

15 38 39 40 41 42

0	-2	-2	-2	-2	-3
0	2	-2	-2	-2	-3
*					

20

0
1
2
3
4
5

21

● ● ● ● ● ●

19 44 45

0	99.5750	100.	0	-2	-2
1	.1023	75.	0	2	-3
	.0036	75.	0	-2	2
2	.0011	50.	0	3	-4
	.0000	50.	0	-3	3
3	.0000	25.	0	4	-5
	.0000	25.	0	-4	4
4	.3180	0.	*		

47 NCTHRSG

22

0	90.0693	100.
1	9.6711	50.
2	.2596	0.

0
1
2

48 N3BFP50

24

0	98.3973	150.
1	1.5940	100.
2	.0086	50.
3	.0000	0.

0
1
2
3

49 NPLANT

1 33 37 35 34 36 43 46

0	72.7800	100.
1	2.9815	84.
	.2583	84.
	7.7993	84.
	2.7405	84.
2	.0920	75.
3	.4715	50.
	.0001	50.
	.0010	50.
	9.3049	50.
	.0075	50.
4	.0396	42.
	.0003	42.
	.3109	42.
	.0439	42.
5	.0000	25.
6	3.1686	0.

0	0	0	0	0	0	0	0
0	1	-2	-2	0	-2	0	0
0	0	1	-2	0	-2	0	0
0	0	0	1	0	-2	0	0
0	0	0	0	0	1	0	0
0	-2	-2	-2	0	-2	0	1
0	-2	-2	-2	1	-2	-2	-3
0	-2	-2	-2	0	-2	1	-3
0	-2	-2	-2	0	-2	0	2
0	-2	-2	-2	0	-2	0	-2
0	-2	-2	-2	0	-2	0	-2
0	2	-3	-3	-2	-3	-2	-3
0	-2	2	-3	-2	-3	-2	-3
0	-2	-2	2	-2	-3	-2	-3
0	-2	-2	-2	-2	2	-2	-3
0	-3	-3	-3	-2	-3	-2	3
*							

23 47 48

0	72.7800	100.
1	2.9815	84.
	.2583	84.
	7.7993	84.
	2.7405	84.
2	.0920	75.
3	.4715	50.
	.0001	50.
	.0010	50.
	9.3049	50.
	.0075	50.
4	.0396	42.
	.0003	42.
	.3109	42.
	.0439	42.
5	.0000	25.

0	0	-2
0	0	-2
0	0	-2
0	0	-2
0	0	-2
0	0	-2
0	-2	-3
0	-2	-3
0	-2	-3
0	-2	-3
0	1	-3
0	0	2
0	-2	-3
0	-2	-3
0	-2	-3
0	-2	-3
0	-2	-3
0	-2	-3

NOTE: "*" INDICATES THAT NO SINGLE DEFINITION CAN COMPLETELY DESCRIBE A GIVEN SUBSTATE.
SEE USER'S GUIDE FOR DETAILED EXPLANATION.

PLANT STATE	OUTPUT STATE PROBABILITY	OUTPUT CAPABILITY	DAYS/ YEAR	POWER (MW) OUTPUT
1	72.78%	100.00%	253.15	400.00
2	13.78%	84.00%	47.93	336.00
3	.09%	75.00%	.32	300.00
4	9.78%	50.00%	34.04	200.00
5	.39%	42.00%	1.37	168.00
6	.00%	25.00%	.00	100.00
7	3.17%	.00%	11.02	.00

PLANT RAM MEASURES				
EQUIVALENT				
FORCED OUTAGE RATE	FORCED OUTAGE RATE	EQUIVALENT AVAILABILITY FACTOR	AVAILABILITY FACTOR	
3.17%	10.52%	85.27%	92.28%	

UNIT-LEVEL R&M VALUES			
UNIT MEAN TIME BETWEEN FULL FORCED OUTAGES (HOURS)	MEAN UNIT FULL FORCED OUTAGE DURATION (HOURS)		
338.9	11.1		

WANSLEY IGCC STUDY II FOR D.O.E. (CASE 1)

 ***** RUN 1 *****

COMPONENT RANKING FOR CRITICALITY						
SUB. NO.	COMP. NO.	RANKING FACTOR	SUBSYSTEM	COMPONENT	HOUR EQUIV	OUTPUT PCT.
9	1	2.8707	CTHRSG50	CTG	239.6	27.0
9	2	1.5937	CTHRSG50	HRSG	133.0	15.0
10	4	1.2387	BOPSTG	STG	103.4	11.7
3	1	.5880	SULFATOR	LASHRECEIVER	49.1	5.5
3	5	.5525	SULFATOR	SULFATORINTERNALS	46.1	5.2
3	4	.5089	SULFATOR	SULFATOR	42.5	4.8
2	14	.2752	CATCHALL6	PRODUCTGASCOOLER	23.0	2.6
2	11	.2538	CATCHALL6	GASIFIER	21.2	2.4
3	9	.1537	SULFATOR	FLUGASBAGHOUSEFILTER	12.8	1.4
7	2	.1446	ZINCFERR	ZINCFERRITEREACTOR	12.1	1.4
7	7	.1446	ZINCFERR	ZINCFERRITEREACTOR2	12.1	1.4
3	11	.1294	SULFATOR	SULFSOLIDSCREWCOOLER	10.8	1.2
10	3	.1274	BOPSTG	CONDENSER	10.6	1.2
3	3	.1228	SULFATOR	SULFN.AIRCOMPRESSOR	10.2	1.2
2	12	.1216	CATCHALL6	CYCLONE	10.2	1.1
2	16	.1216	CATCHALL6	FINESCYCLONE	10.2	1.1
1	4	.1212	BOOST AC	BOOSTERAC	10.1	1.1
6	1	.1063	STARTHEATR	STARTUPHEATER	8.9	1.0
6	2	.0849	STARTHEATR	ST.UP.HEAT.INTERNAL	7.1	.8
3	2	.0830	SULFATOR	SCREWFEEDER	6.9	.8
3	7	.0825	SULFATOR	SULFATORHEATRECOVERY	6.9	.8
2	28	.0651	CATCHALL6	SOLIDSCONVEYOR	5.4	.6
2	4	.0618	CATCHALL6	PRESSZN.AC	5.2	.6
2	10	.0505	CATCHALL6	COALFEEDER	4.2	.5
2	13	.0505	CATCHALL6	ASHFEEDER	4.2	.5
2	23	.0488	CATCHALL6	GASFILTER	4.1	.5
3	8	.0472	SULFATOR	SULFATORSTEAMDRUM	3.9	.4
3	10	.0472	SULFATOR	FLGASOLIDSCREWCOOLER	3.9	.4
19	1	.0430	COAL100	RECLAIMCONVEYOR	3.6	.4
19	2	.0430	COAL100	TRANSFERCONVEYOR1	3.6	.4
19	3	.0430	COAL100	TRANSFERCONVEYOR2	3.6	.4
19	4	.0430	COAL100	TRANSFERCONVEYOR3	3.6	.4
19	5	.0430	COAL100	TRANSFERCONVEYOR4	3.6	.4
19	6	.0430	COAL100	CASCADECONVEYOR	3.6	.4
2	15	.0388	CATCHALL6	STEAMDRUM	3.2	.4
15	1	.0345	SIL0	ASHSIL0	2.9	.3
3	6	.0313	SULFATOR	SOLIDSDISENGAGER	2.6	.3
2	17	.0296	CATCHALL6	CYC.FINE.LOCKHOPPER	2.5	.3
2	20	.0296	CATCHALL6	CFDEPRESSLOCKHOPPER	2.5	.3
2	25	.0296	CATCHALL6	G.F.DEPRSSLOCKHOPPER	2.5	.3
1	2	.0271	BOOST AC	TRIMCOOLER	2.3	.3
1	5	.0271	BOOST AC	AIRRECUPERATOR	2.3	.3
2	7	.0265	CATCHALL6	PRSZN.FEEDHOPPER	2.2	.2
19	8	.0236	COAL100	HOLDINGBIN	2.0	.2
2	9	.0217	CATCHALL6	FEEDHOPPER	1.8	.2

2	26	.0186	CATCHALL6	GFLOCKHOPPERFILTER	1.6	.2
7	11	.0158	ZINCFERR	GUARDCYCLONE	1.3	.1
7	12	.0158	ZINCFERR	GUARDDRUM	1.3	.1
2	1	.0148	CATCHALL6	FEEDSURGEDRUM	1.2	.1
1	3	.0139	BOOST AC	KNOCKOUTDRUM	1.2	.1
7	5	.0139	ZINCFERR	VALVEREGENER3	1.2	.1
7	10	.0139	ZINCFERR	VALVEREGENER6	1.2	.1
7	13	.0138	ZINCFERR	EXITGASCOOLER	1.2	.1
2	5	.0138	CATCHALL6	PRSZN.AIRRECEIVER	1.2	.1
7	1	.0111	ZINCFERR	VALVEZFDS1	.9	.1
7	4	.0111	ZINCFERR	VALVEZFDS2	.9	.1
7	6	.0111	ZINCFERR	VALVEZFDS4	.9	.1
7	9	.0111	ZINCFERR	VALVEZFDS5	.9	.1
2	29	.0080	CATCHALL6	RECYCLEGASCOOLER	.7	.1
2	30	.0080	CATCHALL6	RECYCLEGASRECEIVER	.7	.1
20	1	.0073	MILLINGSYS	EXHAUSTFAN	.6	.1
20	2	.0073	MILLINGSYS	BLOWFAN	.6	.1
20	3	.0063	MILLINGSYS	BAGFILTER	.5	.1
11	1	.0034	BASICBFP	BFP	.3	.0
2	2	.0033	CATCHALL6	VALVECP1	.3	.0
2	18	.0033	CATCHALL6	VALVEFR1	.3	.0
2	6	.0032	CATCHALL6	COALFEEDVENTFILTER	.3	.0
2	21	.0032	CATCHALL6	CYCLOCKHOPPERFILTER	.3	.0
14	2	.0024	VACUUM	VACPUMP	.2	.0
20	7	.0014	MILLINGSYS	COALMILL	.1	.0
2	3	.0012	CATCHALL6	VALVECP2	.1	.0
2	8	.0012	CATCHALL6	VALVECP3	.1	.0
2	19	.0012	CATCHALL6	VALVEFR2	.1	.0
2	22	.0012	CATCHALL6	VALVEFR3	.1	.0
2	24	.0012	CATCHALL6	VALVEFR5	.1	.0
2	27	.0012	CATCHALL6	VALVEFR6	.1	.0
4	2	.0011	CHLORIDEGD	CHLORIDEGUARDDRUM	.1	.0
20	6	.0010	MILLINGSYS	COALFEEDER1	.1	.0
20	4	.0009	MILLINGSYS	CYCLONE	.1	.0
21	3	.0007	LASTFEED	PNEUMATICELEVATOR	.1	.0
13	1	.0007	BAGHOUSE	BAGHOUSES	.1	.0
5	1	.0005	RGCOMPRESS	RECYCLEGASCOMPRESSOR	.0	.0
14	1	.0003	VACUUM	FILTER	.0	.0
20	5	.0003	MILLINGSYS	COALSILO	.0	.0
16	1	.0002	AERATION	BLOWER	.0	.0
21	1	.0002	LASTFEED	LIMESTONEFEEDER	.0	.0
21	2	.0002	LASTFEED	COALFEEDER2	.0	.0
19	7	.0002	COAL100	PNEULIMECONVEYORS	.0	.0
4	1	.0001	CHLORIDEGD	VALVECG1	.0	.0
4	3	.0001	CHLORIDEGD	VALVECG2	.0	.0
10	1	.0001	BOPSTG	CONDPUMP	.0	.0
1	1	.0001	BOOST AC	BFWPREHEATER	.0	.0
12	1	.0000	DISCHARGE	SURGEBIN	.0	.0
7	3	.0000	ZINCFERR	ZNFERRITESORBENT	.0	.0
7	8	.0000	ZINCFERR	ZNFERRITESORBENT2	.0	.0
12	2	.0000	DISCHARGE	OUTCYLVALVE	.0	.0
10	2	.0000	BOPSTG	CIRCWPUMP	.0	.0
17	2	.0000	CONDITION	ASHCONDITIONER	.0	.0
18	2	.0000	DRYUNLOAD	UNLOADERBLOWER	.0	.0
8	1	.0000	DUMMY	DUMMYCOMP	.0	.0
7	1	.0000	CONDITION	CVALVE	.0	.0
18	1	.0000	DRYUNLOAD	DVALVE	.0	.0

 ***** RUN 1 *****

COMPONENT RANKING FOR CRITICALITY						
SUB. NO.	COMP. NO.	RANKING FACTOR	SUBSYSTEM	COMPONENT	HOUR EQUIV	OUTPUT PCT.

9	1	3.0133	CTHRSG50	CTG	251.6	27.3
9	2	1.6715	CTHRSG50	HRSG	139.5	15.1
3	1	.6429	SULFATOR	LASHRECEIVER	53.7	5.8
3	5	.6041	SULFATOR	SULFATORINTERNALS	50.4	5.5
2	11	.5564	CATCHALL6	GASIFIER	46.4	5.0
3	4	.5564	SULFATOR	SULFATOR	46.4	5.0
7	2	.3154	ZINCERR	ZINCERRITEREACTOR	26.3	2.9
7	7	.3154	ZINCERR	ZINCERRITEREACTOR2	26.3	2.9
2	12	.2649	CATCHALL6	CYCLONE	22.1	2.4
2	14	.2649	CATCHALL6	FINESCYCLONE	22.1	2.4
3	9	.1680	SULFATOR	FLUGASBAGHOUSEFILTER	14.0	1.5
2	26	.1414	CATCHALL6	SOLIDSCONVEYOR	11.8	1.3
3	11	.1414	SULFATOR	SULFSOLIDSCREWCOOLER	11.8	1.3
2	4	.1342	CATCHALL6	PRESSZN.AC	11.2	1.2
1	4	.1342	BOOST AC	BOOSTERAC	11.2	1.2
3	3	.1342	SULFATOR	SULFN.AIRCOMPRESSOR	11.2	1.2
2	10	.1096	CATCHALL6	COALFEEDER	9.2	1.0
2	13	.1096	CATCHALL6	ASHFEEDER	9.2	1.0
6	1	.1061	STARTHEATR	STARTUPHEATER	8.9	1.0
2	21	.1059	CATCHALL6	GASFILTER	8.8	1.0
3	2	.0907	SULFATOR	SCREWFEEDER	7.6	.8
3	7	.0901	SULFATOR	SULFATORHEATRECOVERY	7.5	.8
6	2	.0847	STARTHEATR	ST.UP.HEAT. INTERNALS	7.1	.8
2	15	.0642	CATCHALL6	CYC.FINE.LOCKHOPPER	5.4	.6
2	18	.0642	CATCHALL6	CFDEPRESSLOCKHOPPER	5.4	.6
2	23	.0642	CATCHALL6	G.F.DEPRSSLOCKHOPPER	5.4	.6
2	7	.0575	CATCHALL6	PRSZN.FEEDHOPPER	4.8	.5
3	8	.0516	SULFATOR	SULFATORSTEAMDRUM	4.3	.5
3	10	.0516	SULFATOR	FLGASOLIDSCREWCOOLER	4.3	.5
2	9	.0470	CATCHALL6	FEEDHOPPER	3.9	.4
18	1	.0428	COAL100	RECLAIMCONVEYOR	3.6	.4
18	2	.0428	COAL100	TRANSFERCONVEYOR1	3.6	.4
18	3	.0428	COAL100	TRANSFERCONVEYOR2	3.6	.4
18	4	.0428	COAL100	TRANSFERCONVEYOR3	3.6	.4
18	5	.0428	COAL100	TRANSFERCONVEYOR4	3.6	.4
18	6	.0428	COAL100	CASCADECONVEYOR	3.6	.4
2	24	.0403	CATCHALL6	GFLOCKHOPPERFILTER	3.4	.4
14	1	.0344	SILO	ASHSILO	2.9	.3
3	6	.0342	SULFATOR	SOLIDSDISENGAGER	2.9	.3
7	11	.0342	ZINCERR	GUARDCYCLONE	2.9	.3
7	12	.0342	ZINCERR	GUARDDRUM	2.9	.3
2	1	.0321	CATCHALL6	FEEDSURGEDRUM	2.7	.3
7	5	.0302	ZINCERR	VALVEREGENER3	2.5	.3
7	10	.0302	ZINCERR	VALVEREGENER6	2.5	.3
1	2	.0300	BOOST_AC	TRIMCOOLER	2.5	.3

1	5	.0300	BOOST AC	AIRRECUPERATOR	2.5	.3
7	13	.0300	ZINCFERR	EXITGASCOOLER	2.5	.3
2	5	.0300	CATCHALL6	PRSZN.AIRRECEIVER	2.5	.3
7	1	.0241	ZINCFERR	VALVEZFDS1	2.0	.2
7	4	.0241	ZINCFERR	VALVEZFDS2	2.0	.2
7	6	.0241	ZINCFERR	VALVEZFDS4	2.0	.2
7	9	.0241	ZINCFERR	VALVEZFDS5	2.0	.2
18	8	.0235	COAL100	HOLDINGBIN	2.0	.2
2	27	.0174	CATCHALL6	RECYCLEGASCOOLER	1.5	.2
2	28	.0174	CATCHALL6	RECYCLEGASRECEIVER	1.5	.2
1	3	.0154	BOOST AC	KNOCKOUTDRUM	1.3	.1
2	2	.0072	CATCHALL6	VALVECP1	.6	.1
2	16	.0072	CATCHALL6	VALVEFR1	.6	.1
2	6	.0070	CATCHALL6	COALFEEDVENTFILTER	.6	.1
2	19	.0070	CATCHALL6	CYCLOCKHOPPERFILTER	.6	.1
19	4	.0043	MILLINGSYS	EXHAUSTFAN	.4	.0
19	5	.0043	MILLINGSYS	BLOWFAN	.4	.0
10	1	.0038	BASICBFP	BFP	.3	.0
19	6	.0037	MILLINGSYS	BAGFILTER	.3	.0
2	3	.0026	CATCHALL6	VALVECP2	.2	.0
2	8	.0026	CATCHALL6	VALVECP3	.2	.0
2	17	.0026	CATCHALL6	VALVEFR2	.2	.0
2	20	.0026	CATCHALL6	VALVEFR3	.2	.0
2	22	.0026	CATCHALL6	VALVEFR5	.2	.0
2	25	.0026	CATCHALL6	VALVEFR6	.2	.0
13	2	.0024	VACUUM	VACPUMP	.2	.0
4	2	.0023	CHLORIDEGD	CHLORIDEGUARDDRUM	.2	.0
5	1	.0008	RGCOMPRESS	RECYCLEGASCOMPRESSOR	.1	.0
19	3	.0008	MILLINGSYS	COALMILL	.1	.0
12	1	.0007	BAGHOUSE	BAGHOUSES	.1	.0
19	2	.0006	MILLINGSYS	COALFEEDER1	.0	.0
19	7	.0005	MILLINGSYS	CYCLONE	.0	.0
20	3	.0004	LASTFEED	PNEUMATICELEVATOR	.0	.0
4	1	.0003	CHLORIDEGD	VALVECG1	.0	.0
4	3	.0003	CHLORIDEGD	VALVECG2	.0	.0
13	1	.0003	VACUUM	FILTER	.0	.0
15	1	.0002	AERATION	BLOWER	.0	.0
18	7	.0002	COAL100	PNEULIMECONVEYORS	.0	.0
19	1	.0002	MILLINGSYS	COALSILO	.0	.0
20	1	.0001	LASTFEED	LIMESTONEFEEDER	.0	.0
20	2	.0001	LASTFEED	COALFEEDER2	.0	.0
7	3	.0001	ZINCFERR	ZNFERRITESORBENT	.0	.0
7	8	.0001	ZINCFERR	ZNFERRITESORBENT2	.0	.0
1	1	.0001	BOOST AC	BFWPREHEATER	.0	.0
11	1	.0000	DISCHARGE	SURGEBIN	.0	.0
11	2	.0000	DISCHARGE	OUTCYLVALVE	.0	.0
16	2	.0000	CONDITION	ASHCONDITIONER	.0	.0
17	2	.0000	DRYUNLOAD	UNLOADERBLOWER	.0	.0
8	1	.0000	DUMMY	DUMMYCOMP	.0	.0
16	1	.0000	CONDITION	CVALVE	.0	.0
17	1	.0000	DRYUNLOAD	DVALVE	.0	.0

 ***** RUN 1 *****

COMPONENT RANKING FOR CRITICALITY						
SUB. NO.	COMP. NO.	RANKING FACTOR	SUBSYSTEM	COMPONENT	HOUR EQUIV	OUTPUT PCT.

9	1	2.9704	CTHRSG50	CTG	248.0	24.6
9	2	1.6477	CTHRSG50	HRSG	137.5	13.6
10	4	1.2178	BOPSTG	STG	101.7	10.1
3	1	.6130	SULFATOR	LASHRECEIVER	51.2	5.1
3	5	.5760	SULFATOR	SULFATORINTERNALS	48.1	4.8
3	4	.5305	SULFATOR	SULFATOR	44.3	4.4
2	14	.4739	CATCHALL6	PRODUCTGASCOOLER	39.6	3.9
2	11	.4366	CATCHALL6	GASIFIER	36.4	3.6
7	2	.2476	ZINCERR	ZINCERRITEREACTOR	20.7	2.1
7	7	.2476	ZINCERR	ZINCERRITEREACTOR2	20.7	2.1
2	12	.2080	CATCHALL6	CYCLONE	17.4	1.7
2	16	.2080	CATCHALL6	FINESCYCLONE	17.4	1.7
3	9	.1602	SULFATOR	FLUGASBAGHOUSEFILTER	13.4	1.3
3	11	.1348	SULFATOR	SULFSOLIDSCREWCOOLER	11.3	1.1
1	4	.1322	BOOST AC	BOOSTERAC	11.0	1.1
3	3	.1280	SULFATOR	SULFN.AIRCOMPRESSOR	10.7	1.1
10	3	.1253	BOPSTG	CONDENSER	10.5	1.0
2	28	.1111	CATCHALL6	SOLIDSCONVEYOR	9.3	.9
2	4	.1054	CATCHALL6	PRESSZN.AC	8.8	.9
6	1	.1045	STARTHEATR	STARTUPHEATER	8.7	.9
3	2	.0864	SULFATOR	SCREWFEEDER	7.2	.7
2	10	.0861	CATCHALL6	COALFEEDER	7.2	.7
2	13	.0861	CATCHALL6	ASHFEEDER	7.2	.7
3	7	.0860	SULFATOR	SULFATORHEATRECOVERY	7.2	.7
6	2	.0835	STARTHEATR	ST.UP.HEAT.INTERNAL	7.0	.7
2	23	.0832	CATCHALL6	GASFILTER	6.9	.7
2	15	.0662	CATCHALL6	STEAMDUM	5.5	.5
2	17	.0505	CATCHALL6	CYC.FINE.LOCKHOPPER	4.2	.4
2	20	.0505	CATCHALL6	CFDEPRESSLOCKHOPPER	4.2	.4
2	25	.0505	CATCHALL6	G.F.DEPRSSLOCKHOPPER	4.2	.4
3	8	.0492	SULFATOR	SULFATORSTEAMDUM	4.1	.4
3	10	.0492	SULFATOR	FLGASOLIDSCREWCOOLER	4.1	.4
2	7	.0452	CATCHALL6	PRSZN.FEEDHOPPER	3.8	.4
19	1	.0422	COAL100	RECLAIMCONVEYOR	3.5	.3
19	2	.0422	COAL100	TRANSFERCONVEYOR1	3.5	.3
19	3	.0422	COAL100	TRANSFERCONVEYOR2	3.5	.3
19	4	.0422	COAL100	TRANSFERCONVEYOR3	3.5	.3
19	5	.0422	COAL100	TRANSFERCONVEYOR4	3.5	.3
19	6	.0422	COAL100	CASCADECONVEYOR	3.5	.3
2	9	.0369	CATCHALL6	FEEDHOPPER	3.1	.3
15	1	.0339	SIL0	ASHSIL0	2.8	.3
3	6	.0326	SULFATOR	SOLIDSDISENGAGER	2.7	.3
2	26	.0316	CATCHALL6	GFLOCKHOPPERFILTER	2.6	.3
1	2	.0295	BOOST AC	TRIMCOOLER	2.5	.2
1	5	.0295	BOOST AC	AIRRECUPERATOR	2.5	.2

7	11	.0269	ZINCFERR	GUARDCYCLONE	2.2	.2
7	12	.0269	ZINCFERR	GUARDDRUM	2.2	.2
2	1	.0252	CATCHALL6	FEEDSURGEDRUM	2.1	.2
7	5	.0237	ZINCFERR	VALVEREGENER3	2.0	.2
7	10	.0237	ZINCFERR	VALVEREGENER6	2.0	.2
7	13	.0236	ZINCFERR	EXITGASCOOLER	2.0	.2
2	5	.0235	CATCHALL6	PRSZN.AIRRECEIVER	2.0	.2
19	8	.0232	COAL100	HOLDINGBIN	1.9	.2
7	1	.0189	ZINCFERR	VALVEZFDS1	1.6	.2
7	4	.0189	ZINCFERR	VALVEZFDS2	1.6	.2
7	6	.0189	ZINCFERR	VALVEZFDS4	1.6	.2
7	9	.0189	ZINCFERR	VALVEZFDS5	1.6	.2
1	3	.0152	BOOST AC	KNOCKOUTDRUM	1.3	.1
2	29	.0136	CATCHALL6	RECYCLEGASCOOLER	1.1	.1
2	30	.0136	CATCHALL6	RECYCLEGASRECEIVER	1.1	.1
20	4	.0059	MILLINGSYS	BLOWFAN	.5	.0
20	6	.0059	MILLINGSYS	EXHAUSTFAN	.5	.0
2	2	.0056	CATCHALL6	VALVECP1	.5	.0
2	18	.0056	CATCHALL6	VALVEFR1	.5	.0
2	6	.0055	CATCHALL6	COALFEEDVENTFILTER	.5	.0
2	21	.0055	CATCHALL6	CYCLOCKHOPPERFILTER	.5	.0
20	5	.0051	MILLINGSYS	BAGFILTER	.4	.0
11	1	.0035	BASICBFP	BFP	.3	.0
14	2	.0024	VACUUM	VACPUMP	.2	.0
2	3	.0020	CATCHALL6	VALVECP2	.2	.0
2	8	.0020	CATCHALL6	VALVECP3	.2	.0
2	19	.0020	CATCHALL6	VALVEFR2	.2	.0
2	22	.0020	CATCHALL6	VALVEFR3	.2	.0
2	24	.0020	CATCHALL6	VALVEFR5	.2	.0
2	27	.0020	CATCHALL6	VALVEFR6	.2	.0
4	2	.0018	CHLORIDEGD	CHLORIDEGUARDDRUM	.2	.0
20	3	.0011	MILLINGSYS	COALMILL	.1	.0
5	1	.0009	RGCOMPRESS	RECYCLEGASCOMPRESSOR	.1	.0
20	2	.0008	MILLINGSYS	COALFEEDER1	.1	.0
20	7	.0007	MILLINGSYS	CYCLONE	.1	.0
13	1	.0006	BAGHOUSE	BAGHOUSES	.1	.0
21	3	.0005	LASTFEED	PNEUMATICELEVATOR	.0	.0
14	1	.0003	VACUUM	FILTER	.0	.0
4	1	.0002	CHLORIDEGD	VALVECG1	.0	.0
4	3	.0002	CHLORIDEGD	VALVECG2	.0	.0
20	1	.0002	MILLINGSYS	COALSILO	.0	.0
16	1	.0002	AERATION	BLOWER	.0	.0
19	7	.0002	COAL100	PNEULIMECONVEYORS	.0	.0
21	1	.0002	LASTFEED	LIMESTONEFEEDER	.0	.0
21	2	.0002	LASTFEED	COALFEEDER2	.0	.0
1	1	.0001	BOOST AC	BFWPREHEATER	.0	.0
7	3	.0001	ZINCFERR	ZNFERRITESORBENT	.0	.0
7	8	.0001	ZINCFERR	ZNFERRITESORBENT2	.0	.0
10	1	.0001	BOPSTG	CONDUMP	.0	.0
12	1	.0000	DISCHARGE	SURGEIN	.0	.0
10	2	.0000	BOPSTG	CIRCWPUMP	.0	.0
17	2	.0000	CONDITION	ASHCONDITIONER	.0	.0
18	2	.0000	DRYUNLOAD	UNLOADERBLOWER	.0	.0
8	1	.0000	DUMMY	DUMMYCOMP	.0	.0
12	2	.0000	DISCHARGE	OUTCYLVALVE	.0	.0
7	1	.0000	CONDITION	CVALVE	.0	.0
16	1	.0000	DRYUNLOAD	DVALVE	.0	.0

 ***** RUN 1 *****

COMPONENT RANKING FOR CRITICALITY						
SUB. NO.	COMP. NO.	RANKING FACTOR	SUBSYSTEM	COMPONENT	HOUR EQUIV	OUTPUT PCT.
9	1	2.8580	CTHRSG50	CTG	238.6	31.9
9	2	1.5874	CTHRSG50	HRSG	132.5	17.7
3	1	.6122	SULFATOR	LASHRECEIVER	51.1	6.8
3	5	.5752	SULFATOR	SULFATORINTERNALS	48.0	6.4
3	4	.5298	SULFATOR	SULFATOR	44.2	5.9
2	11	.2378	CATCHALL6	GASIFIER	19.9	2.7
3	9	.1600	SULFATOR	FLUGASBAGHOUSEFILTER	13.4	1.8
7	2	.1356	ZINCFERR	ZINCFERRITEREACTOR	11.3	1.5
7	7	.1356	ZINCFERR	ZINCFERRITEREACTOR2	11.3	1.5
3	11	.1347	SULFATOR	SULFSOLIDSCREWCOOLER	11.2	1.5
3	3	.1278	SULFATOR	SULFN.AIRCOMPRESSOR	10.7	1.4
1	4	.1240	BOOST_AC	BOOSTERAC	10.3	1.4
2	12	.1141	CATCHALL6	CYCLONE	9.5	1.3
2	14	.1141	CATCHALL6	FINESCYCLONE	9.5	1.3
6	1	.1085	STARTHEATR	STARTUPHEATER	9.1	1.2
6	2	.0866	STARTHEATR	ST.UP.HEAT.INTERNAL	7.2	1.0
3	2	.0864	SULFATOR	SCREWFEEDER	7.2	1.0
3	7	.0859	SULFATOR	SULFATORHEATRECOVERY	7.2	1.0
2	26	.0611	CATCHALL6	SOLIDSCONVEYOR	5.1	.7
2	4	.0580	CATCHALL6	PRESSZN.AC	4.8	.6
3	8	.0491	SULFATOR	SULFATORSTEAMDUM	4.1	.5
3	10	.0491	SULFATOR	FLGASOLIDSCREWCOOLER	4.1	.5
2	10	.0474	CATCHALL6	COALFEEDER	4.0	.5
2	13	.0474	CATCHALL6	ASHFEEDER	4.0	.5
2	21	.0458	CATCHALL6	GASFILTER	3.8	.5
18	1	.0438	COAL100	RECLAIMCONVEYOR	3.7	.5
18	2	.0438	COAL100	TRANSFERCONVEYOR1	3.7	.5
18	3	.0438	COAL100	TRANSFERCONVEYOR2	3.7	.5
18	4	.0438	COAL100	TRANSFERCONVEYOR3	3.7	.5
18	5	.0438	COAL100	TRANSFERCONVEYOR4	3.7	.5
18	6	.0438	COAL100	CASCADECONVEYOR	3.7	.5
14	1	.0352	SILO	ASHSILO	2.9	.4
3	6	.0326	SULFATOR	SOLIDSDISENGAGER	2.7	.4
2	15	.0278	CATCHALL6	CYC.FINE.LOCKHOPPER	2.3	.3
2	18	.0278	CATCHALL6	CFDEPRESSLOCKHOPPER	2.3	.3
2	23	.0278	CATCHALL6	G.F.DEPRSSLOCKHOPPER	2.3	.3
1	2	.0277	BOOST_AC	TRIMCOOLER	2.3	.3
1	5	.0277	BOOST_AC	AIRRECUPERATOR	2.3	.3
2	7	.0249	CATCHALL6	PRSZN.FEEDHOPPER	2.1	.3
18	8	.0241	COAL100	HOLDINGBIN	2.0	.3
2	9	.0204	CATCHALL6	FEEDHOPPER	1.7	.2
2	24	.0174	CATCHALL6	GFLOCKHOPPERFILTER	1.5	.2
7	11	.0148	ZINCFERR	GUARDCYCLONE	1.2	.2
7	12	.0148	ZINCFERR	GUARDDRUM	1.2	.2
1	3	.0142	BOOST AC	KNOCKOUTDRUM D-92	1.2	.2

2	1	.0139	CATCHALL6	FEEDSURGEDRUM	1.2	.2
7	5	.0131	ZINCERR	VALVEREGENER3	1.1	.1
7	10	.0131	ZINCERR	VALVEREGENER6	1.1	.1
7	13	.0130	ZINCERR	EXITGASCOOLER	1.1	.1
2	5	.0130	CATCHALL6	PRSZN.AIRRECEIVER	1.1	.1
7	1	.0104	ZINCERR	VALVEZFDS1	.9	.1
7	4	.0104	ZINCERR	VALVEZFDS2	.9	.1
7	6	.0104	ZINCERR	VALVEZFDS4	.9	.1
7	9	.0104	ZINCERR	VALVEZFDS5	.9	.1
2	27	.0075	CATCHALL6	RECYCLEGASCOOLER	.6	.1
2	28	.0075	CATCHALL6	RECYCLEGASRECEIVER	.6	.1
19	4	.0073	MILLINGSYS	BLOWFAN	.6	.1
19	6	.0073	MILLINGSYS	EXHAUSTFAN	.6	.1
19	5	.0064	MILLINGSYS	BAGFILTER	.5	.1
10	1	.0035	BASICBFP	BFP	.3	.0
2	2	.0031	CATCHALL6	VALVECP1	.3	.0
2	16	.0031	CATCHALL6	VALVEFR1	.3	.0
2	6	.0030	CATCHALL6	COALFEEDVENTFILTER	.3	.0
2	19	.0030	CATCHALL6	CYCLOCKHOPPERFILTER	.3	.0
13	2	.0025	VACUUM	VACPUMP	.2	.0
19	3	.0014	MILLINGSYS	COALMILL	.1	.0
2	3	.0011	CATCHALL6	VALVECP2	.1	.0
2	8	.0011	CATCHALL6	VALVECP3	.1	.0
2	17	.0011	CATCHALL6	VALVEFR2	.1	.0
2	20	.0011	CATCHALL6	VALVEFR3	.1	.0
2	22	.0011	CATCHALL6	VALVEFR5	.1	.0
2	25	.0011	CATCHALL6	VALVEFR6	.1	.0
4	2	.0010	CHLORIDEGD	CHLORIDEGUARDDRUM	.1	.0
19	2	.0010	MILLINGSYS	COALFEEDER1	.1	.0
19	7	.0009	MILLINGSYS	CYCLONE	.1	.0
20	3	.0007	LASTFEED	PNEUMATICELEVATOR	.1	.0
12	1	.0007	BAGHOUSE	BAGHOUSES	.1	.0
5	1	.0005	RGCOMPRESS	RECYCLEGASCOMPRESSOR	.0	.0
19	1	.0003	MILLINGSYS	COALSILO	.0	.0
13	1	.0003	VACUUM	FILTER	.0	.0
15	1	.0002	AERATION	BLOWER	.0	.0
20	1	.0002	LASTFEED	LIMESTONEFEEDER	.0	.0
20	2	.0002	LASTFEED	COALFEEDER2	.0	.0
18	7	.0002	COAL100	PNEULIMECONVEYORS	.0	.0
4	1	.0001	CHLORIDEGD	VALVECG1	.0	.0
4	3	.0001	CHLORIDEGD	VALVECG2	.0	.0
1	1	.0001	BOOST AC	BFWPREHEATER	.0	.0
7	3	.0000	ZINCERR	ZNFERRITESORBENT	.0	.0
7	8	.0000	ZINCERR	ZNFERRITESORBENT2	.0	.0
11	1	.0000	DISCHARGE	SURGBIN	.0	.0
11	2	.0000	DISCHARGE	OUTCYLVALVE	.0	.0
16	2	.0000	CONDITION	ASHCONDITIONER	.0	.0
17	2	.0000	DRYUNLOAD	UNLOADERBLOWER	.0	.0
8	1	.0000	DUMMY	RECLAIM	.0	.0
16	1	.0000	CONDITION	CVALVE	.0	.0
17	1	.0000	DRYUNLOAD	DVALVE	.0	.0

 ***** RUN 1 *****

COMPONENT RANKING FOR CRITICALITY						
SUB. NO.	COMP. NO.	RANKING FACTOR	SUBSYSTEM	COMPONENT	HOUR EQUIV	OUTPUT PCT.

8	1	2.8391	CTHRSG50	CTG	237.0	25.8
2	14	1.6798	CATCHALL6	PRODUCTGASCOOLER	140.2	15.2
8	2	1.5762	CTHRSG50	HRSG	131.6	14.3
9	4	1.2323	BOPSTG	STG	102.9	11.2
2	11	.4829	CATCHALL6	GASIFIER	40.3	4.4
6	2	.2739	ZINCFERR	ZINCFERRITEREACTOR	22.9	2.5
6	7	.2739	ZINCFERR	ZINCFERRITEREACTOR2	22.9	2.5
2	12	.2301	CATCHALL6	CYCLONE	19.2	2.1
2	16	.2301	CATCHALL6	FINESCYCLONE	19.2	2.1
9	3	.1268	BOPSTG	CONDENSER	10.6	1.2
2	28	.1229	CATCHALL6	SOLIDSCONVEYOR	10.3	1.1
1	4	.1198	BOOST_AC	BOOSTERAC	10.0	1.1
2	4	.1166	CATCHALL6	PRESSZN.AC	9.7	1.1
5	1	.1058	STARTHEATR	STARTUPHEATER	8.8	1.0
2	10	.0953	CATCHALL6	COALFEEDER	8.0	.9
2	13	.0953	CATCHALL6	ASHFEEDER	8.0	.9
2	23	.0920	CATCHALL6	GASFILTER	7.7	.8
5	2	.0845	STARTHEATR	ST.UP.HEAT. INTERNALS	7.1	.8
12	1	.0837	O2PLANT	OXYGENPLANT	7.0	.8
2	15	.0732	CATCHALL6	STEAMDRUM	6.1	.7
2	17	.0558	CATCHALL6	CYC.FINE.LOCKHOPPER	4.7	.5
2	20	.0558	CATCHALL6	CFDEPRESSLOCKHOPPER	4.7	.5
2	25	.0558	CATCHALL6	G.F.DEPRSSLOCKHOPPER	4.7	.5
2	7	.0500	CATCHALL6	PRSZN.FEEDHOPPER	4.2	.5
20	1	.0428	COAL100	RECLAIMCONVEYOR	3.6	.4
20	2	.0428	COAL100	TRANSFERCONVEYOR1	3.6	.4
20	3	.0428	COAL100	TRANSFERCONVEYOR2	3.6	.4
20	4	.0428	COAL100	TRANSFERCONVEYOR3	3.6	.4
20	5	.0428	COAL100	TRANSFERCONVEYOR4	3.6	.4
20	6	.0428	COAL100	CASCADECONVEYOR	3.6	.4
2	9	.0408	CATCHALL6	FEEDHOPPER	3.4	.4
2	26	.0350	CATCHALL6	GFLOCKHOPPERFILTER	2.9	.3
16	1	.0343	SIL0	ASHSIL0	2.9	.3
6	11	.0297	ZINCFERR	GUARDCYCLONE	2.5	.3
6	12	.0297	ZINCFERR	GUARDDRUM	2.5	.3
2	1	.0279	CATCHALL6	FEEDSURGEDRUM	2.3	.3
1	2	.0268	BOOST_AC	TRIMCOOLER	2.2	.2
1	5	.0268	BOOST_AC	AIRRECUPERATOR	2.2	.2
6	5	.0262	ZINCFERR	VALVEREGENER3	2.2	.2
6	10	.0262	ZINCFERR	VALVEREGENER6	2.2	.2
6	13	.0260	ZINCFERR	EXITGASCOOLER	2.2	.2
2	5	.0260	CATCHALL6	PRSZN.AIRRECEIVER	2.2	.2
13	1	.0245	DISCHARGE	SURGEBIN	2.0	.2
20	7	.0235	COAL100	HOLDINGBIN	2.0	.2
5	1	.0210	ZINCFERR	VALVEZFDS1	1.7	.2

6	4	.0210	ZINCFERR	VALVEZFDS2	1.7	.2
6	6	.0210	ZINCFERR	VALVEZFDS4	1.7	.2
6	9	.0210	ZINCFERR	VALVEZFDS5	1.7	.2
2	29	.0151	CATCHALL6	RECYCLEGASCOOLER	1.3	.1
2	30	.0151	CATCHALL6	RECYCLEGASRECEIVER	1.3	.1
1	3	.0138	BOOST AC	KNOCKOUTDRUM	1.1	.1
13	2	.0075	DISCHARGE	OUTCYLVALVE	.6	.1
21	4	.0071	MILLINGSYS	BLOWFAN	.6	.1
21	6	.0071	MILLINGSYS	EXHAUSTFAN	.6	.1
2	2	.0062	CATCHALL6	VALVECP1	.5	.1
2	18	.0062	CATCHALL6	VALVEFR1	.5	.1
21	5	.0062	MILLINGSYS	BAGFILTER	.5	.1
2	6	.0061	CATCHALL6	COALFEEDVENTFILTER	.5	.1
2	21	.0061	CATCHALL6	CYCLOCKHOPPERFILTER	.5	.1
10	1	.0034	BASICBFP	BFP	.3	.0
15	2	.0024	VACUUM	VACPUMP	.2	.0
2	3	.0022	CATCHALL6	VALVECP2	.2	.0
2	8	.0022	CATCHALL6	VALVECP3	.2	.0
2	19	.0022	CATCHALL6	VALVEFR2	.2	.0
2	22	.0022	CATCHALL6	VALVEFR3	.2	.0
2	24	.0022	CATCHALL6	VALVEFR5	.2	.0
2	27	.0022	CATCHALL6	VALVEFR6	.2	.0
3	2	.0020	CHLORIDEGD	CHLORIDEGUARDDRUM	.2	.0
21	3	.0013	MILLINGSYS	COALMILL	.1	.0
4	1	.0013	RGCOMPRESS	RECYCLEGASCOMPRESSOR	.1	.0
21	2	.0009	MILLINGSYS	COALFEEDER1	.1	.0
21	7	.0009	MILLINGSYS	CYCLONE	.1	.0
14	1	.0007	BAGHOUSE	BAGHOUSES	.1	.0
22	2	.0005	LASTFEED	PNEUMATICELEVATOR	.0	.0
.5	1	.0003	VACUUM	FILTER	.0	.0
21	1	.0003	MILLINGSYS	COALSILO	.0	.0
3	1	.0003	CHLORIDEGD	VALVECG1	.0	.0
3	3	.0003	CHLORIDEGD	VALVECG2	.0	.0
17	1	.0002	AERATION	BLOWER	.0	.0
22	1	.0002	LASTFEED	COALFEEDER2	.0	.0
11	1	.0001	SACIDPLANT	SULFURICACIDPLANT	.0	.0
9	1	.0001	BOPSTG	CONDUMP	.0	.0
1	1	.0001	BOOST AC	BFWPREHEATER	.0	.0
6	3	.0001	ZINCFERR	ZNFERRITESORBENT	.0	.0
6	8	.0001	ZINCFERR	ZNFERRITESORBENT2	.0	.0
9	2	.0000	BOPSTG	CIRCWPUMP	.0	.0
18	2	.0000	CONDITION	ASHCONDITIONER	.0	.0
19	2	.0000	DRYUNLOAD	UNLOADERBLOWER	.0	.0
7	1	.0000	DUMMY	DUMMYCOMP	.0	.0
18	1	.0000	CONDITION	CVALVE	.0	.0
19	1	.0000	DRYUNLOAD	DVALVE	.0	.0

 ***** RUN 1 *****

COMPONENT RANKING FOR CRITICALITY						
SUB. NO.	COMP. NO.	RANKING FACTOR	SUBSYSTEM	COMPONENT	HOUR EQUIV	OUTPUT PCT.

22	1	2.9612	CTHRSG50	CTG	247.2	30.0
22	2	1.6433	CTHRSG50	HRSG	137.2	16.7
23	4	1.2515	BOPSTG	STG	104.5	12.7
6	3	.4228	SULFATOR	FBHEATXCHANGE	35.3	4.3
6	7	.2600	SULFATOR	TRANSPORTSULFATOR	21.7	2.6
5	1	.2541	RECYCLEGAS	PRODUCTGASCOOLER	21.2	2.6
6	4	.2342	SULFATOR	HPASHCOOLER	19.6	2.4
6	6	.2342	SULFATOR	SULFATORSTEAMDRUM	19.6	2.4
3	1	.2334	TRANSPGAS	TRANSPORTGASIFIER	19.5	2.4
3	2	.1967	TRANSPGAS	STANDPIPE	16.4	2.0
2	1	.1316	BOOSTERAC	BOOSTERAIRCOMP	11.0	1.3
1	11	.1291	COAL&LIME	LOCKHOPPERAC	10.8	1.3
23	3	.1288	BOPSTG	CONDENSER	10.7	1.3
5	6	.1134	RECYCLEGAS	RECYCLEGASCOMPR	9.5	1.1
1	9	.1111	COAL&LIME	COALFEEDER	9.3	1.1
1	10	.1111	COAL&LIME	LIMEFEEDER	9.3	1.1
1	13	.1074	COAL&LIME	STARTUPHEATER	9.0	1.1
6	1	.1046	SULFATOR	FLUIDBED2NDCYCLONE	8.7	1.1
6	2	.1046	SULFATOR	FLUIDBED1STCYCLONE	8.7	1.1
1	14	.0858	COAL&LIME	SUHEATERINTERNAL	7.2	.9
1	2	.0789	COAL&LIME	LIMEPRESSHOPPER	6.6	.8
1	6	.0789	COAL&LIME	COALPRESSHOPPER	6.6	.8
1	4	.0523	COAL&LIME	LIMEFEEDHOPPER	4.4	.5
1	8	.0523	COAL&LIME	COALFEEDHOPPER	4.4	.5
1	1	.0486	COAL&LIME	LIMESURGEBIN	4.1	.5
1	5	.0486	COAL&LIME	CFSURGEBIN	4.1	.5
5	3	.0450	RECYCLEGAS	GASFILTER	3.8	.5
19	1	.0434	COAL100	RECLAIMCONVEYOR	3.6	.4
19	2	.0434	COAL100	TRANSFERCONVEYOR1	3.6	.4
19	3	.0434	COAL100	TRANSFERCONVEYOR2	3.6	.4
19	4	.0434	COAL100	TRANSFERCONVEYOR3	3.6	.4
19	5	.0434	COAL100	TRANSFERCONVEYOR4	3.6	.4
19	6	.0434	COAL100	CASCADECONVEYOR	3.6	.4
6	5	.0419	SULFATOR	ASHCOOLERCYCLONE	3.5	.4
5	2	.0359	RECYCLEGAS	STEAMDRUM	3.0	.4
15	1	.0349	SIL0	ASHSIL0	2.9	.4
1	12	.0303	COAL&LIME	AIRRECEIVER	2.5	.3
2	2	.0280	BOOSTERAC	AIRRECUPERATOR	2.3	.3
2	4	.0280	BOOSTERAC	TRIMCOOLER	2.3	.3
19	8	.0238	COAL100	HOLDINGBIN	2.0	.2
9	2	.0205	ASHFILTERS	ASHSCREWCONVEYOR	1.7	.2
2	5	.0144	BOOSTERAC	KNOCKOUTDRUM	1.2	.1
9	1	.0128	ASHFILTERS	ASHVENTFILTER	1.1	.1
20	5	.0100	MILLINGSYS	BLOWFAN	.8	.1
20	6	.0100	MILLINGSYS	EXHAUSTFAN	.8	.1

20	4	.0087	MILLINGSYS	BAGFILTER	.7	.1
5	5	.0074	RECYCLEGAS	RECYCLEGASCOOLER	.6	.1
5	7	.0074	RECYCLEGAS	RECYCLEGASRECEIVER	.6	.1
1	3	.0071	COAL&LIME	LIMEFVENTFILTER	.6	.1
1	7	.0071	COAL&LIME	CFVENTFILTER	.6	.1
4	2	.0063	TRNGASCYCL	PRIMARYCYCLONE	.5	.1
4	3	.0063	TRNGASCYCL	SECONDARYCYCLONE	.5	.1
7	1	.0057	SULFACYCL	SULFATORCYCLONE1	.5	.1
7	2	.0057	SULFACYCL	SULFATORCYCLONE2	.5	.1
20	3	.0041	MILLINGSYS	COALMILL	.3	.0
24	1	.0036	BASICBFP	BFP	.3	.0
5	4	.0031	RECYCLEGAS	GFILTERVALVE	.3	.0
9	4	.0028	ASHFILTERS	AIRRECEIVER	.2	.0
4	1	.0027	TRNGASCYCL	STANDPIPECYCLONE	.2	.0
14	2	.0024	VACUUM	VACPUMP	.2	.0
8	1	.0019	BAYONET	BAYONETBOILER	.2	.0
20	2	.0013	MILLINGSYS	COALFEEDER1	.1	.0
20	7	.0012	MILLINGSYS	CYCLONE	.1	.0
21	2	.0007	LASTFEED	PNEUMATICELEVATOR	.1	.0
13	1	.0007	BAGHOUSE	BAGHOUSES	.1	.0
20	1	.0004	MILLINGSYS	COALSILO	.0	.0
14	1	.0003	VACUUM	FILTER	.0	.0
21	1	.0003	LASTFEED	COALFEEDER2	.0	.0
16	1	.0002	AERATION	BLOWER	.0	.0
19	7	.0002	COAL100	PNEULIMECONVEYORS	.0	.0
10	1	.0002	BLOWBACK	BLOWBACKAC	.0	.0
2	3	.0001	BOOSTERAC	BFWPREHEATER	.0	.0
23	1	.0001	BOPSTG	CONDPUMP	.0	.0
9	3	.0000	ASHFILTERS	SULFBARRIERFILTER	.0	.0
12	1	.0000	DISCHARGE	SURGEBIN	.0	.0
11	1	.0000	ASHHOPPER	ASHDEPRESZNHOPPER	.0	.0
17	2	.0000	CONDITION	ASHCONDITIONER	.0	.0
18	2	.0000	DRYUNLOAD	UNLOADERBLOWER	.0	.0
23	2	.0000	BOPSTG	CIRCPUMP	.0	.0
12	2	.0000	DISCHARGE	OUTCYLVALVE	.0	.0
17	1	.0000	CONDITION	CVALVE	.0	.0
18	1	.0000	DRYUNLOAD	DVALVE	.0	.0

APPENDIX E
DETAILED COST OF ELECTRICITY
RESULTS (DOE MODEL)

TABLE E-1
WANSLEY IGCC/HOT GAS CLEANUP CASE 1
458.4 MW POWER PLANT

CAPITAL REQUIREMENTS (Mid-1990 Dollars)

Total Plant Investment

AREA NO.	PLANT SECTION DESCRIPTION	PROCESS CONT. K\$	COST, K\$ W/O CONT
100	Coal Receiving, Handling & Prep	\$0	\$15,901
150	Limestone Receiving, Hand & Pre	\$0	\$9,072
250	Booster Compression	\$0	\$8,822
300	KRW Gasification	\$5,035	\$55,163
380	Recycle Gas Compression	\$1,430	\$13,939
400	Gas Conditioning	\$4,847	\$35,212
500	External Desulfurization	\$1,548	\$20,171
600	Sulfation	\$2,043	\$13,091
700	Sulfuric Acid Plant	\$0	\$0
900	Gas Turbine System	\$3,630	\$70,780
1000	HRSG System	\$1,277	\$24,929
1100	Steam Turbine System	\$0	\$46,268
1200	Ash & Fines Handling & Disp	\$0	\$5,788
<hr/> Subtotal, Process Plant Cost		<hr/>	\$319,136
General Plant Facilities			\$33,484
Engineering Fees			\$25,470
Process Contingency (Using contingencies listed above)			\$19,810
Project Contingency, 20% of above items			\$79,580
Total Plant Cost (TPC)			\$477,480
Adjustment for Interest and Inflation (AFUDC)			\$93,700
Total Plant Investment (TPI)			\$571,180
Prepaid Royalties			\$1,890
Initial Catalyst and Chemical Inventory			\$10,642
Startup Costs			\$14,727
Spare Parts			\$6,893
Working Capital			\$9,883
Land, 0 Acres			\$0
Total Capital Requirement (TCR)			\$615,215

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

ANNUAL OPERATING COSTS

Capacity Factor = 65%

<u>COST ITEM</u>	<u>QUANTITY</u>	<u>UNIT \$ PRICE</u>	<u>ANNUAL COST, K\$</u>
Ill. No. 6 Coal Fuel Type	4053.0 T/D	\$35.00 /T	\$33,656
Consumable Materials			
Limestone	1085.6 T/D	\$11.20 /T	\$2,885
Nahcolite	4.8 T/D	\$261.25 /T	\$296
Zinc Ferrite	1.1 T/D	\$6,270 /T	\$1,648
Miscellaneous	2509.6 T/D	\$1.00 /T	\$595
Ash/Sorbent Disposal Costs	1231.0 T/D	\$2.94 /T	\$859
Plant Labor			
Oper Labor (incl benef)	23 Men/shift	\$21.80 /Hr.	\$4,380
Supervision & Clerical			\$1,127
Maintenance Costs			\$8,639
Insurance & Local Taxes			\$2,865
Royalties			\$0
Other Operating Costs			\$564
Total Operating Costs			\$57,514
By-Product Credits	0.0 T/D	\$0.00 /T	\$0
Total By-Product Credits			\$0
Net Operating Costs			\$57,514

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

BASES AND ASSUMPTIONS

A. CAPITAL BASES AND DETAILS

	QUANTITY	UNIT \$ PRICE	COST, K\$
Initial Catalyst Inventory			
Gasification	___ lb.	___ /lb.	\$10,583
Initial Chemical Inventory			
Miscellaneous	59000 lb.	\$1.00 /lb.	\$59
Total Catalyst and Chemical Inventory			\$10,642
Startup costs			
Plant modifications,	2% TPI		\$11,424
Operating costs			\$2,239
Fuel			\$1,064
Total Startup Costs			\$14,727
Working capital			
Fuel & Consumables inv.,	60 days supply		\$9,883
By-Product inventory,	30 days supply		\$0
Direct expenses,	0 days		\$0
Total Working Capital			\$9,883

B. ECONOMIC ASSUMPTIONS

Project life	30 Years
Book life	30 Years
Tax life	20 Years

Federal and state income tax rate	37.7%
Tax depreciation method	ACRS
Investment Tax Credit	0.0%

Financial structure

<u>Type of Security</u>	<u>% of Total</u>	<u>Current Dollar</u>		<u>Constant Dollar</u>	
		<u>Cost. %</u>	<u>Ret. %</u>	<u>Cost. %</u>	<u>Ret. %</u>
Debt	45	10.3	4.6	5.6	2.5
Preferred Stock	10	9.8	1.0	5.1	0.5
Common Stock	45	13.8	6.2	8.9	4.0
Discount rate (cost of capital)			11.8		7.0
Inflation rate, % per year			4.5		
Real escalation rates (over inflation)					
Fuel, % per year			0.5		
Operating & Maintenance, % per year			0.0		

C. COST OF ELECTRICITY

The approach to determining the cost of electricity is based upon the methodology described in the Technical Assessment Guide, published by the Electric Power Research Institute, Volume I, EPRI-4463-SR, December 1986. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current \$	Constant \$
Levelizing Factors		
Capital Carrying Charge, 10th year	0.169	0.113
Fuel, 10th year	1.265	1.025
Operating & Maintenance, 10th year	1.234	1.000
	mills/kWh	mills/kWh
Cost of Electricity - Levelized		
Capital Charges	39.8	26.7
Fuel costs	16.3	13.2
Operating & Maintenance	11.3	9.1
Total Cost of Electricity	67.4	49.1

TABLE E-2
WANSLEY IGCC/HOT GAS CLEANUP CASE 2
99.6 MW POWER PLANT

CAPITAL REQUIREMENTS (Mid-1990 Dollars)

Total Plant Investment

AREA NO.	PLANT SECTION DESCRIPTION	PROCESS CONT. K\$	COST, K\$ W/O CONT
100	Coal Receiving, Handling & Prep	\$0	\$8,907
150	Limestone Receiving, Hand & Pre	\$0	\$6,607
250	Booster Compression	\$0	\$2,886
300	KRW Gasification	\$1,527	\$14,178
380	Recycle Gas Compression	\$657	\$4,091
400	Gas Conditioning	\$1,369	\$9,553
500	External Desulfurization	\$593	\$7,524
600	Sulfation	\$482	\$2,768
700	Sulfuric Acid Plant	\$0	\$0
900	Gas Turbine System	\$1,210	\$23,702
1000	HRSB System	\$311	\$6,106
1100	Steam Turbine System	\$0	\$1,945
1200	Ash & Fines Handling & Disp	\$0	\$5,788
Subtotal, Process Plant Cost			\$91,188
General Plant Facilities			\$15,743
Engineering Fees			\$13,176
Process Contingency (Using contingencies listed above)			\$6,149
Project Contingency, 20% of above items			\$25,251
Total Plant Cost (TPC)			\$151,507
Adjustment for Interest and Inflation (AFUDC)			\$28,881
Total Plant Investment (TFI)			\$180,388
Prepaid Royalties			\$601
Initial Catalyst and Chemical Inventory			\$1,883
Startup Costs			\$4,684
Spare Parts			\$4,032
Working Capital			\$2,329
Land, 0 Acres			\$0
Total Capital Requirement (TCR)			\$193,917

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

ANNUAL OPERATING COSTS

Capacity Factor = 65%

<u>COST ITEM</u>	<u>QUANTITY</u>	<u>UNIT \$ PRICE</u>	<u>ANNUAL COST, K\$</u>
Ill. No. 6 Coal Fuel Type	953.4 T/D	\$35.00 /T	\$7,917
Consumable Materials			
Limestone	255.7 T/D	\$11.20 /T	\$679
Nahcolite	1.1 T/D	\$261.25 /T	\$71
Zinc Ferrite	0.2 T/D	\$6,270 /T	\$299
Miscellaneous	1024.5 T/D	\$1.00 /T	\$243
Ash/Sorbent Disposal Costs	289.0 T/D	\$2.94 /T	\$202
Plant Labor			
Oper Labor (incl benef)	15 Men/shift	\$21.80 /Hr.	\$2,857
Supervision & Clerical			\$697
Maintenance Costs			\$2,941
Insurance & Local Taxes			\$909
Royalties			\$0
Other Operating Costs			\$348
Total Operating Costs			\$17,162
By-Product Credits	0.0 T/D	\$0.00 /T	\$0
Total By-Product Credits			\$0
Net Operating Costs			\$17,162

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

BASES AND ASSUMPTIONS

A. CAPITAL BASES AND DETAILS

	QUANTITY	UNIT \$ PRICE	COST, K\$
Initial Catalyst Inventory			
Gasification	___ lb.	___ /lb.	\$1,863
Initial Chemical Inventory			
Miscellaneous	20000 lb.	\$1.00 /lb.	\$20
Total Catalyst and Chemical Inventory			\$1,883
Startup costs			
Plant modifications,	2% TPI		\$3,608
Operating costs			\$826
Fuel			\$250
Total Startup Costs			\$4,684
Working capital			
Fuel & Consumables inv.,	60 days supply		\$2,329
By-Product inventory,	30 days supply		\$0
Direct expenses,	0 days		\$0
Total Working Capital			\$2,329

B. ECONOMIC ASSUMPTIONS

Project life	30 Years				
Book life	30 Years				
Tax life	20 Years				
Federal and state income tax rate	37.7%				
Tax depreciation method	ACRS				
Investment Tax Credit	0.0%				
Financial structure					
	Current Dollar	Constant Dollar			
<u>Type of Security</u>	<u>% of Total</u>	<u>Cost. %</u> <u>Ret. %</u>	<u>Cost. %</u> <u>Ret. %</u>		
Debt	45	10.3	4.6	5.6	2.5
Preferred Stock	10	9.8	1.0	5.1	0.5
Common Stock	45	13.8	6.2	8.9	4.0
Discount rate (cost of capital)			11.8		7.0
Inflation rate, % per year			4.5		
Real escalation rates (over inflation)					
Fuel, % per year			0.5		
Operating & Maintenance, % per year			0.0		

C. COST OF ELECTRICITY

The approach to determining the cost of electricity is based upon the methodology described in the Technical Assessment Guide, published by the Electric Power Research Institute, Volume I, EPRI 4463-SR, December 1986. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current \$	Constant \$
Levelizing Factors		
Capital Carrying Charge, 10th year	0.169	0.113
Fuel, 10th year	1.265	1.025
Operating & Maintenance, 10th year	1.234	1.000
	mills/kWh	mills/kWh
Cost of Electricity - Levelized		
Capital Charges	57.7	38.8
Fuel costs	17.7	14.3
Operating & Maintenance	20.1	16.3
Total Cost of Electricity	95.5	69.4

TABLE E-3
WANSLEY IGCC/HOT GAS CLEANUP CASE 3
231.7 MW POWER PLANT

CAPITAL REQUIREMENTS (Mid-1990 Dollars)

Total Plant Investment

AREA NO.	PLANT SECTION DESCRIPTION	PROCESS CONT. K\$	COST, K\$ W/O CONT
100	Coal Receiving, Handling & Prep	\$0	\$11,405
150	Limestone Receiving, Hand & Pre	\$0	\$7,532
250	Booster Compression	\$0	\$4,450
300	KRW Gasification	\$2,654	\$28,643
380	Recycle Gas Compression	\$736	\$7,054
400	Gas Conditioning	\$2,486	\$17,819
500	External Desulfurization	\$796	\$10,213
600	Sulfation	\$1,132	\$7,062
700	Sulfuric Acid Plant	\$0	\$0
900	Gas Turbine System	\$1,860	\$36,123
1000	HRSO System	\$671	\$13,054
1100	Steam Turbine System	\$0	\$25,643
1200	Ash & Fines Handling & Disp	\$0	\$3,942
<hr/> Subtotal, Process Plant Cost		<hr/>	\$172,940
General Plant Facilities			\$26,307
Engineering Fees			\$19,488
Process Contingency (Using contingencies listed above)			\$10,335
Project Contingency, 20% of above items			\$45,814
Total Plant Cost (TPC)			\$274,884
Adjustment for Interest and Inflation (AFUDC)			\$52,400
Total Plant Investment (TPI)			\$327,284
Prepaid Royalties			\$1,094
Initial Catalyst and Chemical Inventory			\$5,322
Startup Costs			\$8,392
Spare Parts			\$5,142
Working Capital			\$4,957
Land, 0 Acres			\$0
Total Capital Requirement (TCR)			\$352,191

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

ANNUAL OPERATING COSTS

Capacity Factor = 65%

<u>COST ITEM</u>	<u>QUANTITY</u>	<u>UNIT \$ PRICE</u>	<u>ANNUAL COST, K\$</u>
III. No. 6 Coal Fuel Type	2026.6 T/D	\$35.00 /T	\$16,828
Consumable Materials			
Limestone	542.3 T/D	\$11.20 /T	\$1,441
Nahcolite	2.4 T/D	\$261.25 /T	\$148
Zinc Ferrite	0.6 T/D	\$6,270 /T	\$824
Miscellaneous	1517.7 T/D	\$1.00 /T	\$360
Ash/Sorbent Disposal Costs	615.0 T/D	\$2.94 /T	\$429
Plant Labor			
Oper Labor (incl benef)	17 Men/shift	\$21.80 /Hr.	\$3,238
Supervision & Clerical			\$815
Maintenance Costs			\$4,953
Insurance & Local Taxes			\$1,649
Royalties			\$0
Other Operating Costs			\$407
Total Operating Costs			\$31,093
By-Product Credits	0.0 T/D	\$0.00 /T	\$0
Total By-Product Credits			\$0
Net Operating Costs			\$31,093

BASES AND ASSUMPTIONS

A. CAPITAL BASES AND DETAILS

	QUANTITY	UNIT \$ PRICE	COST, K\$
Initial Catalyst Inventory			
Gasification	___ lb.	___ /lb.	\$5,292
Initial Chemical Inventory			
Miscellaneous	30000 lb.	\$1.00 /lb.	\$30
Total Catalyst and Chemical Inventory			\$5,322
Startup costs			
Plant modifications,	2% TPI		\$6,546
Operating costs			\$1,314
Fuel			\$532
Total Startup Costs			\$8,392
Working capital			
Fuel & Consumables inv.,	60 days supply		\$4,957
By-Product inventory,	30 days supply		\$0
Direct expenses,	0 days		\$0
Total Working Capital			\$4,957

B. ECONOMIC ASSUMPTIONS

Project life	30 Years
Book life	30 Years
Tax life	20 Years

Federal and state income tax rate	37.7%
Tax depreciation method	ACRS
Investment Tax Credit	0.0%

Financial structure

Type of Security	% of Total	Current Dollar		Constant Dollar	
		Cost, %	Ret. %	Cost, %	Ret. %
Debt	45	10.3	4.6	5.6	2.5
Preferred Stock	10	9.8	1.0	5.1	0.5
Common Stock	45	13.8	6.2	8.9	4.0
Discount rate (cost of capital)			11.8		7.0
Inflation rate, % per year			4.5		
Real escalation rates (over inflation)					
Fuel, % per year			0.5		
Operating & Maintenance, % per year			0.0		

C. COST OF ELECTRICITY

The approach to determining the cost of electricity is based upon the methodology described in the Technical Assessment Guide, published by the Electric Power Research Institute, Volume I, EPRI-4463-SR, December 1986. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current \$	Constant \$
Levelizing Factors		
Capital Carrying Charge, 10th year	0.169	0.113
Fuel, 10th year	1.265	1.025
Operating & Maintenance, 10th year	1.234	1.000
	mills/kWh	mills/kWh
Cost of Electricity - Levelized		
Capital Charges	45.1	30.3
Fuel costs	16.1	13.1
Operating & Maintenance	13.3	10.8
Total Cost of Electricity	74.5	54.2

TABLE E-4
WANSLEY IGCC/HOT GAS CLEANUP CASE 4
231.7 MW POWER PLANT

CAPITAL REQUIREMENTS (Mid- 1990 Dollars)

Total Plant Investment

AREA NO.	PLANT SECTION DESCRIPTION	PROCESS CONT. K\$	COST, K\$ W/O CONT
100	Coal Receiving, Handling & Prep	\$0	\$15,945
150	Limestone Receiving, Hand & Pre	\$0	\$9,141
250	Booster Compression	\$0	\$11,456
300	KRW Gasification	\$5,150	\$49,781
380	Recycle Gas Compression	\$2,312	\$14,965
400	Gas Conditioning	\$5,117	\$37,144
500	External Desulfurization	\$2,237	\$29,336
600	Sulfation	\$1,442	\$9,133
700	Sulfuric Acid Plant	\$0	\$0
900	Gas Turbine System	\$4,676	\$91,005
1000	HRSG System	\$1,132	\$22,077
1100	Steam Turbine System	\$0	\$6,995
1200	Ash & Slimes Handling & Disp	\$0	\$5,658
<hr/> Subtotal, Process Plant Cost			<hr/> \$302,638
General Plant Facilities			\$29,379
Engineering Fees			\$24,464
Process Contingency (Using contingencies listed above)			\$22,066
Project Contingency, 20% of above items			\$75,709
Total Plant Cost (TPC)			\$454,256
Adjustment for Interest and Inflation (AFUDC)			\$89,142
Total Plant Investment (TPI)			\$543,398
Prepaid Royalties			\$1,782
Initial Catalyst and Chemical Inventory			\$7,504
Startup Costs			\$13,960
Spare Parts			\$7,084
Working Capital			\$9,213
Land, 0 Acres			\$0
Total Capital Requirement (TCR)			\$582,941

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

ANNUAL OPERATING COSTS

Capacity Factor = 65%

<u>COST ITEM</u>	<u>QUANTITY</u>	<u>UNIT \$ PRICE</u>	<u>ANNUAL COST. K\$</u>
Ill. No. 6 Coal Fuel Type	3812.6 T/D	\$35.00 /T	\$31,659
Consumable Materials			
Limestone	1020.5 T/D	\$11.20 /T	\$2,712
Nahcolite	4.6 T/D	\$261.25 /T	\$284
Zinc Ferrite	0.8 T/D	\$6,270 /T	\$1,193
Miscellaneous	2452.2 T/D	\$1.00 /T	\$582
Ash/Sorbent Disposal Costs	1158.0 T/D	\$2.94 /T	\$808
Plant Labor			
Oper Labor (incl benef)	21 Men/shift	\$21.80 /Hr.	\$3,999
Supervision & Clerical			\$1,051
Maintenance Costs			\$8,558
Insurance & Local Taxes			\$2,726
Royalties			\$0
Other Operating Costs			\$526
Total Operating Costs			\$54,097
By-Product Credits	0.0 T/D	\$0.00 /T	\$0
Total By-Product Credits			\$0
Net Operating Costs			\$54,097

BASES AND ASSUMPTIONS

A. CAPITAL BASES AND DETAILS

	QUANTITY	UNIT \$ PRICE	COST, K\$
Initial Catalyst Inventory			
Gasification	— lb.	— /lb.	\$7,445
Initial Chemical Inventory			
Miscellaneous	59000 lb.	\$1.00 /lb.	\$59
Total Catalyst and Chemical Inventory			\$7,504
Startup costs			
Plant modifications,	2% TPI		\$10,868
Operating costs			\$2,091
Fuel			\$1,001
Total Startup Costs			\$13,960
Working capital			
Fuel & Consumables inv.,	60 days supply		\$9,213
By-Product inventory,	30 days supply		\$0
Direct expenses,	0 days		\$0
Total Working Capital			\$9,213

B. ECONOMIC ASSUMPTIONS

Project life	30 Years
Book life	30 Years
Tax life	20 Years

Federal and state income tax rate	37.7%
Tax depreciation method	ACRS
Investment Tax Credit	0.0%

Financial structure

Type of Security	% of Total	Current Dollar		Constant Dollar	
		Cost. %	Ret. %	Cost. %	Ret. %
Debt	45	10.3	4.6	5.6	2.5
Preferred Stock	10	9.8	1.0	5.1	0.5
Common Stock	45	13.8	6.2	8.9	4.0
Discount rate (cost of capital)			11.8		7.0

Inflation rate, % per year	4.5
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Real escalation rates (over inflation)

Fuel, % per year	0.5
Operating & Maintenance, % per year	0.0

C. COST OF ELECTRICITY

The approach to determining the cost of electricity is based upon the methodology described in the Technical Assessment Guide, published by the Electric Power Research Institute, Volume I, EPRI-4463-SR, December 1986. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current \$	Constant \$
Levelizing Factors		
Capital Carrying Charge, 10th year	0.169	0.113
Fuel, 10th year	1.265	1.025
Operating & Maintenance, 10th year	1.234	1.000
	mills/kWh	mills/kWh
Cost of Electricity - Levelized		
Capital Charges	43.3	29.1
Fuel costs	17.6	14.3
Operating & Maintenance	12.2	9.9
Total Cost of Electricity	73.1	53.2

TABLE E-5
WANSLEY IGCC/HOT GAS CLEANUP CASE 5
420.2 MW POWER PLANT

CAPITAL REQUIREMENTS (Mid- 1990 Dollars)

Total Plant Investment

AREA NO.	PLANT SECTION DESCRIPTION	PROCESS CONT. K\$	COST, K\$ W/O CONT
100	Coal Receiving, Handling & Prep	\$0	\$16,245
150	Limestone Receiving, Hand & Pre	\$0	\$0
250	Booster Compression	\$0	\$9,272
300	KRW Gasification	\$7,953	\$63,074
380	Recycle Gas Compression	\$1,853	\$18,044
400	Gas Conditioning	\$5,336	\$38,645
500	External Desulfurization	\$3,804	\$32,897
600	Sulfation	\$0	\$0
700	Sulfuric Acid Plant	\$0	\$25,739
900	Gas Turbine System	\$3,637	\$70,965
1000	HRSG System	\$1,643	\$32,092
1100	Steam Turbine System	\$0	\$41,604
1200	Ash & Fines Handling & Disp	\$0	\$4,491
<hr/> Subtotal, Process Plant Cost		<hr/>	<hr/> \$353,068
General Plant Facilities			\$33,365
Engineering Fees			\$26,103
Process Contingency (Using contingencies listed above)			\$24,226
Project Contingency, 20% of above items			\$87,352
Total Plant Cost (TPC)			\$524,114
Adjustment for Interest and Inflation (AFUDC)			\$102,851
Total Plant Investment (TPI)			\$626,965
Prepaid Royalties			\$2,063
Initial Catalyst and Chemical Inventory			\$13,512
Startup Costs			\$16,700
Spare Parts			\$6,219
Working Capital			\$11,813
Land, 0 Acres			\$0
Total Capital Requirement (TCR)			\$677,271

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

ANNUAL OPERATING COSTS

Capacity Factor = 65%

<u>COST ITEM</u>	<u>QUANTITY</u>	<u>UNIT \$ PRICE</u>	<u>ANNUAL COST, K\$</u>
Ill. No. 6 Coal Fuel Type	4043.5 T/D	\$35.00 /T	\$33,576
Consumable Materials			
Limestone	0.0 T/D	\$11.20 /T	\$0
Nahcolite	4.8 T/D	\$261.25 /T	\$296
Zinc Ferrite	6.5 T/D	\$6,270 /T	\$9,653
Miscellaneous	6105.9 T/D	\$1.00 /T	\$1,449
Ash/Sorbent Disposal Costs	540.0 T/D	\$4.40 /T	\$564
Plant Labor			
Oper Labor (incl benef)	24 Men/shift	\$21.80 /Hr.	\$4,571
Supervision & Clerical			\$1,165
Maintenance Costs			\$9,839
Insurance & Local Taxes			\$3,145
Royalties			\$0
Other Operating Costs			\$583
 Total Operating Costs			\$64,652
By-Product Credits			
SULFURIC ACID	280.0 T/D	\$52.25 /T	\$3,471
 Total By-Product Credits			\$3,471
Net Operating Costs			\$61,368

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

BASES AND ASSUMPTIONS

A. CAPITAL BASES AND DETAILS

	QUANTITY	UNIT \$ PRICE	COST, K\$
Initial Catalyst Inventory			
Gasification	___ lb.	___ /lb.	\$13,422
Initial Chemical Inventory			
Miscellaneous	90000 lb.	\$1.00 /lb.	\$90
Total Catalyst and Chemical Inventory			\$13,512
Startup costs			
Plant modifications,	2% TPI		\$12,539
Operating costs			\$3,099
Fuel			\$1,061
Total Startup Costs			\$16,700
Working capital			
Fuel & Consumables inv.,	60 days supply		\$11,374
By-Product inventory,	30 days supply		\$439
Direct expenses,	0 days		\$0
Total Working Capital			\$11,813

B. ECONOMIC ASSUMPTIONS

Project life	30 Years
Book life	30 Years
Tax life	20 Years

Federal and state income tax rate	37.7%
Tax depreciation method	ACRS
Investment Tax Credit	0.0%

Financial structure

<u>Type of Security</u>	<u>% of Total</u>	<u>Current Dollar</u>		<u>Constant Dollar</u>	
		<u>Cost. %</u>	<u>Ret. %</u>	<u>Cost. %</u>	<u>Ret. %</u>
Debt	45	10.3	4.6	5.6	2.5
Preferred Stock	10	9.8	1.0	5.1	0.5
Common Stock	45	13.8	6.2	8.9	4.0
Discount rate (cost of capital)			11.8		7.0
Inflation rate, % per year			4.5		
Real escalation rates (over inflation)					
Fuel, % per year			0.5		
Operating & Maintenance, % per year			0.0		

C. COST OF ELECTRICITY

The approach to determining the cost of electricity is based upon the methodology described in the Technical Assessment Guide, published by the Electric Power Research Institute, Volume I, EPRI-4463-SR, December 1986. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current \$	Constant \$
Levelizing Factors		
Capital Carrying Charge, 10th year	0.169	0.113
Fuel, 10th year	1.265	1.025
Operating & Maintenance, 10th year	1.234	1.000
	mills/kWh	mills/kWh
Cost of Electricity - Levelized		
Capital Charges	47.8	32.1
Fuel costs	17.7	14.4
Operating & Maintenance	14.3	11.6
Total Cost of Electricity	79.9	58.1

TABLE E-6
WANSLEY IGCC/HOT GAS CLEANUP CASE 6
409.6 MW POWER PLANT

CAPITAL REQUIREMENTS (Mid-1990 Dollars)

Total Plant Investment

AREA NO.	PLANT SECTION DESCRIPTION	PROCESS CONT. K\$	COST, K\$ W/O CONT
100	Coal Receiving, Handling & Prep	\$0	\$18,553
150	Limestone Receiving, Hand & Pre	\$0	\$7,628
250	Booster Compression	\$0	\$7,933
300	KRW Gasification	\$6,358	\$40,622
380	Recycle Gas Compression	\$619	\$5,855
400	Gas Conditioning	\$1,925	\$12,740
500	External Desulfurization	\$0	\$0
600	Sulfation	\$3,488	\$21,383
700	Sulfuric Acid Plant	\$0	\$0
900	Gas Turbine System	\$3,629	\$70,737
1000	HRSG System	\$1,277	\$24,914
1100	Steam Turbine System	\$0	\$41,926
1200	Ash & Fines Handling & Disp	\$0	\$4,780
<hr/> Subtotal, Process Plant Cost		<hr/>	\$257,071
General Plant Facilities			\$35,591
Engineering Fees			\$36,121
Process Contingency (Using contingencies listed above)			\$17,296
Project Contingency, 20% of above items			\$69,216
Total Plant Cost (TPC)			\$415,295
Adjustment for Interest and Inflation (AFUDC)			\$81,496
Total Plant Investment (TPI)			\$496,791
Prepaid Royalties			\$1,644
Initial Catalyst and Chemical Inventory			\$59
Startup Costs			\$12,400
Spare Parts			\$5,378
Working Capital			\$7,688
Land, 0 Acres			\$0
Total Capital Requirement (TCR)			\$523,960

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

ANNUAL OPERATING COSTS

Capacity Factor = 65%

<u>COST ITEM</u>	<u>QUANTITY</u>	<u>UNIT \$ PRICE</u>	<u>ANNUAL COST, K\$</u>
Ill. No. 6 Coal Fuel Type	3433.8 T/D	\$35.00 /T	\$28,513
Consumable Materials			
Limestone	444.7 T/D	\$11.20 /T	\$1,182
Nahcolite	0.0 T/D	\$261.25 /T	\$0
Zinc Ferrite	0.0 T/D	\$6,270 /T	\$0
Miscellaneous	2963.5 T/D	\$1.00 /T	\$703
Ash/Sorbent Disposal Costs	798.0 T/D	\$2.94 /T	\$557
Plant Labor			
Oper Labor (incl benef)	21 Men/shift	\$21.80 /Hr.	\$3,999
Supervision & Clerical			\$1,051
Maintenance Costs			\$7,189
Insurance & Local Taxes			\$2,492
Royalties			\$0
Other Operating Costs			\$526
Total Operating Costs			\$46,212
By-Product Credits			
	0.0 T/D	\$0.00 /T	\$0
Total By-Product Credits			\$0
Net Operating Costs			\$46,212

**Assessment of Coal Gasification/Hot Gas
Cleanup Based Advanced Gas Turbine Systems**

BASES AND ASSUMPTIONS

A. CAPITAL BASES AND DETAILS

	QUANTITY	UNIT	\$ PRICE	COST, K\$
Initial Catalyst Inventory				
Gasification	—	lb.	— /lb.	\$0
Initial Chemical Inventory				
Miscellaneous	59000	lb.	\$1.00 /lb.	\$59
Total Catalyst and Chemical Inventory				\$59
Startup costs				
Plant modifications,	2%	TPI		\$9,936
Operating costs				\$1,563
Fuel				\$901
Total Startup Costs				\$12,400
Working capital				
Fuel & Consumables inv.,	60	days supply		\$7,688
By-Product inventory,	30	days supply		\$0
Direct expenses,	0	days		\$0
Total Working Capital				\$7,688

B. ECONOMIC ASSUMPTIONS

Project life	30 Years
Book life	30 Years
Tax life	20 Years

Federal and state income tax rate	37.7%
Tax depreciation method	ACRS
Investment Tax Credit	0.0%

Financial structure

Type of Security	% of Total	Current Dollar		Constant Dollar	
		Cost. %	Ret. %	Cost. %	Ret. %
Debt	45	10.3	4.6	5.6	2.5
Preferred Stock	10	9.8	1.0	5.1	0.5
Common Stock	45	13.8	6.2	8.9	4.0
Discount rate (cost of capital)			11.8		7.0
Inflation rate, % per year			4.5		
Real escalation rates (over inflation)					
Fuel, % per year			0.5		
Operating & Maintenance, % per year			0.0		

C. COST OF ELECTRICITY

The approach to determining the cost of electricity is based upon the methodology described in the Technical Assessment Guide, published by the Electric Power Research Institute, Volume I, EPRI-4463-SR, December 1986. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current \$	Constant \$
Levelizing Factors		
Capital Carrying Charge, 10th year	0.169	0.113
Fuel, 10th year	1.265	1.025
Operating & Maintenance, 10th year	1.234	1.000
	mills/kWh	mills/kWh
Cost of Electricity - Levelized		
Capital Charges	37.9	25.5
Fuel costs	15.5	12.5
Operating & Maintenance	9.4	7.6
Total Cost of Electricity	62.8	45.6

APPENDIX F
THERMOECONOMIC ANALYSIS DETAILS

**ASSESSMENT OF COAL GASIFICATION/HOT GAS
CLEANUP BASED ADVANCED GAS TURBINE SYSTEMS**

EXERGETIC AND THERMOECONOMIC EVALUATION

Prepared for

**SOUTHERN COMPANY SERVICES, INC.
BIRMINGHAM, ALABAMA**

**DOE Contract No. DE-F021-89MC26019
SCS Contract No. 196-89-004**

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ABSTRACT

This study provides an evaluation, from the thermodynamic and economic viewpoints, of two design options for an integrated gasification-combined-cycle (IGCC) power plant. The two design options refer to an IGCC plant that uses an oxygen-blown KRW gasifier and cold gas cleanup [1] (reference case) and to case 1, which uses an air-blown KRW gasifier and hot gas cleanup and which was developed in the current DOE project.

The detailed design configurations including material and heat balances and cost data were developed by the M. W. Kellogg Company for the gasification island, and by Southern Company Services for the power island. The data referring to the gas turbine system were provided by the General Electric Company.

The objective of this study is to

- Compare the two design options from the thermodynamic viewpoint, and
- Conduct a detailed evaluation of the air-blown/hot-gas-cleanup IGCC design option from the thermoeconomic viewpoint.

The first point includes a comparison of the real energy waste in the corresponding areas of each design option and an evaluation of the potential for improving each option from the thermodynamic viewpoint. The exergy destruction in each plant component and the exergy losses associated with material streams rejected to the environment provide the basis for comparing the overall plant

performance as well as the performance of oxygen-blown versus air-blown KRW gasifiers and cold gas versus hot gas cleanup.

The thermoeconomic evaluation of case 1 discusses the performance of the most important plant components or group of components from the thermodynamic and economic viewpoints as well as the potential for improving the overall plant cost effectiveness through design parameter changes. Design inefficiencies are identified and suggestions for improving the cost effectiveness are presented.

This report was prepared in order to be incorporated by Southern Company Services into the final report to be submitted to the U.S. DOE as a separate Section F (with an associated Appendix F1) and with a minimum number of changes. This report is not intended to be used alone. The overall technical approach, the assumptions, and the description of each design option are discussed in the reports being prepared by the M. W. Kellogg Company and Southern Company Services.

Section F

EXERGETIC AND THERMOECONOMIC EVALUATION (OF CASE 1 AND REFERENCE CASE)

This section describes the results of detailed exergy (second-law) analyses and thermoeconomic analyses conducted for cases 1, 1A and 1B and for the reference case. Cases 1A, 1 and 1B refer to a gasification temperature of 1800°, 1900° and 2000° F, respectively, whereas the reference case uses the IGCC power plant presented in Reference [1] as a basis. Some modifications in the capacity and the steam cycle design of the plant presented in Reference [1] were made in the reference case in order to make this design comparable with the design of case 1.

Initially, a brief introduction to the thermoeconomic analysis is given. Subsequently, the results of the analyses are presented and design options that may lead to a reduction in the cost of electricity are discussed. Finally, the conclusions from the exergetic and thermoeconomic analyses are summarized. More detailed results than the ones shown in this section are presented in Appendix Y.

F-1 INTRODUCTION TO THERMOECONOMIC ANALYSIS

Exergy is that part of energy that can be transformed into any other form of energy under the conditions imposed by a given environment. The exergy (useful energy) of an energy carrier expresses the maximum capability of the carrier to cause changes. Thus, exergy is not only an objective measure of the thermodynamic

value of an energy carrier, but is also very closely related to the economic value of the carrier.

A thermoeconomic (or exergoeconomic) analysis is an appropriate combination of exergetic and economic analyses conducted at the plant component level. A complete thermoeconomic analysis consists of (a) an exergy analysis, (b) an economic analysis, (c) exergy costing, and (d) a thermoeconomic evaluation of each plant component. This study discusses the results of the exergy analysis and the thermoeconomic evaluation.

In the exergy analysis we calculate, among others, the exergy of all streams in the plant as well as the rate of exergy destruction, $\dot{E}_D$, and the exergetic (second-law) efficiency, ζ , for each plant component. For the entire thermoeconomic analysis, it is helpful to define a "fuel" and a "product" for each component. The "product" represents the desired result produced by the plant component being considered. The definition of the "product" must be consistent with the purpose for using the component. The "fuel" represents the resources which were spent to generate the "product". In the following, the words "fuel" and "product" for a component are used without quotation marks.

The exergetic efficiency of the k-th plant component is defined as the ratio between the exergy value of the product, $\dot{E}_{P,k}$, and the exergy value of the fuel, $\dot{E}_{F,k}$.

$$\zeta_k = \frac{\dot{E}_{P,k}}{\dot{E}_{F,k}} \quad (1)$$

More details on this approach to the thermoeconomic analysis are given in References [2], [3] and [4].

The rate of exergy destruction in the k-th component (including, where appropriate, the exergy loss, which is associated with streams not further used in the plant) is then calculated from the following Equation

$$\dot{E}_{D,k} = \dot{E}_{F,k} - \dot{E}_{P,k} \quad (2)$$

Exergy costing is based on the finding that exergy is the only rational basis for assigning costs to streams and to "energy waste". In exergy costing, we calculate the cost of each stream in the plant, as well as the cost of (a) providing the fuel, (b) generating the product, and (c) the cost of exergy destruction in each plant component.

The thermoeconomic evaluation of the k-th plant component is based on the following variables:

1. The rate of exergy destruction, $\dot{E}_{D,k}$, in this component, Equation (2).
2. The dimensionless ratio

$$r_k = \frac{\dot{E}_{D,k}}{\dot{E}_{\text{coal}}} \quad (3)$$

which indicates what percentage of the coal exergy was destroyed in the component being considered.

3. The exergetic efficiency, ζ_k , Equation (1), which shows

what percentage of the fuel exergy of the k-th component can be found in the exergy of the product.

4. The relative cost difference, d , between cost per exergy unit of product ($C_{P,k}$) and cost per exergy unit of fuel ($C_{F,k}$) defined by

$$d_k \equiv \frac{C_{P,k} - C_{F,k}}{C_{F,k}} = \frac{1 - \zeta_k}{\zeta_k} + \frac{\dot{Z}_k}{C_{F,k} \dot{E}_{F,k}} \quad (4)$$

Here, $\dot{Z}_k$ expresses the contribution of the investment costs and the operating (excluding fuel) and maintenance (O&M) costs associated with the k-th plant component. Equation (4) reveals the real cost sources in the k-th component, which are irreversibilities (first term on the right side) and capital and O&M expenditures. In general, the higher the relative cost difference, d_k , the more attention should be paid to the k-th component.

5. The cost of exergy destruction in the component being considered, $\dot{D}_k$. In this study, $\dot{D}_k$ is calculated from the following relationship

$$\dot{D}_k = C_{F,k} \dot{E}_{D,k} \quad (5)$$

6. The thermoeconomic factor f_k

$$f_k = \frac{\dot{Z}_k}{\dot{Z}_k + \dot{D}_k} \quad (6)$$

which expresses the contribution of the capital and O&M costs to the cost increase d_k .

7. The optimal exergetic efficiency from the economic viewpoint, ζ_k^{opt} . A condition for calculating this efficiency is that the investment costs associated with the k-th component (I_k) can be approximated by the following Equation

$$I_k = I_{o,k} + g_k \left(\frac{\zeta_k}{1 - \zeta_k} \right)^{n_k} E_{p,k}^{m_k} \quad (7)$$

I_k , g_k , n_k and m_k are constants depending only on the plant component being considered. The cost values provided for cases 1, 1A and 1B do not allow an equation of the form of Equation (7) to be calculated for all plant components. The reasons for this are discussed in section F-2. The optimal exergetic efficiency from the economic viewpoint, ζ_k^{opt} , was calculated here only for a small number of plant components of the gasification island. A comparison of the optimal efficiency with the actual one for a plant component indicates whether it is more cost effective for the electricity generated by the total plant to increase the capital expenditures associated with the component and reduce the cost of exergy destruction in the component, or vice versa.

In the following, we discuss the results obtained from a detailed thermoeconomic evaluation of case 1 (including cases 1A and 1B) and from the exergy analysis of the reference case.

F-2 THERMOECONOMIC EVALUATION OF THE DESIGN OF CASE 1

Table 1 shows a comparison of selected design variables for cases 1A, 1 and 1B. Table 2 shows the exergy destruction, the dimensionless ratio r_x given in Equation (3), and the exergetic efficiency of selected plant components. More detailed results are shown in Tables 1 through 8 of Appendix F1. The investment cost values used for equipment items include only direct material and subcontract costs. Thermoeconomic analyses were conducted for capacity factor values of 65% and 85%. If the capacity factor is not given, a value of 65% is implied.

1. Gasifier

As the gasifier operating temperature increases from 1800°F (case 1A) to 1900°F (case 1) and then to 2000°F (case 1B) the thermoeconomic variables vary as indicated in Table 3.

In the three cases being considered here, the gasifier has the highest exergy destruction among all plant components. Approximately 18% to 19% (r value) of the coal exergy is destroyed in the gasifier alone. Out of the 217.0 MW of exergy destruction in case 1, 28.3 MW and 8.0 MW are due to the preheating (from the inlet temperature to the gasifier operating temperature) of nitrogen and oxygen in the air, respectively, and 5.1 MW of exergy destruction are caused by the vaporization and superheating of the water contained in the coal (4.98 wt %). The pressure drop in the gasifier causes approximately 2 MW of exergy destruction. The remaining 173.6 MW of exergy destruction are mainly due to the irreversibilities of the chemical reactions in the gasifier.

TABLE 1

COMPARISON OF SELECTED VARIABLES
FOR FOUR IGCC DESIGN CONFIGURATIONS

VARIABLE	CASE 1A	CASE 1	CASE 1B	REF CASE
Gasifier operating temp. [F]	1800.00	1900.00	2000.00	1850.00
Coal flow rate [kg/s]	38.89	39.81	41.06	40.37
Coal exergy flow rate [MW]	1149.96	1177.13	1213.92	1116.57
Gasification pressure [bar]	27.58	27.58	27.58	32.06
Temperature at the outlet of:				
- product gas cooler [K]	935.93	946.48	929.26	510.93
Air to Gas Turbine system [kg/s]	755.26	753.32	750.72	760.00
Gas turbine firing temp. [K]	1497.00	1497.00	1497.00	1497.00
Gas temp. at the HRSG inlet	865.37	865.37	865.37	873.71
Ambient temperature [K]	305.37	305.37	305.37	305.37
Main steam temperature [K]	810.93	810.93	810.93	794.26
Main steam pressure [bar]	99.97	99.97	99.97	93.08
Reheat steam pressure [bar]	24.13	24.13	24.13	24.13
Condenser pressure [bar]	0.1186	0.1186	0.1186	0.1186
Steam turbine gross power [MW]	175.02	182.20	191.65	159.08
Gas turbine gross power [MW]	311.64	311.64	311.64	298.80
Total net power [MW]	451.75	458.41	467.19	413.19
Ratio of electric power generated in the gas turbines and steam turbines	1.78	1.71	1.63	1.87
Thermal efficiency [%]	40.05	39.70	39.24	37.72
Total exergetic efficiency [%]	39.28	38.94	38.49	37.01
Gasification Island Exergetic Efficiency [%]	77.68	76.91	76.09	81.73
Heat Rate [BTU/KWH]	8520	8594	8696	9045

TABLE 2

EXERGY DESTRUCTION, DIMENSIONLESS RATIO BETWEEN EXERGY DESTRUCTION AND EXERGY OF COAL (r)
AND EXERGETIC EFFICIENCY FOR CASES 1A, 1, 1B AND THE REFERENCE CASE

COMPONENT	CASE 1A			CASE 1			CASE 1B			REFERENCE CASE		
	ED [MW]	RATIO [%]	EXERG. EFF. [%]	ED [MW]	RATIO [%]	EXERG. EFF. [%]	ED [MW]	RATIO [%]	EXERG. EFF. [%]	ED [MW]	RATIO [%]	EXERG. EFF. [%]
OXID. FEED SYSTEM	7.59	0.66	87.9	8.07	0.69	87.9	8.71	0.72	87.9	17.49	1.57	39.8
OXYGEN PREHEATER	--	--	--	--	--	--	--	--	--	0.22	0.02	70.9
GASIFIER	208.13	18.10	82.7	216.96	18.43	82.4	231.33	19.06	81.8	135.42	12.13	89.3
CYCLONES & MIX	1.67	0.15	99.2	1.74	0.15	99.2	1.84	0.15	99.3	0.33	0.00	99.9
RADIANT COOLER	--	--	--	--	--	--	--	--	--	15.79	1.41	66.5
CONVECTIVE COOLER	--	--	--	--	--	--	--	--	--	9.80	0.88	73.3
PROD. GAS COOLER (INCLUDING DRUM)	19.68	1.71	67.3	23.92	2.03	67.0	33.67	2.61	65.8	--	--	--
GAS CONDITIONING	20.39	1.77	89.1	20.83	1.77	89.3	21.00	1.73	89.5	--	--	--
SULFATOR	14.33	1.23	56.8	15.10	1.28	55.5	16.17	1.33	53.6	--	--	--
ZINC FERR SYSTEM	6.81	0.59	99.3	8.66	0.74	99.1	5.86	0.48	99.4	--	--	--
EXIT GAS COOLER	6.81	0.59	73.5	6.96	0.59	73.9	7.38	0.61	73.6	--	--	--
RECYCLE GAS CLR	1.57	0.14	70.0	1.55	0.13	70.0	1.58	0.13	70.0	--	--	--
REC. GAS COMPR. 1	0.43	0.04	70.0	0.42	0.04	70.0	0.43	0.04	70.0	0.02	0.00	83.7
REC. GAS COMPR. 2	0.12	0.01	58.6	0.12	0.01	58.8	0.13	0.01	58.8	--	--	--
G.T. & AIR COMPR.	80.52	7.00	79.5	82.65	7.02	79.0	85.37	7.03	78.5	82.88	7.42	78.3
COMBUST. CHAMBER	192.77	16.76	82.5	192.57	16.36	82.6	192.29	15.84	82.6	225.17	20.17	80.2
HRSG	35.82	3.12	84.2	35.24	2.99	84.6	35.60	2.93	84.4	38.79	3.47	82.9
STEAM TURBINES	28.97	2.52	85.8	30.10	2.55	85.8	31.56	2.60	85.9	27.97	2.51	85.0
REM. STEAM CYCLE	19.05	1.66	--	19.68	1.67	--	20.72	1.71	--	14.72	1.31	--
OTHER COMPONENTS	0.27	0.02	--	0.59	0.05	--	0.64	0.05	--	45.28	4.06	--
SUM	644.95	55.98	--	665.16	56.50	--	692.23	57.02	--	613.88	54.98	--
EXERGY LOSSES	55.13	4.79	--	55.46	4.71	--	56.39	4.65	--	91.70	8.21	--
EXERGY OTHER THAN COAL EXERGY TO TOTAL PLANT	698.20	60.72	39.3	718.72	61.06	38.9	746.73	61.51	38.5	703.38	63.00	37.0

TABLE 3
VARIATION OF THERMOECONOMIC VARIABLES FOR THE GASIFIER
IN CASES 1A, 1 AND 1B

The values given in parentheses for cases 1A and 1B refer
to the percentage change compared with case 1

VARIABLE	CASE				
	1A		1	1B	
Exergy Destruction, $\dot{E}_D$ [MW]	208.1	(-4.10%)	217.0	231.3	(+6.59%)
Dimensionless Ratio r	0.181	(-1.63%)	0.184	0.191	(+3.80%)
Exergetic Efficiency, ξ [%]	82.70	(+0.36%)	82.40	81.80	(-0.73%)
Fuel Exergy, $\dot{E}_F$ [MW]	1204	(-2.19%)	1231	1270	(+3.20%)
Product Exergy, $\dot{E}_P$ [MW]	996	(-1.80%)	1014	1039	(+2.49%)
Investment Cost Contri- bution, I [K\$]	20801	(+19.53%)	17402	16178	(-7.04%)
Cost per Unit of Fuel Exergy, c_F [\$/GJ]	2.549	(-0.27%)	2.556	2.571	(+0.59%)
Cost per Unit of Product Exergy, c_P [\$/GJ]	3.238	(+0.02%)	3.231	3.260	(+0.90%)
Relative Cost Differ- ence, d [%]	27.0	(+2.3%)	26.4	26.8	(+1.5%)
Cost Rate of Exergy De- struction, $\dot{D}$ [\$/hr]	1910	(-4.31%)	1996	2141	(+7.26%)
Thermoeconomic Factor, f [%]	22.64	(+19.28)	18.98	16.88	(-11.06%)

With increasing gasifier operating temperature, T_0 , both the exergetic efficiency, ζ , and the investment costs, I , decrease, whereas the coal throughput and, consequently the exergy flow rates of fuel, $\dot{E}_f$, and product, $\dot{E}_p$, increase. The data for I , ζ and $\dot{E}_p$ given in Table 3 do not support an equation of the form of Equation (7). When going from a gasifier operating temperature of 1900° F to a value of 1800° F, the effect of efficiency (exponent n in Equation (7)) on the investment costs seems to be extremely high. The costs increase by 19.5%, whereas the efficiency increases by only 0.36%. In the T_0 range between 1900° F and 2000° F, however, the effect of efficiency (-0.78%) on investment costs (-7.04%) is much more weak than between 1800° F and 1900° F. Also, the effect of capacity (exponent m in Equation (6)) on the investment costs is stronger in the former than in the latter interval of gasifier operating temperature. For these reasons, the optimal exergetic efficiency from the cost viewpoint cannot be calculated for the gasifier.

If we, however, use the gasifier operating temperature as the only decision variable, the values of cost per unit of product exergy, c_p , (Table 3 and Figure 1) indicate that the cost minimum for the gasifier is obtained in the T_0 range between 1850° F and 1870° F. This optimum value is obtained by considering the gasifier only and does not necessarily need to be the optimum gasifier operating temperature when the total plant is considered. However, because of the position of the gasifier (near the total plant inlet) and its importance to the IGCC plant, the optimal T_0 value

Figure 1. Cost per Unit of Product Exergy in the Gasifier as a Function of Operating Temperature in Case 1

for the total plant should not be far from the above T_0 range. In fact, detailed simulations and economic analyses conducted for several total plant configurations (similar to those of cases 1A, 1 and 1B) in T_0 steps of 10°F in the range of $T_0 = 1800$ -2000°F confirm this hypothesis. For a total plant capacity factor of 65%, the minimum cost of electricity is obtained at $T_0^{\text{opt}} = 1870^\circ\text{F}$ while at a capacity factor of 85% the optimal gasifier operating temperature is 1855°F. The design changes discussed in the following might slightly affect these optimal values.

2. Cyclone (320L) And Recycle Gas Mixing

The exergy destruction in the cyclone ranges from 1.67 MW in case 1A to 1.84 MW in case 1B. It is mainly caused by the mixing of the recycle gas with the solids removed from the gas stream. Although the cyclone is not an energy conversion component, the actual cyclone exergetic efficiency (99.2%-99.3%) is very close to the calculated optimal one from the economic viewpoint.

3. Product Gas Cooler (360C) and Heat Recovery Steam Drum (360F)

Depending on the case being considered, the exergy destruction in the product gas cooler and the heat recovery steam drum (HRSD) varies from 19.67 MW (case 1A) to 31.67 MW (case 1B). This represents 1.71% and 2.61% of the coal exergy, respectively. The exergy destruction is mainly caused by the heat transfer from gas to steam.

With increasing gasifier operating temperature both the investment costs, I , and the amount of steam raised in the components being considered (and consequently the exergy flow rate of product, $\dot{E}_p$) increase, whereas the exergetic efficiency, ζ , decreases. The values of I , ζ and $\dot{E}_p$ given in Table 4 do not support an equation similar to Equation (7), and, thus, the economically optimal exergetic efficiency cannot be calculated. A more complex cost equation must be assumed for the product gas cooler and steam drum, provided that the investment cost data supplied for cases 1A, 1 and 1B are accurate.

TABLE 4

VARIATION OF THERMOECONOMIC VARIABLES FOR THE PRODUCT GAS
COOLER AND STEAM DRUM IN CASES 1A, 1 AND 1B

VARIABLE	CASE		
	1A	1	1B
Exergy Destruction, $\dot{E}_D$ [MW]	19.67	23.92	31.67
Dimensionless Ratio, r [%]	1.71	2.03	2.61
Exergetic Efficiency, ζ [%]	67.3	67.0	65.8
Fuel Exergy, $\dot{E}_F$ [MW]	59.79	72.12	91.10
Product Exergy, $\dot{E}_P$ [MW]	40.10	48.19	60.41
Investment Costs, I [K\$]	15103	16625	19538
Cost per Unit of Fuel Exergy, c_F [\$/GJ]	3.273	3.270	3.303
Cost per Unit of Product Exergy, c_P [\$/GJ]	7.691	7.468	7.448
Relative Cost Difference, d [%]	135.0	128.4	125.5
Cost Rate of Exergy Destruction, $\dot{D}$ [\$/hr]	232	282	316
Thermoeconomic Factor, f [%]	63.63	61.34	58.24

The values of cost per unit of product exergy suggest that if (a) the product gas cooler and the HRSD are considered isolated, and (b) the gasifier operating temperature is the only decision variable, then the cost performance of these components is best around $T_0 = 1900^\circ\text{F}$.

The relative cost difference d between cost of product and cost of fuel is relatively high. As the thermoeconomic factor f indicates, approximately 56% to 64% of this difference is caused by the investment costs. This is a relatively high number for a heat exchanger. At lower plant capacity factors (below 65%) elimination of these two components might be cost effective.

5a. Mixing of Steam and Gas from the Product Gas Cooler

The exergy destruction in the mixing process of steam (stream 39) with the gas exiting the product gas cooler (stream 12) is considerable, 17.0 MW for all cases. In a new design developed for case 1, every effort should be made to reduce the irreversibilities at this point by reducing the mass flow rate of steam to the absolute minimum required for the operation of the zinc ferrite unit and/or by cooling the gas to a lower temperature before the gas is mixed with the steam.

5. Cyclones and Filters (421L/F; 422L/F; 430L/F/V; 431L)

These components together with the chloride guards which are discussed below form the area 400-gas conditioning. The exergy destruction in these components excluding the mixing process discussed above is 2.2 MW, 2.3 MW and 2.4 MW for cases 1A, 1 and 1B respectively. When the mixing of steam and gas from the product gas cooler is considered together with the cyclones and filters, the high values of exergy destruction, relative cost difference d (a unit of product exergy costs 20%-21% more than a unit of fuel exergy) and cost of exergy destruction (\$333-\$340 per hour of plant operation) indicate that more attention should be devoted to this area in future studies.

6. Area 600 -- Sulfation

The exergy destruction in the area 600 is 14.3 MW, 15.1 MW and

16.2 MW for cases 1A, 1 and 1B, respectively. The exergetic efficiency for each of the three cases is approximately 55%. The calculated economically optimal efficiency is above 80%. Thus, it might be cost effective to increase the capital expenditures for this area, if this would lead to an increase in the efficiency of this area. One option to accomplish that might be to reduce the temperature at which the sulfator flue gas leaves this area by generating more steam in the sulfator and then mixing this flue gas with the gas entering the HRSG. It does not seem cost effective first to dry the coal in coal preparation and then to add high-pressure steam to the gas exiting the product gas cooler.

8. Chloride Guard Reactors (410D, DA)

The exergy destruction in the chloride guards is of the order of 1.5 MW. The cost per product exergy unit increases by approximately 13% compared to the cost per fuel exergy unit and the cost of exergy destruction is approximately \$29 per hour of plant operation.

9-12. Area 250 -- Booster Air Compression

The exergy destruction rate in the air recuperator (#9; 251C), BFW preheater (#10; 252CA), trim cooler (#11; 252CB) and booster air compressor (#12; 251J) is approximately 2.0 MW, 0.7 MW, 2.2 MW and 4.5 MW, respectively, and the cost associated with the exergy destruction about \$55/hr., \$17/hr., \$63/hr. and \$139/hr.,

respectively. All components in this area operate near their economically optimal exergetic efficiency values.

14. Zinc Ferrite Reactors (510D,DA)

These reactors are responsible for 8.7 MW of exergy destruction and for an associated cost rate of \$119/hr. The capital expenditures cause approximately 65% (f value) of the relative cost difference $d = 2.5\%$. No specific recommendations based on the thermoeconomic analysis can be made for the zinc ferrite reactors. The dimensionless ratio r for the zinc ferrite reactors and the chloride guard reactors, which is 0.86% is lower than the ratio r for the acid gas removal and sulfur recovery units in the reference case, which is 1.06%.

16. Exit Gas Cooler (510C)

The exergy destruction in the exit gas cooler is about 7 MW and the associated cost \$193/hr. The actual exergetic efficiency of 73.9% is only slightly higher than the economically optimal efficiency of 70-71% at a plant capacity factor of 65%.

18. Recycle Gas Cooler (361C)

The recycle gas cooler operates at an exergetic efficiency of about 70% in cases 1A and 1, and 77% in case 1B. The exergy rate of the product is 3.7 MW, 3.6 MW and 3.9 MW for cases 1A, 1 and 1B, respectively, whereas the corresponding investment costs are 3.23, 3.19 and 3.23 million dollars. These numbers indicate that the

cost data are not consistent with the thermodynamic data because case 1B has approximately the same investment costs with case 1A although both the exergetic efficiency and the mass flow rates in case 1B are higher than in case 1A for the recycle gas cooler. Thus, the economically optimal exergetic efficiency could not be calculated for this component.

The high values of the relative cost difference d (126.7%) and the thermoeconomic factor f (66.2%) lead to the conclusion that temperature differences in this cooler higher than the actual ones might be economically preferable since they will lead to a reduction in the investment costs of the recycle gas cooler. This conclusion, however, is drawn when the cooler is considered separately from the other heat exchangers in the total plant.

The analysis of the heat exchanger network of the total plant based on the "pinch method" [5] (see composite curves in Figure 2) shows that, in the recycle gas cooler, some heat transfer occurs through the pinch, which is at $T_{hot} = 520K$ and $T_{cold} = 495K$. This ineffectiveness increases both the energy and capital costs and must be corrected in future IGCC designs. The lower part of the heat transfer in the recycle gas cooler must be used as an IP economizer to preheat the liquid water entering the IP drum up to the saturation point.

19&21. Recycle Gas Compressor (381J)

The exergy destruction in the recycle gas compressor is 0.5 MW and the associated cost \$6.80/hr. The large values of the relative

Figure 2. Composite Curves for the Heat Exchanger Network of the IGCC Plant in Case 1

cost difference d and the thermoeconomic factor f are caused by the investment costs. These high values are typical for compressors and pumps.

22&23. Area 900 -- Gas Turbine System

This area is responsible for the highest exergy destruction rate among all IGCC plant areas. A total of 275.2 MW of exergy (i.e., 23.4% of the coal exergy) are destroyed in this system. The estimated exergy destruction in the combustion chamber of 192.6 MW (16.4% of the coal exergy) represents the second highest value of exergy destruction in a single plant component. The cost rate associated with the exergy destruction is \$3,134/hr (the highest among all plant components) for the combustion chamber and \$1,306/hr for the system air compressor/gas turbine. These numbers combined with the extremely low contribution of the investment costs to the d value of the combustion chamber indicate that the cost effectiveness of the total plant might be improved significantly if a gas turbine process with a reheat stage is used. There is an urgent need to study the economic feasibility of mature gas turbine system designs including a reheat stage.

Area 1000 -- Heat Recovery Steam Generator

The exergy destruction rate in the total HRSG varies from 35.8 MW in case 1A to 35.6 MW in case 1B. The corresponding exergetic efficiency is of the order of 84.6%. Significant exergy destruction occurs in the superheater/reheater/blast steam reheater

(12.9 MW with an associated cost rate of \$245/hr), the feedwater heater #1 (4.2 MW with an associated cost rate of \$80/hr), the HP1 economizer (4.0 MW with an associated cost rate of \$76/hr), the feedwater heater #2 (3.5 MW and \$64/hr), the HP drum (3.4 MW and \$65/hr) and the IP drum (2.9 MW and \$72/hr).

In the IP drum, 1.3 MW of heat is transferred through the heat exchanger network pinch (495 K/520 K) which is shown in Figure 2. This inefficiency must be corrected in future design configurations of the HRSG.

Figures 3 and 4 show the composite curves and the temperature-heat transfer profile for the HRSG. The minimum temperature differences occur in the HP2 economizer, IP superheater and HP drum. Estimates based on the values of the HRSG thermoeconomic variables and on a comparison of Figures 3 and 4 indicate that it might be cost effective to increase the temperature differences in the HP2 economizer and IP superheater and decrease the minimum temperature difference in the HP drum. At this point it must be emphasized that the HRSG should not be optimized by itself. An optimization study must simultaneously consider the HRSG, all heat recovery devices in the gasification island and the steam turbine performance. The temperature, pressure and mass flow rate values of the main steam and the reheat steam must be kept variable in such an optimization study. Our studies indicate that a good match of temperature profiles in the HRSG has the potential to increase the overall plant efficiency significantly.

Figure 3. Composite Curves for the HRSG in Case 1

Figure 4. Q-T Diagram for the HRSG in Case 1

Area 1100 Steam Cycle

The total exergy destruction in the steam cycle (including the exergy losses in the condenser) is 48.0 MW, 49.8 MW and 52.7 MW for cases 1A, 1 and 1B, respectively. The actual steam cycle design considered by itself is cost efficient. The cost optimal design for this area, however, can be obtained through an integrated optimization of the overall heat exchanger network and the steam turbines. It is apparent that the exergy destruction in the desuperheater can and must be eliminated.

F-3 COMPARISON BETWEEN CASE 1 AND REFERENCE CASE

The major differences between case 1 and reference case lie in the following

- Oxidant feed
 - The air feed to the oxidant feed system is taken from the gas turbine system in case 1.
- Coal gasifier
 - An oxygen-blown gasifier is used in the reference case as opposed to an air-blown gasifier in case 1.
 - The coal moisture at the gasifier inlet is 11.12 wt% in the reference case versus 4.98 wt% in case 1.
 - The gasification temperature is 1900F in case 1 versus 1850F in the reference case.
- Solids to be disposed

Stream No. 16 (from gas filter to fines handling) in the reference case contains 4.11% of the carbon supplied to the gasifier with the coal whereas stream No. 24 (solids to disposal) in case 1 contains less than 0.07% of the carbon fed with the coal into the gasifier.
- Process used for gas cleaning

Case 1 uses a hot gas cleanup process whereas the reference case uses commercially available cold gas cleanup processes. The temperature of the fuel gas

at the inlet to the combustion chamber is, therefore, 534°F higher in case 1 than in the reference case.

- By-products

Case 1 has no useful by-products as opposed to the reference case where almost pure nitrogen and elemental sulfur are generated.

The comparison between the reference case and case 1 is conducted with the aid of (a) the dimensionless ratio r , the exergy destruction rate in a plant component divided by the coal exergy rate supplied to the corresponding power plant, and (b) the dimensionless ratio between the exergy loss (the exergy rate of a stream leaving the total plant) and the exergy rate of the coal stream supplied to the corresponding power plant. From the thermodynamic viewpoint, these dimensionless variables represent objective measures of the component performance and the value of a stream, respectively. These variables allow for the comparison of power plants with different capacities, different design configurations and different coal moisture content. A comparison of the values of these dimensionless variables is shown in Table 5.

The oxygen-blown KRW gasifier has a considerably higher thermodynamic efficiency than the air-blown KRW gasifier. This difference is caused by the heating of nitrogen up to the gasifier operating temperature, and, mainly, by the differences in the reactions occurring in the gasifier. The product gas from the

TABLE 5

Comparison of the values of exergy destruction in selected groups of plant components and of the exergy loss associated with streams leaving the plant in the reference case and case 1. The values in the first two columns are given as a percentage of the coal exergy supplied to the corresponding IGCC power plant.

VARIABLE	REFERENCE CASE	CASE 1	DIFFERENCE
I. Exergy Destruction			
Gasification System (incl. Oxidant Feed System)	13.72	19.27	-5.55
Gas Turbine System	27.59	23.38	+4.21
High-Temperature Gas Cooling	2.71	2.03	+0.68
Gas Cleaning and Low-Temperature Gas Cooling	2.15	3.28	-1.13
Heat Recovery Steam Generator	3.47	2.99	+0.48
Steam Turbines	2.51	2.55	-0.04
Remaining Steam Cycle	1.31	1.67	-0.36
Other Components	1.52	1.33	+0.19
Sum I	54.98	56.50	-1.52
II. Exergy Loss			
Solids to Disposal	3.70	0.07	3.63
Flue Gas	2.36	2.83	-0.47
Sulfur	1.39	---	1.39
Other Exergy Losses	0.76	1.90	1.14
Sum II	8.21	4.80	3.41
Sum of Total Exergy Destruction and Exergy Loss	63.19	61.30	1.89

gasifier contains relatively more carbon monoxide and hydrogen and considerably more methane in the reference case than in case 1.

From the thermodynamic viewpoint, the lower gasifier operating temperature and the higher coal moisture in the reference case represent an advantage and disadvantage, respectively, compared with case 1.

The thermodynamic advantage of the oxygen-blown KRW gasifier is, however, partially offset in the gas turbine system. In the reference case a larger air flow rate than in case 1 must be compressed in the air compressor and, more importantly, heated in the combustion chamber. The values shown in Table 5 and the exergetic efficiencies calculated for various groups of plant components allow for the overall effect of using oxygen-blown KRW gasifiers to be estimated. Our calculations indicate that the difference in the overall efficiency between two IGCC power plants which have been designed under identical technical approaches and assumptions with the only exception that the first plant uses an oxygen-blown KRW gasifier and the second plant an air-blown KRW gasifier, should be of the order of 0.2 to 0.8 percentage points, with the first plant having a higher overall thermodynamic efficiency.

In order to determine if IGCC plants based on an oxygen-blown KRW gasifier are more cost effective than plants based on an air-blown KRW gasifier, the relative fuel savings resulting from the efficiency difference must be compared with the relative increase

in the overall investment costs for the former plants compared with the latter ones.

The figures shown in Table 5 and additional estimates used to compare the two cases indicate that the overall plant efficiency improvement that can be attributed to the high-temperature gas cleanup process used in case 1 compared with the low-temperature gas cleanup process of the reference case is probably of the order of 0.4-0.8 percentage points. The relatively small value of the total efficiency improvement is attributed to the relatively high irreversibilities in the areas of gas conditioning and external desulfurization for case 1. As shown in Table 5, if only gas cleaning and low-temperature gas cooling are compared, the exergy destruction in case 1 is considerably higher than in the reference case.

The exergy losses, associated with material streams exiting the boundaries of the total plants considered here, are mainly responsible for the difference in the overall efficiency between reference case and case 1. In the reference case approximately 3.7% of the total exergy supplied to the plant by the coal is rejected to the environment with the solids-to-disposal stream as opposed to 0.07% in case 1. This loss should be reduced in future design options based on the oxygen-blown KRW gasifier.

The comparison of the values given for the HRSG in Table 5 shows that the HRSG design in case 1 is thermodynamically more efficient than in the reference case. This is due in part to the

fact that the temperature profiles in the HRSG can be better matched in case 1 than in the reference case.

Both design options analyzed here can be improved from the thermodynamic and cost viewpoints. The potential for improvement is considerably higher for the reference case than for case 1.

F-4 CONCLUSIONS

The exergy analysis and the thermoeconomic analysis applied here identify and evaluate the real energy waste and the real cost sources in an IGCC design option. They are very useful tools in evaluating the potential for improving the plant efficiency and cost effectiveness. With the aid of these analyses, conclusions can be drawn without redesigning the total plant.

This study shows that the cost effectiveness of case 1 might be improved if the following design changes were made.

1. Operate the gasifier at a temperature ranging from 1850° F to 1870° F.
2. Cool the gas at the outlet of the product gas cooler to a lower temperature before steam is added to the gas. Keep the steam amount to the absolute minimum required by the zinc ferrite unit.
3. Redesign the recycle gas cooler, IP economizer, and IP drum to avoid heat transfer through the pinch. Optimize the heat exchanger network in the total plant in conjunction with the steam turbines. This optimization might result in a reduction of both capital expenditures and energy costs. In addition, a reduction in the overall pinch temperature from about 34° C (61° F) in the current design to about 25°-30° C (45°-54° F) might prove to be cost effective.
4. Study the economic feasibility of a gas turbine system with a reheat stage. The thermoeconomic analysis

indicates that this change might result in considerable cost savings.

5. Enhance the heat recovery operations in the sulfation area at the expense of coal drying. Compare the cost effectiveness of coal drying versus steam raising.

The gasification island produces two products: The steam, which is used in the steam turbines, and the fuel gas, which is used in the gas turbine and in the heat recovery steam generator. The ratio of cost per exergy unit between these two energy carriers can be calculated for the gasification island and, assuming the same contribution to the electricity cost, for the power island. This ratio is approximately 2.1 in the gasification island and 3.7 in the power island. In an ideally optimized IGCC design, the difference between these two values should be smaller. This is an additional indication that the overall design of case 1 is not optimal.

The comparison of the thermodynamic performance between case 1 and the reference case shows that the oxygen-blown KRW gasifier has a higher overall efficiency than the air-blown KRW gasifier. Part of these thermodynamic gains in the gasifier is offset in the gas turbine system. Also, an IGCC plant based on hot gas cleanup is thermodynamically more efficient than a similar plant which uses cold gas cleanup even when the irreversibilities associated with the hot gas cleanup are significant, as it happens in case 1.

The exergy losses in the reference case have a considerable impact on the overall efficiency. The efficiency of this case can

be significantly improved by reducing the losses associated with the solids-to-disposal stream and optimizing the heat exchanger network.

From the above discussion it is apparent that an IGCC power plant which uses

- a. an oxygen-blown, KRW fluidized-bed gasifier employing in-bed desulfurization,
- b. an optimized hot gas cleanup process,
- c. a device to (a) burn any residual carbon remaining in the char fines, and (b) use the generated heat to raise HP steam, and
- d. an optimized heat exchanger network integrated with the steam cycle

should demonstrate significant performance and cost improvements compared with the two design options discussed in this study.

F-5 REFERENCES

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4. Tsatsaronis, G., Thermoeconomic Analysis and Optimization of Energy Systems, to be published in Progress in Energy and Combustion Science, 1991.
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APPENDIX F1

Appendix F1 contains the detailed results of the thermodynamic analysis of case 1 (Tables 1 through 3) and reference case (Tables 9 through 11), as well as the results of the detailed thermoeconomic analysis of case 1 for a capacity factor of 65% (Tables 4 through 8). Figures 1 through 4 refer to case 1, and Figures 5 through 7 to the reference case.

Figure 2. Simplified Flow Diagram of the Gasification Island in Case 1

Figure 3. Simplified Flow Diagram of the Gas Turbine System and the HRSG in Case 1

Figure 4. Simplified Flow Diagram of the Steam Cycle in Case 1

TABLE 1
THERMODYNAMIC PROPERTIES

STREAM NO.	m [kg/s]	T [K]	P [bar]	H [MW]	S [kW/K]	E [MW]
1	39.812	366.5	1.01	-97.135	69.202	1177.128
2	10.800	305.4	1.01	-130.665	9.669	0.000
3	9.830	533.2	29.30	-101.104	12.655	7.401
4	135.317	677.6	13.79	53.423	942.531	49.432
5	134.416	616.5	31.03	44.307	891.420	54.035
6	238.285	1310.9	27.58	-451.755	1826.201	1769.482
7	192.858	1310.9	27.23	-223.372	1680.791	1051.453
8	2.015	496.6	29.65	-9.343	16.046	7.398
9	1.899	496.6	29.65	-8.806	15.124	6.973
10	1.297	783.2	31.03	-16.260	13.971	1.654
11	6.305	496.6	29.65	-29.235	50.208	23.149
12	192.858	946.5	26.54	-322.468	1593.967	978.871
13	1.123	591.5	110.32	-16.348	7.811	0.466
14	103.398	591.4	110.32	-1505.143	719.026	42.872
15	2.059	546.9	51.71	-9.402	16.269	7.744
16	112.319	591.4	110.32	-1493.753	1019.746	114.927
17	6.818	922.0	29.65	-72.094	72.246	10.187
18	13.162	438.7	20.34	-62.137	104.258	47.381
19	1.598	533.2	31.03	0.397	10.320	0.620
20	13.162	922.0	21.03	-52.560	118.745	52.534
21	24.651	344.3	1.38	1.151	170.618	0.756
22	11.801	533.2	1.01	-112.543	15.995	42.841
23	23.787	1144.3	1.38	-14.276	193.290	13.004
24	12.671	338.7	1.01	-143.638	10.766	0.871
25	1.773	816.1	1.01	-9.421	4.297	32.402
26	243.921	811.5	23.79	-974.731	2170.392	984.306
27	243.852	922.0	21.03	-973.750	2199.917	973.275
29	0.900	616.5	31.03	0.297	5.972	0.362
30	5.846	783.2	31.03	-73.266	62.953	7.452
31	32.377	1283.2	29.65	-114.415	118.240	709.302
32	230.690	810.9	20.34	-961.085	2037.833	894.079
33	32.366	591.4	110.32	-430.437	293.848	33.117
34	32.366	591.4	110.32	-471.138	225.069	13.420
35	243.905	811.5	25.16	-974.605	2165.280	985.796
36	10.044	390.2	113.76	-155.781	50.292	0.558
37	10.027	529.8	1.01	-102.293	12.921	10.895
39	49.102	533.2	31.03	-644.670	483.259	47.158
40	240.187	816.1	26.54	-957.717	2128.582	976.633
42	11.103	496.6	31.03	-51.489	88.237	40.823
43	1.900	546.5	51.71	-8.676	15.011	7.149
44	0.159	546.5	51.71	-0.726	1.256	0.598
45	1.124	366.5	1.01	0.080	7.951	0.026
46	19.511	390.2	113.76	-302.614	97.695	1.084
47	19.317	591.4	110.32	-256.907	175.383	19.766
48	23.787	1033.2	1.01	-17.388	192.489	10.137
49	31.509	390.2	113.76	-488.721	157.778	1.751

TABLE 1 (CONT'D)
THERMODYNAMIC PROPERTIES

STREAM NO.	m [kg/s]	T [K]	P [bar]	H [MW]	S [kW/K]	E [MW]
50	31.509	493.1	110.32	-474.635	189.798	6.060
51	1.316	591.4	110.32	-19.160	9.154	0.546
52	61.064	390.2	113.76	-947.116	305.765	3.394
69	79.953	591.4	110.32	-1163.860	555.991	33.151
70	79.953	591.4	110.32	-1063.316	725.898	81.810
71	10.044	591.4	110.32	-146.204	69.843	4.164
72	753.316	305.4	1.01	5.561	5189.324	0.995
73	574.975	677.6	13.79	226.999	4004.920	210.041
74	805.824	1497.0	13.38	-734.812	6687.556	912.140
75	848.848	865.4	1.03	-1358.385	7078.956	259.368
76	848.848	676.3	1.03	-1549.044	6831.031	144.418
77	848.848	609.5	1.03	-1614.839	6729.104	109.749
78	848.848	580.1	1.03	-1643.536	6681.843	95.484
79	848.848	524.6	1.02	-1697.232	6585.048	71.346
80	848.848	519.7	1.02	-1701.937	6576.535	69.241
81	848.848	466.2	1.02	-1753.107	6473.134	49.648
82	848.848	428.9	1.02	-1788.363	6394.814	38.308
83	848.848	409.8	1.02	-1806.386	6352.327	33.259
84	230.690	922.0	21.03	-921.190	2081.172	920.740
85	135.317	616.5	31.03	44.604	897.392	54.397
86	13.162	496.6	31.03	-61.044	104.598	48.370
87	2.059	496.6	31.03	-9.550	16.363	7.567
88	0.193	591.4	110.32	-2.812	1.343	0.080
89	0.885	496.6	29.65	-4.107	7.046	3.236
90	848.848	605.8	1.03	-1618.433	6723.688	107.809
101	169.372	810.9	99.97	-2123.917	1734.522	241.821
102	0.072	810.9	99.97	-0.897	0.733	0.102
103	0.291	810.9	99.97	-3.655	2.985	0.416
104	116.578	613.5	24.13	-1504.016	1206.414	120.471
105	129.252	810.9	22.68	-1610.689	1421.555	164.772
106	131.443	701.2	10.34	-1667.483	1453.572	135.644
107	0.213	613.5	24.13	-2.742	2.200	0.220
108	131.655	701.0	10.34	-1670.225	1455.857	135.837
109	0.081	701.2	10.34	-1.025	0.893	0.083
110	0.127	495.2	24.13	-1.907	0.768	0.024
111	0.152	733.3	10.34	-1.922	1.699	0.163
112	1.694	591.4	110.32	-24.655	11.778	0.702
113	50.239	659.5	34.47	-643.870	518.783	56.548
114	13.928	322.3	1.72	-220.118	58.688	0.063
115	131.655	322.4	0.12	-1769.944	1518.642	16.945
116	58.066	299.8	1.00	-923.121	227.142	0.161
117	189.874	322.4	0.12	-2695.440	1747.164	16.913
118	189.874	322.3	0.12	-3000.763	800.008	0.826
119	0.152	372.8	1.01	-2.375	0.735	0.005
120	175.946	322.3	1.72	-2780.601	741.361	0.795

TABLE 1 (CONT'D)
THERMODYNAMIC PROPERTIES

STREAM NO.	m [kg/s]	T [K]	P [bar]	H [MW]	S [kW/K]	E [MW]
121	175.946	323.0	1.72	-2780.084	742.963	0.823
122	175.946	347.1	1.72	-2762.328	795.984	2.388
123	13.928	388.8	1.72	-185.383	149.007	7.217
124	189.874	388.7	1.72	-2947.751	949.459	8.200
125	110.002	390.6	113.76	-1706.005	551.232	6.149
126	7.144	659.5	34.47	-91.555	73.768	8.041
127	43.096	659.5	34.47	-552.315	445.015	48.507
128	7.144	783.2	31.03	-89.526	76.924	9.106
129	12.801	389.1	24.13	-198.695	64.046	0.586
130	12.801	472.6	24.13	-194.059	74.829	1.929
131	1.980	613.5	24.13	-25.539	20.486	2.046
132	12.675	495.2	24.13	-167.420	124.064	11.368
133	129.252	610.9	22.70	-1667.895	1340.463	132.331
134	141.512	570.9	110.32	-2077.324	954.333	50.387
135	169.372	591.4	110.32	-2252.524	1537.738	173.306
136	12.675	599.8	22.70	-163.879	130.914	12.817
137	177.073	390.6	113.76	-2746.188	887.329	9.898
138	169.009	806.7	89.98	-2119.364	1738.495	238.955
139	118.770	613.5	24.13	-1532.297	1229.099	122.737
140	131.523	807.9	22.68	-1639.883	1445.428	167.111
141	131.523	701.2	10.34	-1668.507	1454.465	135.727
142	131.636	591.4	110.32	-1750.660	1195.129	134.693
143	0.474	533.2	31.03	0.118	3.061	0.184
144	189.874	322.3	1.72	-3000.719	800.049	0.858
145	110.002	495.9	112.06	-1655.592	665.381	21.704
146	141.512	495.3	110.32	-2130.226	855.226	27.750
147	12.801	388.7	1.72	-198.738	64.013	0.553
148	177.073	388.7	1.72	-2749.013	885.446	7.648
149	6.007	390.6	113.76	-93.157	30.100	0.336
150	135.317	504.3	13.35	28.617	901.539	37.144
151	135.317	403.7	12.92	14.530	871.655	32.183
152	135.317	322.4	12.51	3.325	841.933	30.054
153	135.317	441.5	31.23	19.799	849.842	44.112
154	169.009	810.9	99.97	-2119.365	1730.805	241.303

TABLE 2

THERMAL, MECHANICAL, REACTION, ENVIRONMENTAL, CHEMICAL,
PHYSICAL AND TOTAL EXERGY VALUES

STREAM NO.	ETH [MW]	EM [MW]	ER [MW]	EE [MW]	ECH [MW]	EPH [MW]	ETOT [MW]
1	0.321	0.000	1143.309	33.498	1176.807	0.321	1177.128
2	0.000	0.000	0.000	0.000	0.000	0.000	0.000
3	0.555	0.000	6.320	0.526	6.846	0.555	7.401
4	18.284	30.969	0.000	0.179	0.179	49.253	49.432
5	13.540	40.318	0.000	0.178	0.178	53.858	54.035
6	193.114	64.712	1453.664	57.992	1511.656	257.826	1769.482
7	146.509	64.465	830.515	9.964	840.479	210.974	1051.453
8	0.148	0.752	6.453	0.045	6.498	0.900	7.398
9	0.140	0.708	6.082	0.042	6.125	0.848	6.973
10	1.054	0.596	0.000	0.003	0.003	1.650	1.654
11	0.465	2.352	20.193	0.140	20.333	2.816	23.149
12	74.429	63.963	830.515	9.964	840.479	138.392	978.871
13	0.451	0.012	0.000	0.003	0.003	0.463	0.466
14	41.476	1.131	0.000	0.265	0.265	42.607	42.872
15	0.212	0.895	6.591	0.046	6.637	1.107	7.744
16	66.522	48.120	0.000	0.287	0.287	114.540	114.927
17	5.828	3.030	0.000	1.328	1.328	8.858	10.187
18	0.594	4.362	42.132	0.293	42.425	4.956	47.381
19	0.098	0.494	0.000	0.028	0.028	0.592	0.620
20	5.699	1.410	42.132	0.293	42.425	10.109	52.534
21	0.057	0.666	0.000	0.033	0.033	0.723	0.756
22	0.711	0.000	39.069	3.061	42.130	0.711	42.841
23	11.676	0.630	0.002	0.698	0.699	12.305	13.004
24	0.019	0.000	0.764	0.087	0.851	0.019	0.871
25	0.611	0.000	29.516	2.275	31.791	0.611	32.402
26	81.443	85.022	810.258	7.582	817.841	166.465	984.306
27	105.580	81.702	780.557	5.435	785.992	187.283	973.275
29	0.091	0.270	0.000	0.001	0.001	0.361	0.362
30	4.751	2.686	0.000	0.015	0.015	7.437	7.452
31	38.125	0.000	623.149	48.028	671.177	38.125	709.302
32	74.070	76.442	738.425	5.142	743.567	150.513	894.079
33	19.169	13.866	0.000	0.083	0.083	33.034	33.117
34	12.983	0.354	0.000	0.083	0.083	13.337	13.420
35	81.421	86.536	810.258	7.580	817.839	167.957	985.796
36	0.419	0.113	0.000	0.026	0.026	0.532	0.558
37	0.556	0.000	9.553	0.787	10.339	0.556	10.895
39	26.462	20.572	0.000	0.126	0.126	47.033	47.158
40	81.258	86.562	800.999	7.814	808.813	167.820	976.633
42	0.817	4.198	35.562	0.246	35.808	5.015	40.823
43	0.195	0.826	6.086	0.042	6.128	1.021	7.149
44	0.016	0.069	0.509	0.001	0.513	0.085	0.598
45	0.006	0.000	0.000	0.020	0.020	0.006	0.026
46	0.814	0.220	0.000	0.050	0.050	1.034	1.084
47	11.441	8.276	0.000	0.049	0.049	19.717	19.766
48	9.437	0.001	0.002	0.698	0.699	9.438	10.137
49	1.315	0.356	0.000	0.081	0.081	1.670	1.751

TABLE 2 (CONT'D)

THERMAL, MECHANICAL, REACTION, ENVIRONMENTAL, CHEMICAL,
PHYSICAL AND TOTAL EXERGY VALUES

STREAM NO.	ETH [MW]	EM [MW]	ER [MW]	EE [MW]	ECH [MW]	EPH [MW]	ETOT [MW]
50	5.635	0.345	0.000	0.081	0.081	5.979	6.060
51	0.528	0.015	0.000	0.003	0.003	0.543	0.546
52	2.548	0.689	0.000	0.156	0.156	3.237	3.394
69	32.072	0.875	0.000	0.205	0.205	32.946	33.151
70	47.353	34.254	0.000	0.205	0.205	81.605	81.810
71	4.029	0.110	0.000	0.026	0.026	4.139	4.164
72	0.000	0.000	0.000	0.995	0.995	0.000	0.995
73	77.691	131.590	0.000	0.760	0.760	209.281	210.041
74	710.280	186.517	0.506	14.837	15.343	896.797	912.140
75	242.995	1.556	0.506	14.310	14.816	244.552	259.368
76	128.198	1X1.404	0.506	14.310	14.816	129.602	144.418
77	93.681	1.252	0.506	14.310	14.816	94.933	109.749
78	79.721	0.947	0.506	14.310	14.816	80.668	95.484
79	55.735	0.795	0.506	14.310	14.816	56.530	71.346
80	53.782	0.643	0.506	14.310	14.816	54.425	69.241
81	34.341	0.490	0.506	14.310	14.816	34.832	49.648
82	23.154	0.338	0.506	14.310	14.816	23.492	38.308
83	18.257	0.186	0.506	14.310	14.816	18.443	33.259
84	99.882	77.292	738.425	5.142	743.567	177.174	920.740
85	13.630	40.588	0.000	0.179	0.179	54.219	54.397
86	0.969	4.976	42.132	0.293	42.425	5.945	48.31
87	0.152	0.778	6.591	0.046	6.637	0.930	7.567
88	0.077	0.002	0.000	0.000	0.000	0.080	0.080
89	0.065	0.330	2.822	0.019	2.841	0.395	3.236
90	91.894	1.100	0.506	14.310	14.816	92.993	107.809
101	143.875	97.517	0.000	0.433	0.433	241.388	241.821
102	0.061	0.041	0.000	0.000	0.000	0.102	0.102
103	0.248	0.168	0.000	0.001	0.001	0.415	0.416
104	71.620	48.555	0.000	0.298	0.298	120.173	120.471
105	109.795	54.650	0.000	0.331	0.331	164.442	164.772
106	93.433	41.877	0.000	0.336	0.336	135.307	135.644
107	0.131	0.089	0.000	0.001	0.001	0.219	0.220
108	93.559	41.944	0.000	0.337	0.337	135.500	135.837
109	0.057	0.026	0.000	0.000	0.000	0.083	0.083
110	0.023	0.000	0.000	0.000	0.000	0.024	0.024
111	0.114	0.049	0.000	0.000	0.000	0.163	0.163
112	0.679	0.019	0.000	0.004	0.004	0.698	0.702
113	33.318	23.103	0.000	0.129	0.129	56.420	56.548
114	0.026	0.001	0.000	0.036	0.036	0.027	0.063
115	55.478	-38.868	0.000	0.337	0.337	16.608	16.945
116	0.012	0.000	0.000	0.149	0.149	0.012	0.161
117	54.623	-38.193	0.000	0.486	0.486	16.428	16.913
118	0.357	-0.017	0.000	0.486	0.486	0.340	0.826
119	0.004	0.000	0.000	0.000	0.000	0.004	0.005
120	0.332	0.012	0.000	0.450	0.450	0.345	0.

TABLE 2 (CONT'D)

THERMAL, MECHANICAL, REACTION, ENVIRONMENTAL, CHEMICAL,
PHYSICAL AND TOTAL EXERGY VALUES

STREAM NO.	ETH [MW]	EM [MW]	ER [MW]	EE [MW]	ECH [MW]	EPH [MW]	ETOT [MW]
121	0.360	0.012	0.000	0.450	0.450	0.372	0.823
122	1.925	0.012	0.000	0.450	0.450	1.938	2.388
123	6.180	1.002	0.000	0.036	0.036	7.181	7.217
124	7.701	0.013	0.000	0.486	0.486	7.714	8.200
125	4.626	1.241	0.000	0.281	0.281	5.867	6.149
126	4.738	3.285	0.000	0.018	0.018	8.023	8.041
127	28.580	19.818	0.000	0.110	0.110	48.397	48.507
128	5.805	3.282	0.000	0.018	0.018	9.087	9.106
129	0.523	0.030	0.000	0.033	0.033	0.553	0.586
130	1.866	0.030	0.000	0.033	0.033	1.896	1.929
131	1.216	0.825	0.000	0.005	0.005	2.041	2.046
132	6.443	4.892	0.000	0.032	0.032	11.335	11.368
133	79.063	52.939	0.000	0.331	0.331	132.000	132.331
134	48.478	1.548	0.000	0.362	0.362	50.025	50.387
135	100.312	72.564	0.000	0.433	0.433	172.872	173.306
136	7.612	5.173	0.000	0.032	0.032	12.784	12.817
137	7.447	1.998	0.000	0.453	0.453	9.445	9.898
138	142.597	95.929	0.000	0.432	0.432	238.522	238.955
139	72.966	49.469	0.000	0.304	0.304	122.433	122.737
140	111.183	55.594	0.000	0.336	0.336	166.775	167.111
141	93.490	41.903	0.000	0.336	0.336	135.391	135.727
142	77.962	56.396	0.000	0.337	0.337	134.356	134.693
143	0.029	0.147	0.000	0.008	0.008	0.176	0.184
144	0.359	0.013	0.000	0.486	0.486	0.372	0.858
145	20.200	1.223	0.000	0.281	0.281	21.423	21.704
146	25.840	1.548	0.000	0.362	0.362	27.387	27.750
147	0.519	0.001	0.000	0.033	0.033	0.520	0.553
148	7.182	0.013	0.000	0.453	0.453	7.195	7.648
149	0.253	0.068	0.000	0.015	0.015	0.320	0.336
150	6.382	30.583	0.000	0.179	0.179	36.965	37.144
151	1.807	30.198	0.000	0.179	0.179	32.004	32.183
152	0.063	29.812	0.000	0.179	0.179	29.875	30.054
153	3.266	40.668	0.000	0.179	0.179	43.934	44.112
154	143.566	97.308	0.000	0.432	0.432	240.871	241.303

TABLE 3

EXERGY RATE OF FUEL, PRODUCT, EXERGY DESTRUCTION AND
THE EXERGETIC EFFICIENCY FOR EACH PLANT COMPONENT

DEVICE NO.	NAME	FUEL EXERGY [MW]	PRODUCT EXERGY [MW]	EXERGY DESTRUCTION [MW]	EXERGETIC EFFICIENCY [%]
1	GASIFIER	1231.087	1014.133	216.954	82.377
2	CYCLONE & MIX	221.838	220.096	1.742	99.215
3	PRODUCT GAS COOLER & DRUM	72.116	48.194	23.923	66.828
5	CYCLONE AND FILTER & MIX	194.878	175.538	19.341	90.076
6	SULFATOR	33.802	18.682	15.120	55.268
8	CHLORIDE GUARD	175.538	174.048	1.490	99.151
9	AIR RECUPERATOR	12.288	10.285	2.003	83.702
10	BFW PREHEATER	4.961	4.309	0.652	86.851
11	TRIM COOLER	32.183	30.053	2.129	0.000
12	BOOSTER AIR COMPRESSOR	17.341	14.059	3.282	81.073
14	ZINC FERRITE SYSTEM	992.120	983.461	8.658	99.127
16	EXIT GAS COOLER	26.661	19.698	6.963	73.882
18	RECYCLE GAS COOLER	5.153	3.606	1.547	69.983
19	RECYCLE GAS COMPRESSOR I	1.413	0.989	0.424	69.996
21	RECYCLE GAS COMPRESSOR II	0.301	0.177	0.124	58.771
22	GAS TURBINE AND AIR COMPR.	394.295	311.640	82.655	79.037
23	COMBUSTION CHAMBER	1104.718	912.140	192.578	82.568
24	S/H, RH AND BLAST STEAM RH	114.950	102.022	12.928	88.75
27	HP TURBINE	59.670	55.898	3.772	93.6
30	DESUPERHEATER	48.843	47.159	1.685	96.551
31	MIX BEFORE IP TURBINE	167.234	167.111	0.123	99.926
32	IP TURBINE	31.384	28.168	3.216	89.753
34	MIX BEFORE LP TURBINE	135.863	135.837	0.026	99.981
35	LP TURBINE	118.892	98.130	20.762	82.537
36	SSR	0.186	0.163	0.022	87.978
37	MIX BEFORE CONDENSER	17.111	16.913	0.197	98.848
39	PUMP AFTER CONDENSER	0.046	0.032	0.014	68.683
41	GLAND SEAL CONDENSER	0.159	0.028	0.131	17.391
42	FWH # 1	11.340	7.154	4.185	63.090
43	FWH # 2	5.049	1.565	3.483	31.007
44	DEAERATOR	9.605	8.200	1.405	85.375
45	IP PUMP	0.045	0.033	0.012	72.528
46	HP PUMP	2.974	2.250	0.724	75.668
49	HP DRUM	33.967	30.550	3.416	89.942
50	IP SUPERHEATER	1.940	1.449	0.491	74.694
51	MIX AFTER IP SUPERHEATER	133.288	132.331	0.957	99.282
52	IP DRUM	12.301	9.439	2.862	76.731
53	MIX BEFORE HP2 ECONOMIZER	27.764	27.750	0.014	99.948
54	HP2 ECONOMIZER	24.138	22.638	1.500	93.787
55	IP ECONOMIZER	2.105	1.343	0.763	63.779
56	HP1 ECONOMIZER	19.594	15.555	4.038	79.390
57	FIRST STAGE PRESSURE DROP	241.303	238.955	2.348	99.027
58	COAL FEED AIR COMPRESSOR	0.653	0.210	0.444	32.7
60	TOTAL PLANT	1177.128	458.413	718.715	38.9

TABLE 4

COSTS ASSOCIATED WITH THERMAL, MECHANICAL AND REACTION EXERGY

NO.	THERMAL EXERGY MONETARY FLOW [\$/hr]	COST OF THERMAL EXERGY UNIT [\$/GJ]	MECHANICAL EXERGY MONETARY FLOW [\$/hr]	COST OF MECHANICAL EXERGY UNIT [\$/GJ]	REACTION EXERGY MONETARY FLOW [\$/hr]	COST OF REACTION EXERGY UNIT [\$/GJ]
1	0.000	0.000	0.000	0.000	8872.565	2.156
2	0.000	0.000	0.000	0.000	0.000	0.000
3	0.000	0.000	0.000	0.000	0.000	0.000
4	509.067	7.734	862.243	7.734	0.000	0.000
5	546.636	11.215	1340.014	9.232	0.000	0.000
6	2184.494	3.142	758.297	3.255	14438.132	2.759
7	1725.481	3.271	759.924	3.274	9658.093	3.230
8	12.188	22.799	26.826	9.914	76.072	3.274
9	11.491	22.799	25.284	9.914	71.700	3.274
10	29.901	7.877	18.079	8.425	0.000	0.000
11	38.140	22.799	83.939	9.914	238.032	3.274
12	876.576	3.271	759.924	3.300	9658.093	3.230
13	0.000	0.000	0.000	0.000	0.000	0.000
14	1278.782	8.564	64.639	15.871	0.000	0.000
15	27.290	35.707	56.205	17.452	77.695	3.274
16	2212.461	9.239	1579.909	9.120	0.000	0.000
17	166.641	7.942	51.598	4.730	0.000	3.274
18	16.992	7.942	75.091	4.782	496.656	3.274
19	0.000	0.000	0.000	0.000	0.000	0.000
20	162.944	7.942	75.091	4.730	496.656	3.274
21	0.000	0.000	0.000	0.000	0.000	0.000
22	4.807	1.878	0.000	3.274	303.189	2.156
24	0.097	1.391	0.000	3.274	5.930	2.156
26	2382.547	8.126	1443.323	4.716	9551.396	3.274
27	3018.771	7.942	1391.173	4.730	9201.272	3.274
29	3.662	11.215	8.977	9.232	0.000	0.000
30	134.731	7.877	81.466	8.425	0.000	0.000
31	446.753	3.255	3.868	3.255	4835.901	2.156
32	2117.827	7.942	1316.081	4.782	8704.617	3.274
33	1185.248	17.176	107.608	2.156	0.000	0.000
34	400.298	8.565	2.748	2.156	0.000	0.000
35	1963.620	6.699	1443.323	4.633	9551.396	3.274
36	25.481	16.889	6.277	15.388	0.000	0.000
39	849.172	8.914	604.633	8.164	0.000	0.000
43	25.118	35.707	51.871	17.452	71.739	3.274
44	2.095	35.707	4.341	17.452	6.001	3.274
45	0.000	0.000	0.000	0.000	0.000	0.000
46	49.503	16.889	12.199	15.388	0.000	0.000
47	626.055	15.200	12.199	0.409	0.000	0.000
48	0.000	0.000	0.000	0.000	0.000	0.000
49	79.945	16.889	19.700	15.388	0.000	0.000
50	243.021	11.981	19.700	15.875	0.000	0.000
51	0.000	0.000	0.000	0.000	0.000	0.000
52	154.934	16.889	38.169	15.388	0.000	0.000

TABLE 4 (CONT'D)

COSTS ASSOCIATED WITH THERMAL, MECHANICAL AND REACTION EXERGY

NO.	THERMAL EXERGY MONETARY FLOW [\$/hr]	COST OF THERMAL EXERGY UNIT [\$/GJ]	MECHANICAL EXERGY MONETARY FLOW [\$/hr]	COST OF MECHANICAL EXERGY UNIT [\$/GJ]	REACTION EXERGY MONETARY FLOW [\$/hr]	COST OF REACTION EXERGY UNIT [\$/GJ]
71	257.118	17.727	6.277	15.864	0.000	0.000
72	0.000	0.000	0.000	0.000	0.000	0.000
73	2163.079	7.734	3663.766	7.734	0.000	0.000
74	13430.247	5.252	4630.671	6.896	0.000	0.000
75	4594.728	5.252	38.640	6.897	0.000	0.000
76	2424.056	5.252	34.861	6.897	0.000	0.000
77	1771.390	5.252	34.861	7.736	0.000	0.000
78	1507.417	5.252	34.861	10.222	0.000	0.000
79	1053.884	5.252	34.861	12.181	0.000	0.000
80	1016.953	5.252	34.861	15.065	0.000	0.000
81	649.344	5.252	34.861	19.743	0.000	0.000
82	437.862	5.253	34.865	28.636	0.000	0.000
83	345.263	5.253	34.865	52.068	0.000	0.000
84	2855.825	7.942	1316.081	4.730	8704.617	3.274
85	550.298	11.215	1348.991	9.232	0.000	0.000
86	79.539	22.799	177.589	9.914	496.656	3.274
87	12.443	22.799	27.782	9.914	77.695	3.274
88	0.000	0.000	0.000	0.000	0.000	0.000
89	5.360	22.799	11.782	9.914	33.261	3.274
90	1737.589	5.252	34.861	8.807	0.000	0.000
101	4197.449	8.104	2886.508	8.222	0.000	0.000
102	1.774	8.104	1.220	8.222	0.000	0.000
103	7.224	8.104	4.967	8.222	0.000	0.000
104	2089.459	8.104	1472.217	8.422	0.000	0.000
105	3078.967	7.790	1648.503	8.379	0.000	0.000
106	2624.802	7.804	1264.308	8.386	0.000	0.000
107	3.810	8.104	2.683	8.422	0.000	0.000
108	2628.989	7.806	1266.991	8.391	0.000	0.000
109	1.613	7.804	0.776	8.386	0.000	0.000
110	0.000	0.000	0.000	0.000	0.000	0.000
111	6.396	15.558	1.995	11.405	0.000	0.000
112	0.000	0.000	0.000	0.000	0.000	0.000
113	972.036	8.104	700.476	8.422	0.000	0.000
114	0.027	0.285	0.694	192.644	0.000	0.000
115	1558.915	7.806	-1174.078	8.391	0.000	0.000
116	0.000	0.000	0.000	0.000	0.000	0.000
117	1561.999	7.943	-1174.078	8.539	0.000	0.000
118	0.000	0.000	0.000	0.000	0.000	0.000
119	0.235	15.558	0.000	11.405	0.000	0.000
120	0.341	0.285	8.600	192.644	0.000	0.000
121	13.996	10.806	8.600	192.644	0.000	0.000
122	159.641	23.033	8.600	192.644	0.000	0.000
123	278.090	12.500	0.694	0.192	0.000	0.000
125	281.262	16.889	68.766	15.388	0.000	0.000

TABLE 4 (CONT'D)

COSTS ASSOCIATED WITH THERMAL, MECHANICAL AND REACTION EXERGY

NO.	THERMAL EXERGY MONETARY FLOW [\$/hr]	COST OF THERMAL EXERGY UNIT [\$/GJ]	MECHANICAL EXERGY MONETARY FLOW [\$/hr]	COST OF MECHANICAL EXERGY UNIT [\$/GJ]	REACTION EXERGY MONETARY FLOW [\$/hr]	COST OF REACTION EXERGY UNIT [\$/GJ]
126	138.216	8.104	99.605	8.422	0.000	0.000
127	833.814	8.104	600.877	8.422	0.000	0.000
128	164.629	7.877	99.545	8.425	0.000	0.000
129	32.995	17.508	8.980	83.988	0.000	0.000
130	71.857	10.695	8.980	83.988	0.000	0.000
131	35.479	8.104	24.999	8.422	0.000	0.000
132	188.949	8.146	133.388	7.573	0.000	0.000
133	2318.310	8.145	1605.605	8.425	0.000	0.000
134	1494.649	8.564	88.465	15.871	0.000	0.000
135	3120.343	8.641	2269.535	8.688	0.000	0.000
136	228.851	8.352	133.388	7.163	0.000	0.000
137	452.751	16.889	110.691	15.388	0.000	0.000
138	4160.182	8.104	2908.595	8.422	0.000	0.000
139	2128.736	8.104	1499.911	8.422	0.000	0.000
140	3123.447	7.804	1678.469	8.386	0.000	0.000
141	2626.406	7.804	1265.099	8.386	0.000	0.000
142	2838.516	10.114	1592.107	7.842	0.000	0.000
143	59.880	573.566	0.000	0.000	0.000	0.000
144	0.368	0.285	9.293	192.644	0.000	0.000
145	688.828	9.472	68.766	15.624	0.000	0.000
146	931.938	10.018	88.465	15.871	0.000	0.000
147	31.747	16.985	0.624	192.644	0.000	0.000
148	439.146	16.985	8.669	192.644	0.000	0.000
149	15.358	16.889	3.756	15.388	0.000	0.000
150	177.687	7.734	862.261	7.832	0.000	0.000
151	50.298	7.734	862.261	7.932	0.000	0.000
152	57.127	252.282	862.261	8.034	0.000	0.000
153	200.735	17.073	1348.991	9.214	0.000	0.000
154	4188.440	8.104	2880.336	8.222	0.000	0.000

TABLE 5

COSTS ASSOCIATED WITH TOTAL, PHYSICAL, AND CHEMICAL EXERGY

NO.	TOTAL EXERGY MONETARY FLOW [\$/hr]	COST OF TOTAL EXERGY UNIT [\$/GJ]	PHYSICAL EXERGY MONETARY FLOW [\$/hr]	COST OF PHYSICAL EXERGY UNIT [\$/GJ]	CHEMICAL EXERGY MONETARY FLOW [\$/hr]	COST OF CHEMICAL EXERGY UNIT [\$/GJ]
1	8872.565	2.094	0.000	0.000	8872.565	2.094
3	0.000	0.000	0.000	0.000	0.000	0.000
4	1371.310	7.706	1371.310	7.734	0.000	0.000
5	1886.650	9.699	1886.650	9.731	0.000	0.000
6	17380.922	2.729	2942.791	3.171	14438.132	2.653
7	12143.498	3.208	2485.405	3.272	9658.093	3.192
8	115.086	4.321	39.014	12.040	76.072	3.252
9	108.474	4.321	36.774	12.040	71.700	3.252
10	47.980	8.059	47.980	8.075	0.000	0.000
11	360.111	4.321	122.079	12.040	238.032	3.252
12	11294.593	3.205	1636.500	3.285	9658.093	3.192
13	0.000	0.000	0.000	0.000	0.000	0.000
14	1343.421	8.704	1343.421	8.758	0.000	0.000
15	161.190	5.782	83.495	20.953	77.695	3.252
16	3792.370	9.166	3792.370	9.189	0.000	0.000
17	218.239	5.951	218.239	6.843	0.000	0.000
18	588.739	3.452	92.083	5.161	496.656	3.252
19	0.000	0.000	0.000	0.000	0.000	0.000
20	734.690	3.885	238.034	6.541	496.656	3.252
21	0.000	0.000	0.000	0.000	0.000	0.000
22	307.996	1.997	4.807	1.878	303.189	1.999
24	6.026	1.923	0.097	1.391	5.930	1.935
26	13377.266	3.775	3825.870	6.384	9551.396	3.244
27	13611.217	3.885	4409.944	6.541	9201.272	3.252
29	12.639	9.698	12.639	9.731	0.000	0.000
30	216.197	8.059	216.197	8.075	0.000	0.000
31	5282.655	2.069	446.753	3.255	4835.901	2.001
32	12138.525	3.771	3433.908	6.337	8704.617	3.252
33	1292.856	10.844	1292.856	10.871	0.000	0.000
34	403.046	8.343	403.046	8.394	0.000	0.000
35	12958.339	3.651	3406.943	5.635	9551.396	3.244
36	31.757	15.804	31.757	16.569	0.000	0.000
39	1453.804	8.563	1453.804	8.586	0.000	0.000
43	148.727	5.779	76.989	20.946	71.739	3.252
44	12.438	5.777	6.437	20.936	6.001	3.251
45	0.000	0.000	0.000	0.000	0.000	0.000
46	61.701	15.807	61.701	16.569	0.000	0.000
47	638.253	8.970	638.253	8.992	0.000	0.000
48	0.000	0.000	0.000	0.000	0.000	0.000
49	99.644	15.807	99.644	16.569	0.000	0.000
50	262.720	12.043	262.720	12.205	0.000	0.000
51	0.000	0.000	0.000	0.000	0.000	0.000
52	193.103	15.807	193.103	16.569	0.000	0.000

TABLE 5 (CONT'D)

COSTS ASSOCIATED WITH TOTAL, PHYSICAL, AND CHEMICAL EXERGY

NO.	TOTAL EXERGY MONETARY FLOW [\$/hr]	COST OF TOTAL EXERGY UNIT [\$/GJ]	PHYSICAL EXERGY MONETARY FLOW [\$/hr]	COST OF PHYSICAL EXERGY UNIT [\$/GJ]	CHEMICAL EXERGY MONETARY FLOW [\$/hr]	COST OF CHEMICAL EXERGY UNIT [\$/GJ]
71	263.395	17.569	263.395	17.678	0.000	0.000
72	0.000	0.000	0.000	17.678	0.000	0.000
73	5826.845	7.706	5826.845	7.734	0.000	0.000
74	18060.918	5.500	18060.918	5.594	0.000	0.000
75	4633.368	4.962	4633.368	5.263	0.000	0.000
76	2458.917	4.730	2458.917	5.270	0.000	0.000
77	1806.251	4.572	1806.251	5.285	0.000	0.000
78	1542.278	4.487	1542.278	5.311	0.000	0.000
79	1088.745	4.239	1088.745	5.350	0.000	0.000
80	1051.815	4.220	1051.815	5.368	0.000	0.000
81	684.205	3.828	684.205	5.456	0.000	0.000
82	472.726	3.428	472.726	5.590	0.000	0.000
83	380.128	3.175	380.128	5.725	0.000	0.000
84	12876.523	3.885	4171.906	6.541	8704.617	3.252
85	1899.289	9.699	1899.289	9.731	0.000	0.000
86	753.784	4.329	257.128	12.015	496.656	3.252
87	117.920	4.329	40.225	12.015	77.695	3.252
88	0.000	0.000	0.000	0.000	0.000	0.000
89	50.402	4.326	17.141	12.042	33.261	3.252
90	1772.450	4.567	1772.450	5.294	0.000	0.000
101	7083.958	8.137	7083.958	8.152	0.000	0.000
102	2.993	8.136	2.993	8.152	0.000	0.000
103	12.190	8.138	12.190	8.152	0.000	0.000
104	3561.676	8.212	3561.676	8.233	0.000	0.000
105	4727.470	7.970	4727.470	7.986	0.000	0.000
106	3889.110	7.964	3889.110	7.984	0.000	0.000
107	6.494	8.210	6.494	8.233	0.000	0.000
108	3895.980	7.967	3895.980	7.987	0.000	0.000
109	2.388	7.965	2.388	7.984	0.000	0.000
110	0.000	0.000	0.000	0.000	0.000	0.000
111	8.392	14.283	8.392	14.318	0.000	0.000
112	0.000	0.000	0.000	0.000	0.000	0.000
113	1672.512	8.216	1672.512	8.234	0.000	0.000
114	0.721	3.182	0.721	7.331	0.000	0.000
115	384.837	6.308	384.837	6.436	0.000	0.000
116	0.000	0.000	0.000	0.000	0.000	0.000
117	387.921	6.371	387.921	6.559	0.000	0.000
118	0.000	0.000	0.000	0.000	0.000	0.000
119	0.235	14.521	0.235	15.558	0.000	0.000
120	8.941	3.124	8.941	7.205	0.000	0.000
121	22.596	7.630	22.596	16.864	0.000	0.000
122	168.240	19.569	168.240	24.118	0.000	0.000
123	278.783	10.730	278.783	10.783	0.000	0.000
125	350.027	15.813	350.027	16.571	0.000	0.000

TABLE 5 (CONT'D)

COSTS ASSOCIATED WITH TOTAL, PHYSICAL, AND CHEMICAL EXERGY

NO.	TOTAL EXERGY MONETARY FLOW [\$/hr]	COST OF TOTAL EXERGY UNIT [\$/GJ]	PHYSICAL EXERGY MONETARY FLOW [\$/hr]	COST OF PHYSICAL EXERGY UNIT [\$/GJ]	CHEMICAL EXERGY MONETARY FLOW [\$/hr]	COST OF CHEMICAL EXERGY UNIT [\$/GJ]
126	237.822	8.216	237.822	8.234	0.000	0.000
127	1434.690	8.216	1434.690	8.234	0.000	0.000
128	264.174	8.059	264.174	8.075	0.000	0.000
129	41.975	19.900	41.975	21.077	0.000	0.000
130	80.837	11.642	80.837	11.843	0.000	0.000
131	60.478	8.212	60.478	8.233	0.000	0.000
132	322.337	7.877	322.337	7.899	0.000	0.000
133	3923.914	8.237	3923.914	8.257	0.000	0.000
134	1583.115	8.727	1583.115	8.790	0.000	0.000
135	5389.878	8.639	5389.878	8.661	0.000	0.000
136	362.239	7.851	362.239	7.871	0.000	0.000
137	563.442	15.813	563.442	16.571	0.000	0.000
138	7068.777	8.217	7068.777	8.232	0.000	0.000
139	3628.647	8.212	3628.647	8.233	0.000	0.000
140	4801.917	7.982	4801.917	7.998	0.000	0.000
141	3891.505	7.964	3891.505	7.984	0.000	0.000
142	4430.624	9.137	4430.624	9.160	0.000	0.000
143	59.880	90.497	59.880	94.723	0.000	0.000
144	9.661	3.128	9.661	7.212	0.000	0.000
145	757.593	9.696	757.593	9.823	0.000	0.000
146	1020.403	10.214	1020.403	10.349	0.000	0.000
147	32.371	16.263	32.371	17.289	0.000	0.000
148	447.814	16.266	447.814	17.290	0.000	0.000
149	19.114	15.816	19.114	16.571	0.000	0.000
150	1039.948	7.777	1039.948	7.815	0.000	0.000
151	912.559	7.877	912.559	7.921	0.000	0.000
152	919.388	8.498	919.388	8.549	0.000	0.000
153	1549.726	9.759	1549.726	9.798	0.000	0.000
154	7068.776	8.137	7068.776	8.152	0.000	0.000

TABLE 6

COST OF FUEL, PRODUCT AND EXERGY DESTRUCTION

DVC NO.	FUEL MONETARY FLOW [\$/hr]	COST PER FUEL EXERGY UNIT [\$/GJ]	PRODUCT MONETARY FLOW [\$/hr]	COST PER PRODUCT EXERGY UNIT [\$/GJ]	COST RATE OF EXERGY DESTRUCTION [\$/hr]
1	11328.931	2.556	11796.541	3.231	1996.493
2	2465.736	3.088	2506.352	3.163	19.361
3	848.905	3.270	1295.622	7.468	281.601
5	3152.975	4.494	3406.943	5.391	312.916
6	347.278	2.854	582.577	8.662	155.344
8	3406.943	5.391	3825.870	6.106	28.919
9	331.391	7.491	349.563	9.441	54.011
10	127.389	7.133	163.074	10.513	16.750
11	912.559	7.877	919.388	8.498	60.380
12	531.764	8.518	630.338	12.454	100.646
14	13606.102	3.809	13829.455	3.906	118.743
16	737.998	7.689	889.810	12.548	192.753
18	145.951	7.868	231.637	17.843	43.811
19	43.324	8.518	165.045	46.360	12.999
21	9.230	8.518	43.270	67.945	3.806
22	6229.850	4.389	7809.074	6.961	1305.950
23	17977.809	4.520	18060.918	5.500	3133.952
24	2174.450	5.255	2522.526	6.868	244.545
27	1767.618	8.229	2050.229	10.188	111.739
30	1453.804	8.268	1453.804	8.563	50.142
31	4800.138	7.973	4801.917	7.982	3.537
32	910.412	8.058	1071.837	10.570	93.289
34	3895.603	7.965	3895.980	7.967	0.757
35	3511.143	8.203	4211.700	11.922	613.152
36	5.382	8.059	8.392	14.283	0.647
37	385.072	6.251	387.921	6.371	4.438
39	1.420	8.518	9.661	84.392	0.445
41	8.156	14.276	13.655	137.433	6.738
42	211.567	5.183	278.063	10.797	78.090
43	92.599	5.095	145.644	25.843	63.886
44	447.024	12.928	480.186	16.266	65.375
45	1.395	8.518	9.603	80.837	0.383
46	91.192	8.518	115.627	14.274	22.189
49	652.666	5.337	719.561	6.543	65.642
50	33.801	4.840	39.902	7.649	8.554
51	3923.914	8.178	3923.914	8.237	28.179
52	311.017	7.023	322.337	9.486	72.371
53	1020.403	10.209	1020.403	10.214	0.526
54	453.533	5.219	562.711	6.905	28.179
55	36.931	4.872	38.862	8.039	13.377
56	367.610	5.212	407.566	7.278	75.765
57	7068.776	8.137	7068.777	8.217	68.789
58	1.024	0.435	59.880	79.282	0.695
60	8872.565	2.094	14779.984	8.956	5417.294

TABLE 7
VARIOUS THERMOECONOMIC VARIABLES

DVC NO.	Z [\$ / hr]	$C_P - C_F$ [\$ / GJ]	$(C_P - C_F)$ ----- C_F	$(1 - \text{EFF})$ ----- EFF	Z ----- $E_P * C_F$	f * 100
1	467.61	0.67	0.264	0.214	0.050	18.98
2	40.61	0.08	0.025	0.008	0.017	67.72
3	446.72	4.20	1.284	0.496	0.787	61.34
5	253.97	0.90	0.200	0.110	0.089	44.80
6	235.30	5.81	2.035	0.809	1.226	60.23
8	418.93	0.71	0.133	0.009	0.124	93.54
9	18.17	1.95	0.260	0.195	0.066	25.18
10	35.68	3.38	0.474	0.151	0.323	68.06
11	6.83	0.62	0.008	0.000	0.008	10.16
12	98.57	3.94	0.462	0.233	0.229	49.48
14	223.35	0.10	0.025	0.009	0.017	65.29
16	151.81	4.86	0.632	0.354	0.278	44.06
18	85.69	9.97	1.268	0.429	0.839	66.17
19	121.72	37.84	4.443	0.429	4.014	90.35
21	34.04	59.43	6.976	0.702	6.275	89.94
22	1579.23	2.57	0.586	0.265	0.321	54.74
23	83.11	0.98	0.217	0.211	0.006	2.58
24	201.92	1.61	0.231	0.127	0.105	45.23
27	241.17	1.96	0.213	0.067	0.146	68.34
30	0.00	0.30	0.036	0.036	0.000	0.00
31	0.00	0.01	0.001	0.001	0.000	0.00
32	121.53	2.51	0.263	0.114	0.149	56.57
34	0.00	0.00	0.000	0.000	0.000	0.00
35	423.38	3.72	0.358	0.212	0.146	40.85
36	2.69	6.22	0.704	0.137	0.568	80.59
37	0.00	0.12	0.012	0.012	0.000	0.00
39	8.06	75.87	8.722	0.456	8.267	94.77
41	2.69	123.16	6.644	4.750	1.894	28.51
42	13.52	5.61	0.686	0.585	0.101	14.76
43	6.54	20.75	2.453	2.225	0.228	9.29
44	13.44	3.34	0.206	0.171	0.035	17.05
45	8.06	72.32	8.345	0.379	7.966	95.46
46	16.12	5.76	0.555	0.322	0.234	42.08
49	66.89	1.21	0.226	0.112	0.114	50.43
50	6.10	2.81	0.580	0.339	0.242	41.63
51	0.00	0.06	0.007	0.007	0.000	0.00
52	11.32	2.46	0.351	0.303	0.047	13.53
53	0.00	0.01	0.001	0.001	0.000	0.00
54	109.18	1.69	0.323	0.066	0.257	79.49
55	1.93	3.17	0.650	0.568	0.082	12.62
56	39.96	2.07	0.397	0.260	0.137	34.53
57	0.00	0.08	0.010	0.010	0.000	0.00
58	58.86	78.85	181.054	2.114	178.940	98.83
60	5907.42	6.86	3.278	1.568	1.710	52.16

TABLE 8

COST OF FUEL FOR EACH AREA IN THE GCC POWER PLANT

AREA DESCRIPTION	COST OF FUEL [\$/GJ]	COST OF PRODUCT [\$/GJ]
AREA 250 : BOOSTER AIR COMPRESSION	7.917	9.758
AREA 300 : KRW GASIFICATION	2.703	3.576
AREA 380 : RECYCLE GAS COMPRESSION	4.031	5.474
AREA 400 : GAS CONDITIONING	4.444	6.106
AREA 500 : EXTERNL DESULFURIZATION	3.805	3.982
AREA 600 : SULFATOR	2.406	8.573
AREA 900 : GAS TURBINE SYSTEM	3.283	7.066
AREA 1000: HRSG	5.285	7.314
AREA 1100: STEAM TURBINE SYSTEM	7.810	11.181

Figure 5. Simplified Flow Diagram of the Gasification Island in the Reference Case

Figure 6. Simplified Flow Diagram of the Gas Turbine System and the HRSG in Reference Case

Figure 7. Simplified Flow Diagram of the Steam Cycle in Reference Case

TABLE 9
THERMODYNAMIC PROPERTIES

STREAM NO.	m [kg/s]	T [K]	P [bar]	H [MW]	S [kW/K]	E [MW]
1	40.373	305.4	33.66	-138.928	65.996	1116.574
2	3.897	557.6	32.41	-16.463	34.271	52.463
3	40.373	305.4	33.66	-138.928	65.996	1116.574
4	24.641	505.4	36.54	4.897	149.049	10.769
5	4.792	652.6	37.92	-61.527	49.124	5.398
6	11.040	557.6	32.41	-46.638	97.085	148.622
7	3.166	394.3	1.01	-36.456	3.374	3.015
8	126.000	1227.3	32.06	-409.583	956.279	2139.860
9	126.000	977.6	31.92	-474.803	897.084	2092.716
10	81.568	977.6	31.92	-287.473	778.977	1148.056
11	81.586	510.9	31.58	-350.825	691.822	1111.319
12	79.571	505.4	31.44	-341.244	687.823	1069.636
16	1.997	505.4	1.01	-10.287	2.752	41.358
17	64.634	505.4	31.44	-277.186	558.705	868.844
18	14.937	505.4	31.44	-64.058	129.118	200.792
19	62.015	313.7	31.30	-263.209	484.408	859.513
20	23.634	455.4	31.30	-357.085	134.828	6.848
21	21.015	305.4	31.58	-333.541	83.794	0.118
22	1.290	490.3	37.16	-19.458	7.756	0.234
23	59.415	299.8	31.16	-248.646	465.251	836.783
24	84.158	425.5	24.82	-589.021	708.126	845.377
25	85.020	514.3	24.13	-564.178	799.319	857.107
26	0.862	652.6	37.92	-11.066	8.835	0.971
27	825.710	1497.1	15.52	-269.708	6696.649	914.185
28	740.690	677.6	16.00	294.470	5147.803	282.245
29	1.046	520.3	37.92	-13.815	10.053	0.995
30	1.046	510.3	37.54	-15.673	6.481	0.228
31	0.383	322.0	1.52	-2.648	2.513	2.555
32	2.136	322.0	1.52	-12.003	11.913	20.138
33	0.081	299.8	7.93	-1.292	0.318	0.000
34	0.325	333.1	1.38	-2.315	2.621	3.403
35	0.818	390.0	1.01	0.090	1.068	15.468
36	4.299	305.4	1.01	0.032	29.723	0.009
37	24.641	428.2	38.37	3.038	144.745	10.225
38	5.809	319.1	1.38	-21.101	37.830	0.614
39	0.514	319.1	1.38	-8.130	2.145	0.002
40	845.586	722.4	1.03	-1043.782	6769.173	159.856
41	1.902	388.8	7.64	-29.528	9.513	0.083
42	760.566	305.4	1.01	5.654	5257.997	1.634
43	845.586	873.7	1.03	-894.705	6955.976	251.888
44	845.586	410.4	1.01	-1338.599	6240.762	26.401
45	845.586	595.2	1.03	-1166.085	6583.497	94.253
46	845.586	541.5	1.03	-1216.867	6495.065	70.475
47	845.586	523.5	1.02	-1233.750	6463.847	63.126
48	14.937	510.7	32.40	-63.934	129.179	200.898
49	44.433	977.6	31.92	-187.329	118.107	944.661

TABLE 9 (CONT'D)
THERMODYNAMIC PROPERTIES

STREAM NO.	m [kg/s]	T [K]	P [bar]	H [MW]	S [kW/K]	E [MW]
51	845.586	513.7	1.02	-1242.920	6446.655	59.206
52	845.586	410.4	1.01	-1338.599	6240.762	26.401
53	105.794	305.4	1.01	0.786	731.381	0.227
54	81.153	302.0	1.01	0.337	557.336	1.539
55	324.346	873.7	1.03	-343.186	6955.976	251.888
56	324.346	722.4	1.03	-400.369	6769.173	159.856
57	521.240	873.7	1.03	-551.518	6955.976	251.888
58	9.010	490.3	37.16	-135.854	54.152	1.633
59	521.240	722.4	1.03	-643.413	6769.173	159.856
60	7.990	383.4	1.72	-124.221	39.494	0.308
61	24.243	383.4	1.72	-376.906	119.832	0.934
62	136.282	383.5	4.53	-2118.711	673.675	5.297
63	845.586	477.5	1.02	-1276.656	6379.053	46.114
64	845.586	430.3	1.02	-1320.365	6283.660	31.535
65	76.375	486.0	106.87	-1152.942	455.032	13.734
66	76.375	532.6	105.17	-1136.309	487.723	20.384
67	24.243	470.3	37.92	-367.738	141.138	3.596
68	845.586	457.7	1.02	-1295.032	6340.241	39.589
69	130.722	794.3	103.42	-1645.196	1329.405	183.535
101	131.000	794.3	103.42	-1648.690	1332.228	183.925
102	0.055	794.3	103.42	-0.689	0.557	0.077
103	0.223	794.3	103.42	-2.805	2.267	0.313
104	107.658	595.5	24.13	-1393.439	1106.668	109.029
105	121.658	794.3	22.68	-1520.548	1332.430	152.306
106	123.333	686.6	10.34	-1568.441	1358.369	125.129
107	0.163	595.5	24.13	-2.104	1.671	0.165
108	168.515	322.4	0.12	-2377.725	1595.600	15.775
109	0.062	686.6	10.34	-0.786	0.680	0.063
110	14.000	520.3	37.92	-184.916	134.566	13.320
111	0.117	715.6	10.34	-1.474	1.293	0.122
112	845.586	592.2	1.03	-1168.955	6579.152	92.710
113	10.008	322.3	1.72	-158.161	42.169	0.045
114	158.507	322.3	1.72	-2504.998	667.886	0.716
115	123.496	322.4	0.12	-1662.067	1418.894	15.800
116	44.902	299.8	1.01	-713.842	175.643	0.125
117	168.515	322.3	0.12	-2663.198	710.019	0.733
118	168.515	322.3	1.72	-2663.159	710.055	0.761
119	0.117	372.8	1.01	-1.816	0.562	0.003
120	158.507	322.8	1.72	-2504.656	668.946	0.734
121	14.000	585.5	37.92	-182.089	139.701	14.579
122	158.507	349.9	1.72	-2486.691	722.387	2.380
123	10.008	388.8	1.72	-133.203	107.063	5.186
124	168.515	383.4	1.72	-2619.895	832.956	6.495
125	136.282	385.4	108.59	-2116.584	675.332	6.917
126	7.990	383.5	4.53	-124.218	39.497	0.311

TABLE 9 (CONT'D)
THERMODYNAMIC PROPERTIES

STREAM NO.	m [kg/s]	T [K]	P [bar]	H [MW]	S [kW/K]	E [MW]
127	59.906	385.4	108.59	-930.409	296.853	3.040
128	59.906	449.1	103.42	-914.136	335.994	7.359
129	4.493	586.6	103.42	-65.535	31.030	1.796
130	55.413	586.6	103.42	-736.274	505.394	56.685
131	1.514	595.5	24.13	-19.596	15.563	1.533
132	10.056	520.3	37.92	-132.821	96.655	9.567
133	21.388	652.9	37.92	-274.568	219.259	24.098
134	1.720	652.6	37.92	-22.088	17.635	1.938
136	0.430	652.6	37.92	-5.523	4.410	0.485
137	17.024	652.6	37.92	-218.557	174.499	19.173
138	9.010	520.3	37.92	-119.006	86.602	8.572
139	2.464	420.9	4.48	-32.695	25.575	1.624
140	0.182	420.9	4.48	-2.415	1.889	0.120
141	36.236	586.6	103.42	-528.546	250.256	14.489
142	36.236	586.6	103.42	-481.467	330.489	37.068
143	50.198	586.6	103.42	-732.202	346.683	20.071
144	50.198	586.6	103.42	-666.982	457.830	51.350
145	3.079	420.9	4.48	-40.850	31.953	2.030
146	2.464	420.9	4.48	-37.920	13.161	0.190
147	0.430	520.3	37.92	-6.426	2.706	0.102
148	95.165	355.4	30.85	-1490.549	439.748	1.999
149	95.165	355.5	40.51	-1490.423	439.838	2.098
150	95.165	372.1	39.30	-1483.819	458.026	3.148
151	95.165	389.3	38.12	-1476.897	476.244	4.507
152	95.165	430.8	36.98	-1460.048	517.388	8.792
153	7.719	490.3	37.16	-116.396	46.396	1.399
154	17.024	518.2	37.16	-254.466	106.723	3.961
155	84.158	512.9	24.82	-553.112	788.826	856.643
156	85.020	514.4	24.82	-564.178	798.308	857.416
157	17.276	455.4	31.30	-263.205	98.115	2.162
158	17.276	455.8	48.89	-263.156	98.147	2.201
159	17.276	366.5	47.91	-269.760	82.040	0.516
160	76.375	385.4	108.59	-1186.179	378.471	3.876
161	24.243	384.1	38.30	-376.773	119.936	1.036
162	7.990	383.5	4.53	-124.218	39.497	0.311
163	131.000	586.6	103.42	-1740.585	1194.774	134.005
164	121.658	592.4	24.13	-1575.528	1249.090	122.777
165	7.938	420.9	4.48	-105.310	82.375	5.232
166	3.079	390.9	4.48	-47.773	15.471	0.140
167	17.276	324.8	46.95	-272.771	73.321	0.167
169	0.117	715.6	10.34	-1.474	1.293	0.122
170	84.158	425.5	24.82	-589.021	708.127	845.377
171	0.789	586.6	103.42	-11.502	5.446	0.315
172	0.187	520.3	37.92	-2.797	1.178	0.044
173	0.052	420.9	4.48	-0.803	0.279	0.004
174	130.722	789.6	93.08	-1645.196	1335.300	181.735

TABLE 9 (CONT'D)
THERMODYNAMIC PROPERTIES

STREAM NO.	m [kg/s]	T [K]	P [bar]	H [MW]	S [kW/K]	E [MW]
175	109.335	595.5	24.13	-1415.139	1123.902	110.726
176	123.395	791.7	22.68	-1542.950	1350.579	154.055
177	123.395	686.6	10.34	-1569.226	1359.050	125.192
178	123.496	686.4	10.34	-1570.545	1360.106	125.273
179	2.394	420.9	4.48	-31.765	24.847	1.578

TABLE 10

THERMAL, MECHANICAL, REACTION, ENVIRONMENTAL,
CHEMICAL, PHYSICAL AND TOTAL EXERGY VALUES

STREAM NO.	ETH [MW]	EM [MW]	ER [MW]	EE [MW]	ECH [MW]	EPH [MW]	ETOT [MW]
1	0.015	0.000	1084.693	31.866	1116.559	0.015	1116.574
2	0.410	1.676	49.123	1.255	50.378	2.086	52.463
3	0.015	0.000	1084.693	31.866	1116.559	0.015	1116.574
4	1.102	7.054	0.000	2.614	2.614	8.156	10.769
5	3.142	2.243	0.000	0.012	0.012	5.385	5.398
6	1.161	4.747	139.159	3.554	142.714	5.908	148.622
7	0.041	0.000	0.000	2.975	2.975	0.041	3.015
8	116.937	34.089	1004.490	984.344	1988.833	151.027	2139.860
9	68.836	34.047	1004.490	984.344	1988.833	103.883	2092.716
10	42.636	34.047	1004.490	66.884	1071.373	76.683	1148.056
11	6.006	33.939	1004.490	66.884	1071.373	39.946	1111.319
12	5.592	33.896	1004.490	25.659	1030.148	39.488	1069.636
16	0.133	0.000	0.000	41.225	41.225	0.133	41.358
17	4.542	27.533	815.927	20.842	836.769	32.075	868.844
18	1.050	6.363	188.563	4.817	193.380	7.413	200.792
19	0.007	26.266	812.662	20.579	833.241	26.272	859.513
20	2.805	0.167	3.265	0.611	3.876	2.972	6.848
21	0.000	0.064	0.000	0.054	0.054	0.064	0.118
22	0.226	0.005	0.000	0.003	0.003	0.231	0.234
23	0.002	25.611	797.717	13.453	811.170	25.613	836.783
24	4.279	30.070	797.718	13.311	811.028	34.348	845.377
25	10.937	35.140	797.718	13.314	811.031	46.076	857.107
26	0.565	0.403	0.000	0.002	0.002	0.969	0.971
27	703.112	199.142	0.006	11.926	11.931	902.254	914.185
28	100.787	179.867	0.000	1.591	1.591	280.654	282.245
29	0.552	0.440	0.000	0.003	0.003	0.992	0.995
30	0.221	0.004	0.000	0.003	0.003	0.225	0.228
31	0.000	0.012	2.437	0.106	2.543	0.013	2.555
32	0.001	0.056	12.507	7.574	20.081	0.057	20.138
33	0.000	0.000	0.000	0.000	0.000	0.000	0.000
34	0.000	0.011	2.893	0.498	3.391	0.012	3.403
35	0.012	0.000	0.000	15.456	15.456	0.012	15.468
36	0.000	0.000	0.000	0.009	0.009	0.000	0.009
37	0.462	7.150	0.000	2.614	2.614	7.611	10.225
38	0.002	0.148	0.000	0.464	0.464	0.150	0.614
39	0.001	0.000	0.000	0.001	0.001	0.001	0.002
40	146.707	1.381	0.006	11.763	11.768	148.088	159.856
41	0.077	0.001	0.000	0.005	0.005	0.079	0.083
42	0.000	0.000	0.000	1.634	1.634	0.000	1.634
43	238.590	1.530	0.006	11.763	11.768	240.120	251.888
44	14.598	0.034	0.006	11.763	11.768	14.632	26.401
45	81.254	1.231	0.006	11.763	11.768	82.485	94.253
46	57.775	0.932	0.006	11.763	11.768	58.707	70.475
47	50.576	0.782	0.006	11.763	11.768	51.358	63.126
48	1.099	6.419	188.563	4.817	193.380	7.518	200.898

TABLE 10 (CONT'D)

THERMAL, MECHANICAL, REACTION, ENVIRONMENTAL,
CHEMICAL, PHYSICAL AND TOTAL EXERGY VALUES

STREAM NO.	ETH [MW]	EM [MW]	ER [MW]	EE [MW]	ECH [MW]	EPH [MW]	ETOT [MW]
49	27.200	0.000	0.000	917.461	917.461	27.200	944.661
51	46.805	0.633	0.006	11.763	11.768	47.438	59.206
52	14.598	0.034	0.006	11.763	11.768	14.632	26.401
53	0.000	0.000	0.000	0.227	0.227	0.000	0.227
54	0.002	-0.002	0.000	1.539	1.539	0.000	1.539
55	239.533	0.587	0.006	11.763	11.768	240.120	251.888
56	147.558	0.530	0.006	11.763	11.768	148.088	159.856
57	239.177	0.943	0.006	11.763	11.768	240.120	251.888
58	1.577	0.033	0.000	0.023	0.023	1.610	1.633
59	147.237	0.851	0.006	11.763	11.768	148.088	159.856
60	0.287	0.001	0.000	0.020	0.020	0.288	0.308
61	0.871	0.002	0.000	0.062	0.062	0.872	0.934
62	4.900	0.048	0.000	0.349	0.349	4.948	5.297
63	33.863	0.483	0.006	11.763	11.768	34.346	46.114
64	19.583	0.184	0.006	11.763	11.768	19.767	31.535
65	12.729	0.809	0.000	0.195	0.195	13.538	13.734
66	19.393	0.796	0.000	0.195	0.195	20.188	20.384
67	3.444	0.090	0.000	0.062	0.062	3.534	3.596
68	27.486	0.333	0.006	11.763	11.768	27.821	39.58
69	108.138	75.066	0.000	0.334	0.334	183.201	183.534
101	108.368	75.225	0.000	0.335	0.335	183.590	183.925
102	0.045	0.031	0.000	0.000	0.000	0.077	0.077
103	0.184	0.128	0.000	0.001	0.001	0.312	0.313
104	64.194	44.561	0.000	0.275	0.275	108.753	109.029
105	100.640	51.357	0.000	0.311	0.311	151.995	152.306
106	85.571	39.245	0.000	0.316	0.316	124.813	125.129
107	0.097	0.067	0.000	0.000	0.000	0.164	0.165
108	51.056	-35.711	0.000	0.431	0.431	15.344	15.775
109	0.043	0.020	0.000	0.000	0.000	0.063	0.063
110	7.395	5.890	0.000	0.036	0.036	13.284	13.320
111	0.085	0.037	0.000	0.000	0.000	0.122	0.122
112	79.859	1.082	0.006	11.763	11.768	80.941	92.710
113	0.019	0.001	0.000	0.026	0.026	0.020	0.045
114	0.299	0.011	0.000	0.406	0.406	0.311	0.716
115	51.717	-36.232	0.000	0.316	0.316	15.484	15.800
116	0.010	0.000	0.000	0.115	0.115	0.010	0.125
117	0.317	-0.015	0.000	0.431	0.431	0.302	0.733
118	0.318	0.012	0.000	0.431	0.431	0.330	0.761
119	0.003	0.000	0.000	0.000	0.000	0.003	0.003
120	0.318	0.011	0.000	0.406	0.406	0.329	0.734
121	8.211	6.332	0.000	0.036	0.036	14.543	14.579
122	1.963	0.011	0.000	0.406	0.406	1.974	2.380
123	4.440	0.720	0.000	0.026	0.026	5.161	5.1
124	6.052	0.012	0.000	0.431	0.431	6.064	6.495

TABLE 10 (CONT'D)

THERMAL, MECHANICAL, REACTION, ENVIRONMENTAL,
CHEMICAL, PHYSICAL AND TOTAL EXERGY VALUES

STREAM NO.	ETH [MW]	EM [MW]	ER [MW]	EE [MW]	ECH [MW]	EPH [MW]	ETOT [MW]
125	5.101	1.467	0.000	0.349	0.349	6.568	6.917
126	0.287	0.003	0.000	0.020	0.020	0.290	0.311
127	2.242	0.645	0.000	0.153	0.153	2.887	3.040
128	6.592	0.614	0.000	0.153	0.153	7.206	7.359
129	1.739	0.046	0.000	0.011	0.011	1.785	1.796
130	32.560	23.984	0.000	0.142	0.142	56.543	56.685
131	0.903	0.627	0.000	0.004	0.004	1.529	1.533
132	5.311	4.230	0.000	0.026	0.026	9.542	9.567
133	14.031	10.012	0.000	0.055	0.055	24.043	24.098
134	1.128	0.805	0.000	0.004	0.004	1.933	1.938
136	0.282	0.201	0.000	0.001	0.001	0.483	0.485
137	11.161	7.968	0.000	0.044	0.044	19.130	19.173
138	4.759	3.790	0.000	0.023	0.023	8.549	8.572
139	1.132	0.486	0.000	0.006	0.006	1.618	1.624
140	0.084	0.036	0.000	0.000	0.000	0.120	0.120
141	14.025	0.371	0.000	0.093	0.093	14.396	14.489
142	21.291	15.684	0.000	0.093	0.093	36.975	37.068
143	19.429	0.515	0.000	0.128	0.128	19.943	20.071
144	29.495	21.727	0.000	0.128	0.128	51.222	51.350
145	1.415	0.607	0.000	0.008	0.008	2.022	2.030
146	0.183	0.001	0.000	0.006	0.006	0.184	0.190
147	0.099	0.002	0.000	0.001	0.001	0.101	0.102
148	1.471	0.285	0.000	0.243	0.243	1.756	1.999
149	1.478	0.377	0.000	0.243	0.243	1.855	2.098
150	2.539	0.365	0.000	0.243	0.243	2.904	3.148
151	3.910	0.354	0.000	0.243	0.243	4.264	4.507
152	8.205	0.343	0.000	0.243	0.243	8.548	8.792
153	1.351	0.028	0.000	0.020	0.020	1.379	1.399
154	3.856	0.062	0.000	0.044	0.044	3.918	3.961
155	10.551	35.063	797.718	13.312	811.029	45.614	856.643
156	10.933	35.452	797.718	13.314	811.031	46.385	857.416
157	2.065	0.052	0.000	0.044	0.044	2.118	2.162
158	2.074	0.083	0.000	0.044	0.044	2.157	2.201
159	0.391	0.081	0.000	0.044	0.044	0.472	0.516
160	2.859	0.822	0.000	0.195	0.195	3.681	3.876
161	0.883	0.091	0.000	0.062	0.062	0.974	1.036
162	0.287	0.003	0.000	0.020	0.020	0.290	0.311
163	76.972	56.700	0.000	0.335	0.335	133.670	134.005
164	72.172	50.296	0.000	0.311	0.311	122.466	122.777
165	3.647	1.565	0.000	0.020	0.020	5.212	5.232
166	0.131	0.001	0.000	0.008	0.008	0.132	0.140
167	0.043	0.080	0.000	0.044	0.044	0.122	0.167
169	0.085	0.037	0.000	0.000	0.000	0.122	0.122
170	-0.715	35.063	797.718	13.312	811.029	34.348	845.377
171	0.305	0.008	0.000	0.002	0.002	0.313	0.315

TABLE 10 (CONT'D)

THERMAL, MECHANICAL, REACTION, ENVIRONMENTAL,
CHEMICAL, PHYSICAL AND TOTAL EXERGY VALUES

STREAM NO.	ETH [MW]	EM [MW]	ER [MW]	EE [MW]	ECH [MW]	EPH [MW]	ETOT [MW]
172	0.043	0.001	0.000	0.000	0.000	0.044	0.044
173	0.004	0.000	0.000	0.000	0.000	0.004	0.004
174	107.333	74.071	0.000	0.334	0.334	181.401	181.735
175	65.193	45.255	0.000	0.280	0.280	110.447	110.726
176	101.663	52.078	0.000	0.316	0.316	153.739	154.055
177	85.613	39.265	0.000	0.316	0.316	124.876	125.192
178	85.663	39.296	0.000	0.316	0.316	124.957	125.273
179	1.100	0.472	0.000	0.006	0.006	1.572	1.578

TABLE 11

EXERGY OF FUEL, PRODUCT, EXERGY DESTRUCTION, AND THE
EXERGETIC EFFICIENCY FOR EACH PLANT COMPONENT

DEVICE NO.	NAME	FUEL EXERGY [MW]	PRODUCT EXERGY [MW]	EXERGY DESTRUCTION [MW]	EXERGETIC EFFICIENCY [%]
1	OXIDANT FEED SYSTEM	29.021	11.536	17.485	39.752
2	OXYGEN PREHEATER	0.768	0.544	0.223	70.902
3	GASIFIER	1266.868	1131.444	135.424	89.310
4	RADIANT COOLER	47.143	31.279	15.865	66.348
5	CONVECTIVE COOLER	36.737	26.898	9.839	73.218
6	HEAT RECOVERY STEAM DRUM	53.857	49.325	4.532	91.585
7	GAS FILTER	1069.961	1069.636	0.325	99.970
9	GAS COMPRESSOR	0.126	0.105	0.020	83.743
10	GAS COOLING TOWER	868.844	868.238	0.606	99.930
11	QUENCH CIRC. PUMP	0.049	0.039	0.010	79.901
12	QUENCH WATER COOLER	1.685	1.049	0.636	62.281
14	ACID GAS REMOVAL	49.787	46.815	2.972	94.030
15	CLAUS & SCOT	26.655	18.056	8.598	67.742
16	SAT/RH TOWER	51.155	47.660	3.495	93.168
17	RH TOWER CIRC. PUMP	0.127	0.099	0.028	78.236
18	LP STEAM PREHEATER	1.890	1.360	0.530	71.951
19	SAT. WATER PREHEATER	6.940	4.285	2.655	61.742
21	FUEL GAS SUPERHEATER	15.212	11.266	3.946	74.059
22	MIX AFTER FUEL GAS SH	857.614	857.416	0.198	99.977
23	THROTTLE	857.416	857.107	0.309	99.964
24	GAS TURBINE & AIR COMP	381.685	298.800	82.885	78.284
25	COMBUSTION CHAMBER	1139.353	914.185	225.168	80.237
26	S/H AND R/H	92.032	79.449	12.583	86.327
27	HP STEAM DRUM	142.356	134.005	8.351	94.134
28	IP SUPERHEATER	1.543	1.259	0.284	81.613
29	IP STEAM DRUM	25.786	22.887	2.899	88.759
30	HP2 ECONOMIZER	7.350	6.650	0.699	90.484
31	IP ECONOMIZER	3.920	2.560	1.360	65.305
32	HP1 ECONOMIZER	13.092	9.857	3.234	75.294
33	LP STEAM DRUM	6.831	5.232	1.599	76.595
34	FW PREHEATER 1	8.054	5.141	2.913	63.832
35	FW PREHEATER 2	5.134	1.645	3.489	32.043
36	FIRST STEAGE PRESS DROP	183.535	181.735	1.800	99.019
37	HP TURBINE	46.911	43.625	3.286	92.995
39	MIX BEFORE IP TURBINE	154.152	154.055	0.098	99.937
40	IP TURBINE	28.863	25.754	3.110	89.227
42	MIX BEFORE LP TURBINE	125.293	125.273	0.020	99.984
43	LP TURBINE	109.474	89.700	19.774	81.938
44	SSR	0.139	0.122	0.017	87.610
45	MIX BEFORE CONDENSER	15.928	15.775	0.153	99.042
47	PUMP AFTER CONDENSER	0.039	0.028	0.011	71.323
48	GLAND SEAL CONDENSER	0.119	0.018	0.100	15.541
49	DEAERATOR	7.135	6.064	1.071	84.993

TABLE 11 (CONT'D)

EXERGY OF FUEL, PRODUCT, EXERGY DESTRUCTION, AND THE
EXERGETIC EFFICIENCY FOR EACH PLANT COMPONENT

DEVICE NO.	NAME	FUEL EXERGY [MW]	PRODUCT EXERGY [MW]	EXERGY DESTRUCTION [MW]	EXERGETIC EFFICIENCY [%]
50	HP PUMP	2.126	1.620	0.506	76.194
51	IP PUMP	0.133	0.101	0.032	76.138
52	LP PUMP	0.003	0.003	0.001	75.843
56	TOTAL PLANT	1116.574	413.190	703.384	37.005

Figure 8 Exergy Destruction and Associated Costs for Selected Plant Components in Case 1

$$r = \frac{\dot{E}_{D,k}}{\dot{E}_{Coal}} 100$$

Figure 9 Comparison of the relative Exergy Destruction and Exergy Losses in Various Areas of the IGCC Plant in Case 1 and in the Reference Case.

END

**DATE
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