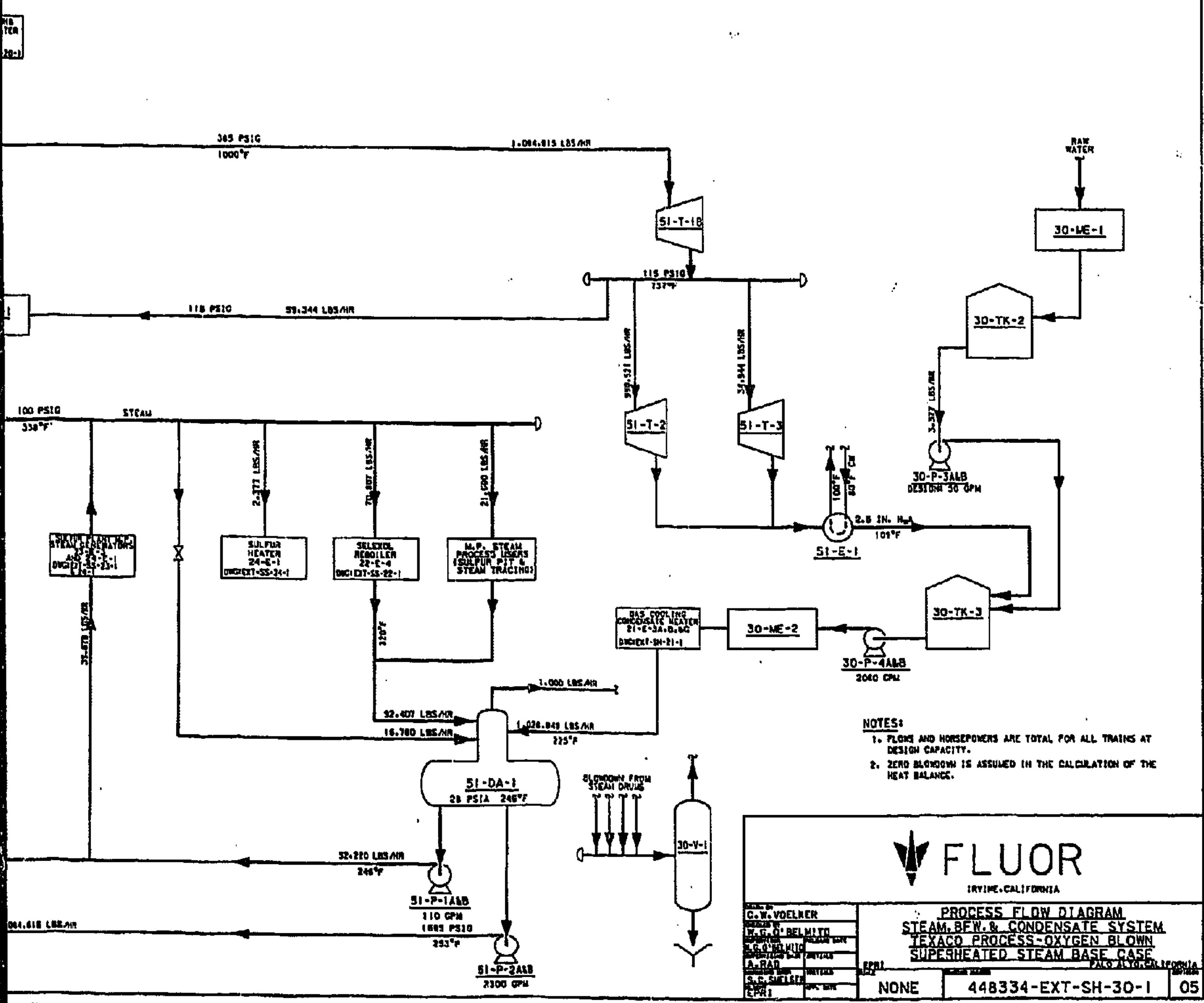
30-V-1 BLOWDOWN FLASH DRUM	51-T-1B I.P. POWER TURBINE	51-T-2 M.P. POWER TURBINE	51-T-3 H.P. BFW PUMP TURBINE	30-TK-3 CONDENSATE SURGE TANK	30-ME-2 CONDENSATE POLISHING UNIT	30-TK-2 DEMINERALIZED WATER STORAGE TANK	30-ME-1 WATER DEMINERALIZATION UNIT
36"ID X 6'-0" T/T DESIGN: 75 PSIG & 350°F CARBON STEEL	53,324 BHP (SHAFT)	126,056 BHP (SHAFT)	4,050 BHP (SHAFT)	CAPACITY: 2000 BBLS	DESIGN CAPACITY: 2100 GPM	CAPACITY: 2000 BBLS EPOXY LINED	DESIGN CAPACITY: 50 GPM



NOTES:

1. FLOWS AND HORSEPOWERS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. ZERO BLOWDOWN IS ASSUMED IN THE CALCULATION OF THE HEAT BALANCE.

FLUOR
IRVINE, CALIFORNIA

C.W. VOELKER		
W.G. O'BELMITO		
W.G. O'BELMITO		
A. RAD		
S.C. SMELSER		
EPRI		

PROCESS FLOW DIAGRAM
STEAM, BFW. & CONDENSATE SYSTEM
TEXACO PROCESS-OXYGEN BLOWN
SUPERHEATED STEAM BASE CASE

EPRI
PALO ALTO, CALIFORNIA

SCALE	DRAWING NUMBER	REVISION
NONE	448334-EXT-SH-30-1	05

483SK1201

COOLING WATER SYSTEMS

This unit provides cooling water for process heat rejection, condensation of exhaust steam from the steam turbines, and cooling of mechanical equipment. Two cooling water systems are provided. The first system, consisting of mechanical draft towers and three low-head pumps, serves only the surface condenser for the steam turbogenerator and high-pressure boiler feedwater pump steam turbine driver. The second system involves one tower and two higher-head pumps. This separation of systems allows the use of the low-head surface condenser water supply pumps by keeping the runs of cooling water lines to a minimum. A further significant advantage is the confinement of contaminants to one cooling tower in the event of a process fluid leak to cooling water.

Makeup water for the surface condenser cooling water system is raw water. The blowdown from this system is treated for calcium hardness in a softener by cold lime-soda addition and is used subsequently as makeup for the process cooling water system. Other makeup flows for the process cooling water system include boiler blowdown and treated effluent from the oily water system. Blowdown from the process cooling tower is an effluent for disposal.

Sulfuric acid is injected into the cooling water system for pH control. A proprietary organic phosphate (nonchromate) solution is injected for corrosion inhibition, scale control, and sludge dispersion. Biocide agents compatible with ammonium ions are injected to maintain clean heat transfer surfaces.

POWER GENERATION

Process Flow Diagrams EXT-SS-50/51-1 and EXT-SH-50/51-1 describe the steam and power generation systems for the saturated and superheated steam base cases respectively.

The power generation system includes two parallel trains of fuel gas expanders (50-1-EX-1) with generators (50-1-G-1) one high-pressure/intermediate-pressure/medium-pressure steam turbine (51-T-1A&B and 51-T-2) with generator 51-G-1 and one fired heater (51-FH-1) in the saturated steam case only. A power block performance summary for all cases is part of Appendix D, which also includes detailed performance information on the power block components, i.e., fuel gas expanders, fired heaters and steam turbines.

EXT-SS. Reheated fuel gas from 20-1-E-3 at 600°F is expanded to 50 psia in two fuel gas expanders 50-1-EX-1, producing 62 MW total net power in two generators 50-1-G-1, while a portion of the expanded gas (3.37×10^6 SCFH) is used in fired heater 50-FH-1, and remainder constitutes product gas for export.

In the fired heater high-pressure saturated steam from gasifier waste heat boilers is superheated to 900°F before flowing to the high-pressure end of back-pressure turbine 51-T-1A. Exhaust steam from 51-T-1A is combined with saturated intermediate-pressure steam from the sulfur plant, and is then reheated to 900°F in the fired heater. This reheated steam is then used in intermediate-pressure power turbine 51-T-1B, a back-pressure machine. There exists a net demand for medium-pressure steam in the process area. Part of exhaust from 51-T-1B is desuperheated and diverted to the process area. The remainder of 51-T-1B exhaust steam is used to drive the medium-pressure power turbine 51-T-2 and the high-pressure boiler feedwater pump driver turbine 51-T-3. These two turbines exhaust at 2-1/2 inches Hg absolute. Total steam turbine generator 51-G-1 output is 259 MW.

The turbine surface condenser 51-E-1, a single-shell single-pass unit with divided water boxes, handles flow from both condensing turbines. The tubes are 90/10 copper/nickel, 7/8 inch OD, 22 BWG wall thickness. The noncondensable gas removal and priming equipment include two positive displacement rotary vacuum pumps and a recirculating ball-type condenser tube cleaning system. Four motor-driven condensate pumps 51-P-3A-D (two spare) transport the condensate to condensate storage tank 30-TK-3, which is sized for 30-minute capacity at design flow rate.

Deaerating steam is provided to deaerator 51-DA-1 by "letting down" medium-pressure steam from header. Medium-pressure feedwater to the process is supplied by pump 51-P-1A&B. The other boiler feedwater users are supplied by high-pressure boiler feedwater pump 51-P-2A&B. High-pressure boiler feedwater is heated in gas cooling boiler feedwater heaters 21-1-E-1 and fired heater 51-FH-1. At which point it is split and one set of streams is fed to the sulfur plant wasteheat boilers and the other is further heated in raw gas cooling boiler feedwater heaters 20-1-E-4, to high-pressure steam saturation temperature. It is then forwarded to gasifier waste heat boilers 20-1-E-1B.

EXT-SH. Reheated fuel gas at 600°F is similarly expanded as in the saturated steam base case generating also 62 MW net power. However, the entire fuel gas exhausting the expander constitutes product gas for export.

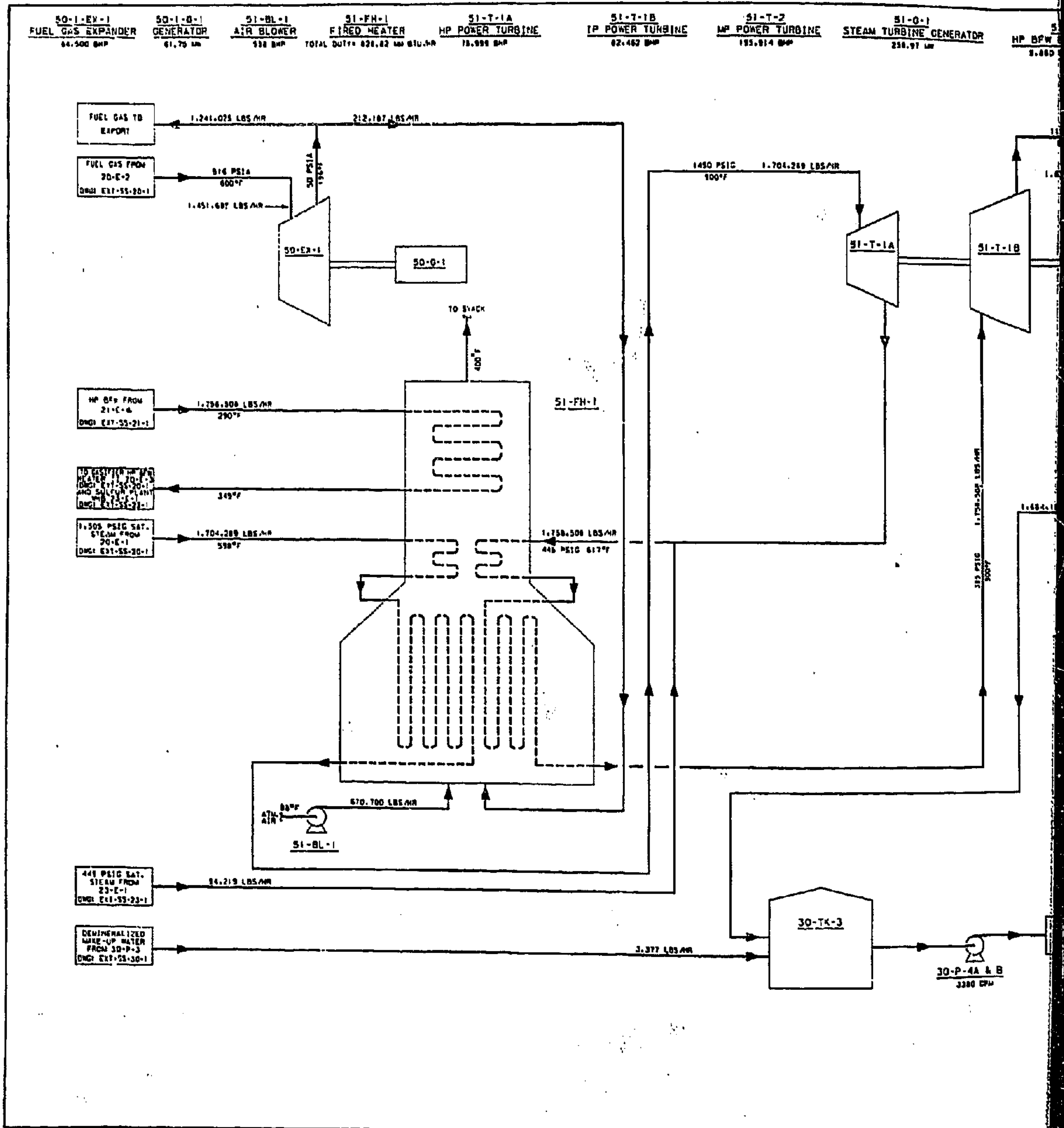
High-pressure superheated steam generated in gasifier waste heat boilers at 1000°F flows to the high-pressure end of back-pressure turbine 51-T-1A. Exhaust steam from 51-T-1A is combined with saturated intermediate-pressure steam from the sulfur plant. It is then reheated to 1000°F in the gasifier waste heat boilers. This reheated steam is then used in a similar fashion as in the SS base case in the power turbine and in the high-pressure boiler feedwater pump driver turbine, and some is desuperheated to provide process steam. The steam turbine generator 51-G-1 output is 167 MW.

The turbine surface condenser 51-E-1 is similar to the saturated steam base case. The condensate is transported and stored in storage tank 30-TK-3 as in the saturated steam base case.

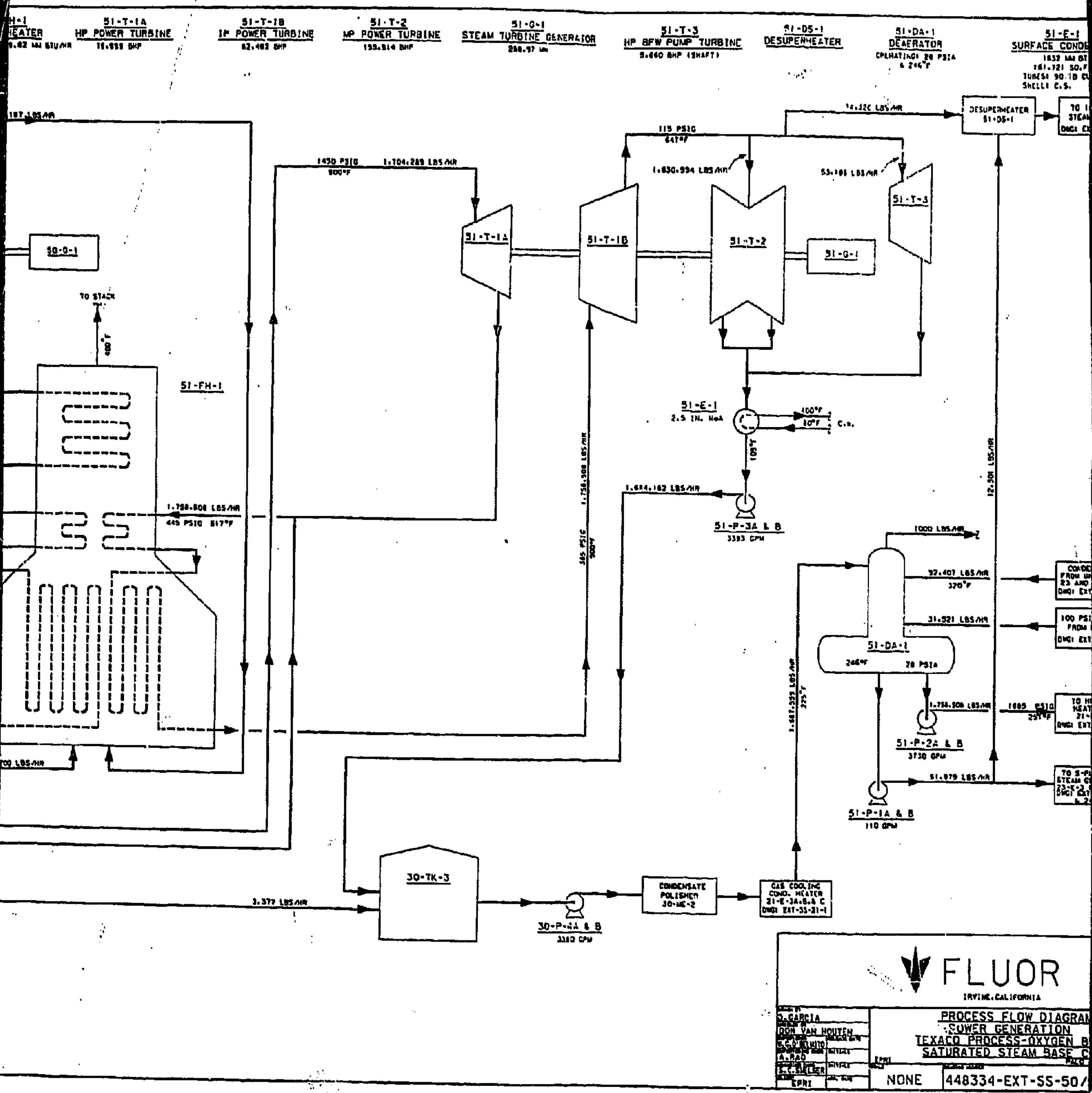
Deaerating steam is provided to deaerator 51-DA-1 by "letting down" medium-pressure steam from header. Medium-pressure feedwater users are supplied by high-pressure boiler feedwater pump 51-P-1A&B. The other boiler feedwater users are supplied by medium-pressure boiler feedwater pump 51-P-2A&B. High-pressure boiler feedwater is heated in a series of exchangers involving gas cooling boiler feedwater heaters 21-1-E-4, at which point it is split and one set of streams fed to the sulfur plant waste heat boilers and the other further heated in raw gas cooling boiler feedwater heaters 20-1-E-4, to high-pressure steam saturation temperature. It is then forwarded to gasifier waste heat boiler 20-1-E-1.

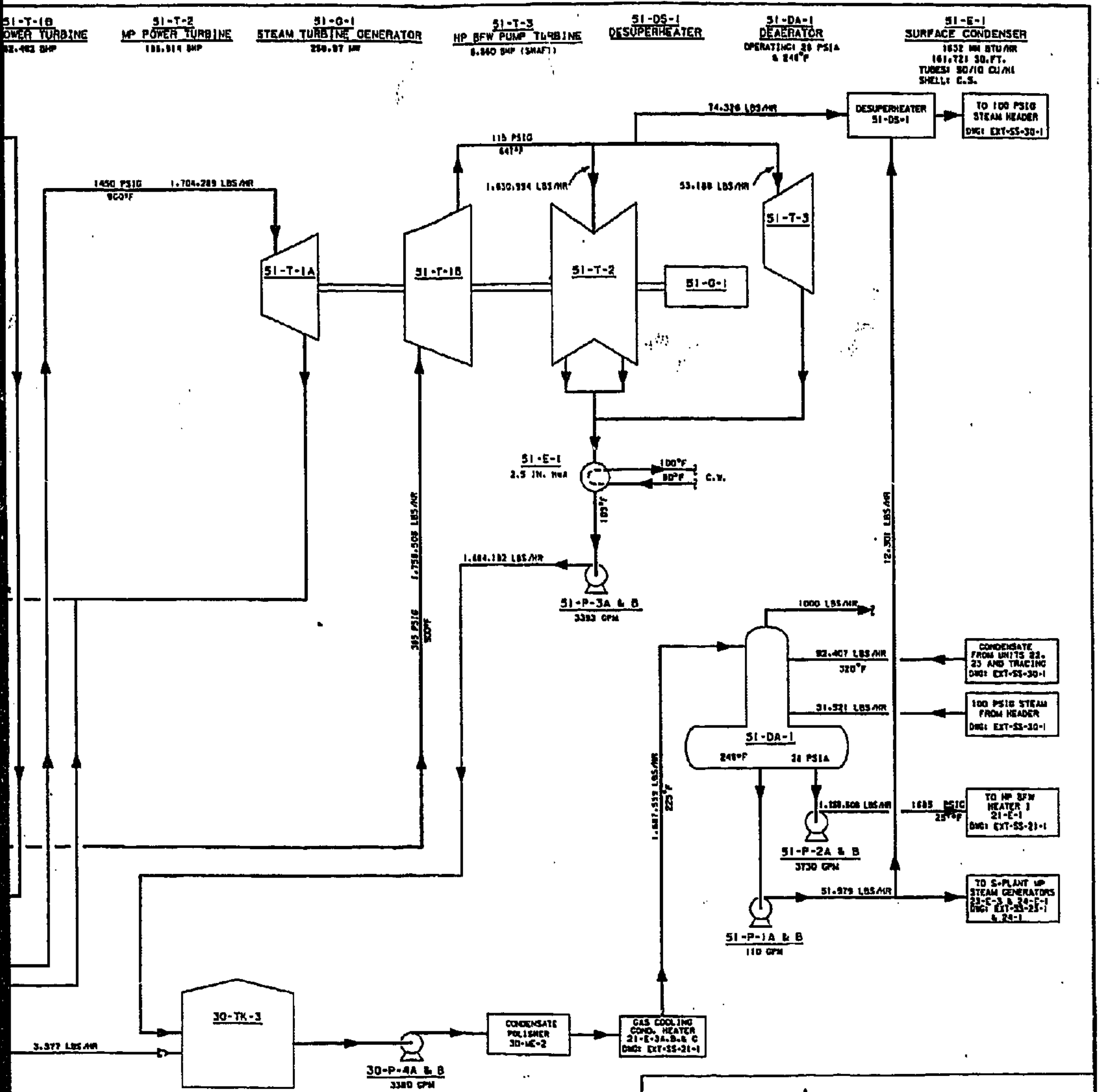
Equipment Notes

With the exception of the high temperature gas coolers, all the equipment is commercially available.



50-1-EX-1
FUEL GAS EXPANDER
50-1-G-1
GENERATOR
51-BL-1
AIR BLOWER
51-FH-1
FIRED HEATER
51-T-1A
HP POWER TURBINE
51-T-1B
IP POWER TURBINE
51-T-2
MP POWER TURBINE
51-G-1
STEAM TURBINE GENERATOR
FUEL GAS TO EXPORT
1,241,025 LBS/HR
212,187 LBS/HR
50-EX-1
50-G-1
TO STACK
51-FH-1
51-T-1A
51-T-1B
51-BL-1
670,700 LBS/HR
30-TK-3
30-P-4A & B
3,377 LBS/HR


51-T-1A
HP POWER TURBINE
51-T-1B
IP POWER TURBINE
51-T-2
MP POWER TURBINE
51-G-1
STEAM TURBINE GENERATOR
51-T-3
HP BFW PUMP TURBINE
51-DS-1
DESUPERHEATER
51-DA-1
DEAERATOR
51-E-1
SURFACE COND
1450 PSIG
900°F
1,704,289 LBS/HR
115 PSIG
1,630,994 LBS/HR
DESUPERHEATER 51-DS-1
50-G-1
TO STACK
51-FH-1
51-E-1
2.5 IN. HgA
100°F
C.W.
1,694,162 LBS/HR
1,758,808 LBS/HR
445 PSIG
51-P-3A & B
3393 GPM
1000 LBS/HR
51-DA-1
28 PSIA
51-P-2A & B
3730 GPM
51-P-1A & B
110 GPM
30-TK-3
30-P-4A & B
CONDENSATE POLISHER 30-ME-2
GAS COOLING COND. HEATER
FLUOR
IRVINE, CALIFORNIA
S. GARCIA
DON VAN HOUTEN
A. RAO
EPRI
PROCESS FLOW DIAGRAM
POWER GENERATION
TEXACO PROCESS-OXYGEN B
SATURATED STEAM BASE C
NONE
448334-EXT-SS-50/



FLUOR
IRVINE, CALIFORNIA

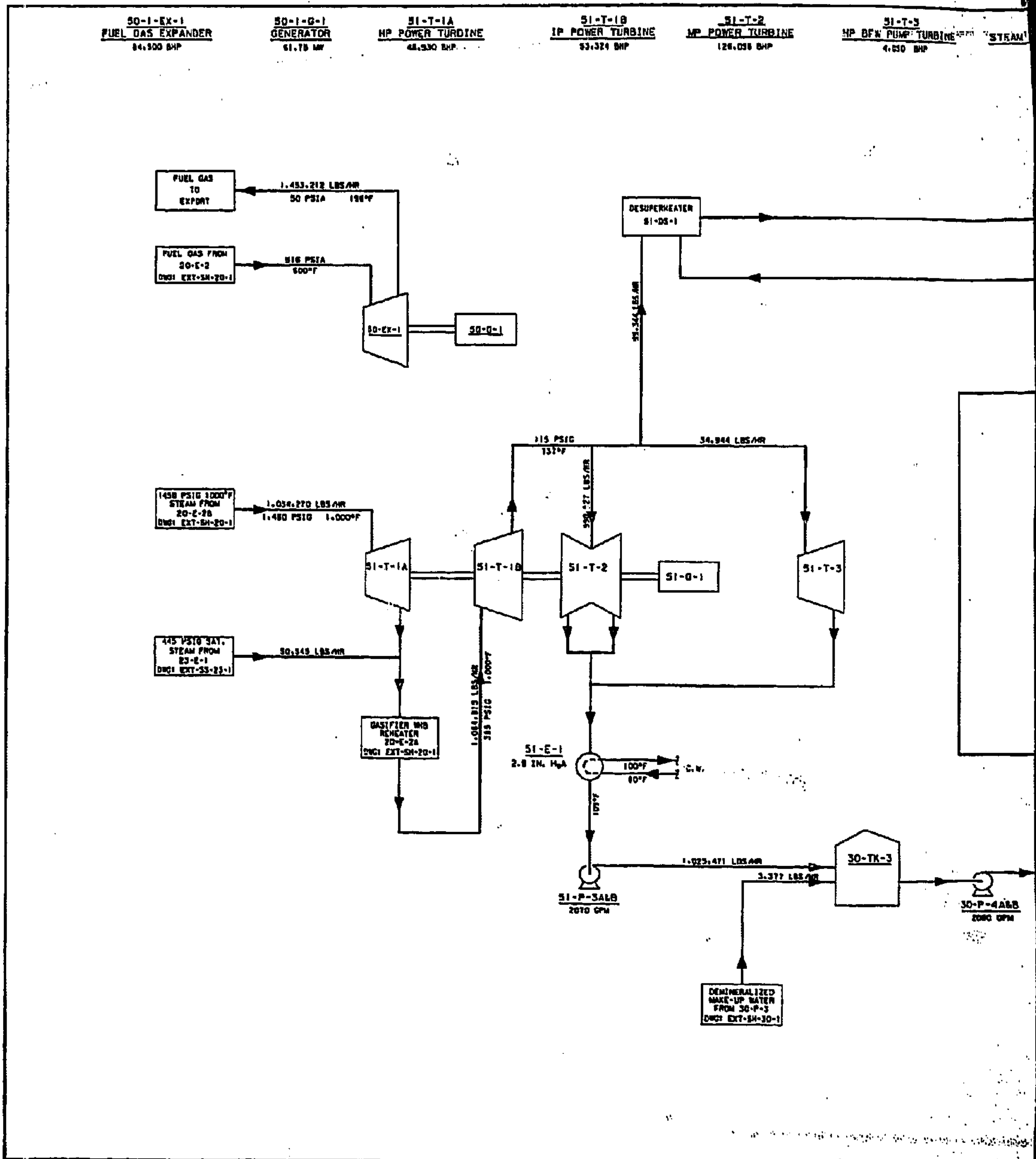
D. GARCIA
DON VAN HOUTEN
A. RAO
S.E. SHELSER
EPRI

PROCESS FLOW DIAGRAM
POWER GENERATION
TEXACO PROCESS-OXYGEN BLOWN
SATURATED STEAM BASE CASE

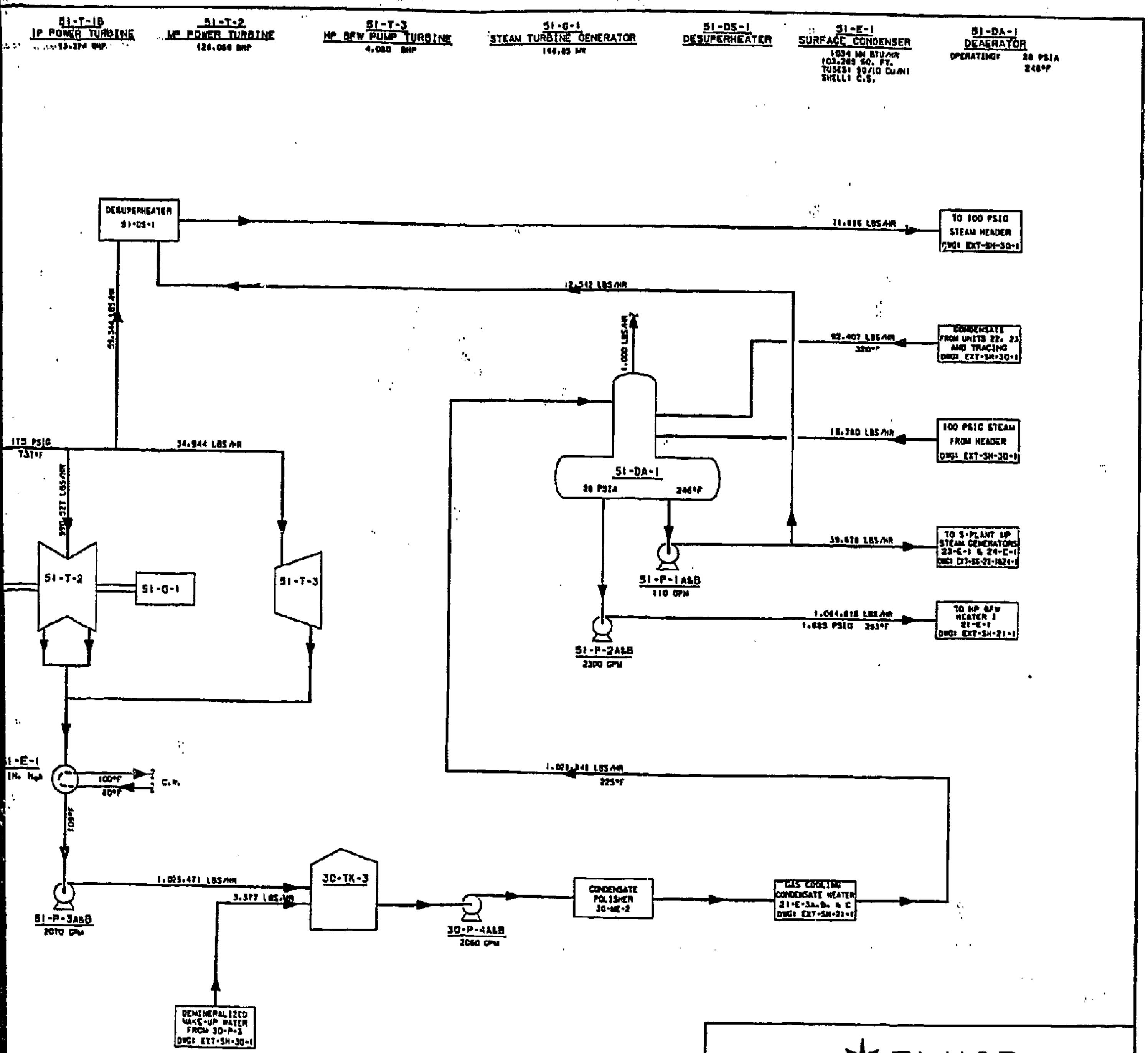
EPRI
PALO ALTO, CALIFORNIA

NONE	448334-EXT-SS-50/51-1A	04

4835K1213



50-1-EX-1
FUEL GAS EXPANDER
84,500 BHP
50-1-G-1
GENERATOR
61.75 MW
51-T-1A
HP POWER TURBINE
48,930 BHP
51-T-1B
IP POWER TURBINE
53,324 BHP
51-T-2
MP POWER TURBINE
126,035 BHP
51-T-3
HP BFW PUMP TURBINE
4,030 BHP
STEAM
FUEL GAS TO EXPORT
1,453,212 LBS/HR
50 PSIA
FUEL GAS FROM 20-E-2 DWG: EXT-SH-20-1
816 PSIA
600°F
50-EX-1
50-G-1
DESUPERHEATER 51-DS-1
315 PSIG
34,944 LBS/HR
1458 PSIG 1000°F STEAM FROM 20-E-2B DWG: EXT-SH-20-1
1,034,270 LBS/HR
1,450 PSIG 1,000°F
51-T-1A
51-T-1B
51-T-2
51-G-1
51-T-3
445 PSIG SAT. STEAM FROM 23-E-1 DWG: EXT-SS-23-1
50,545 LBS/HR
GASIFIER WHB REHEATER 20-E-2A DWG: EXT-SH-20-1
1,000°F
385 PSIG
51-E-1
2.8 IN. HgA
100°F
80°F
C.W.
30-TK-3
1,023,471 LBS/HR
3,377 LBS/HR
51-P-3A&B
2070 GPM
30-P-4A&B
2080 GPM
DEMINERALIZED MAKE-UP WATER FROM 30-P-3 DWG: EXT-SH-30-1



![FLUOR]
IRVINE, CALIFORNIA

D.A. COTTRELL

A. RAO

EPRI

PROCESS FLOW DIAGRAM
POWER GENERATION
TEXACO PROCESS - OXYGEN BLOWN
SUPERHEATED STEAM BASE CASE

NONE	448334-EXT-SH-50/51-1	05

835K1212

GENERAL FACILITIES

The various support systems and services required to produce an operable grass-roots facility are divided into the following units.

- Plant and instrument air
- Potable and utility water
- Fuel system
- Nitrogen system
- Effluent water treating
- Flare system
- Fire water system
- Buildings
- Railroad loading
- Electrical system

Plant and Instrument Air

One motor-driven compressor, rated for 3150 SCFM, is in normal operation supplying plant air and air to instrument air dryer packages. Another motor-driven compressor supplies air on demand, with a third steam turbine-driven compressor on standby for emergency. Two dryer packages are provided; each is designed to supply 1500 SCFM instrument air. Each dryer package is a dual-tower molecular-sieve adsorbent system with air prefilters, air after-filters, and an adsorbent regeneration system.

Potable and Utility Water

The potable water system includes two motor-driven 100 gpm pumps, with in-line chlorination. Water is pumped from raw water storage tank 30-TK-1 to an air-pressurized supply drum. Plant water is supplied by two motor-driven pumps, each rated for 200 gpm, operating on demand.

Fuel System

Fuel oil is used as startup fuel for the fired heater in the saturated steam case only. The fuel system has two 33,000 bbl tanks, two motor-driven pumps and suction heaters (for viscosity reduction). This fuel oil capacity is sufficient to support full operation of the fired heater for two weeks. The multiple tank system allows the use of more than one type of fuel oil.

Nitrogen System

Nitrogen gas is required for blanketing fuel oil storage tanks and for purging process equipment prior to maintenance. Liquid nitrogen obtained from the air separation plant is stored in a double-walled 7200 bbl cryogenic vessel, from which nitrogen gas is supplied upon demand by vaporization of the liquid in air-fin heaters.

Effluent Water Treating

The water streams treated are:

- Storm water
- Utility wastewater
- Cooling tower blowdown

A process condensate blowdown rate of 170 gpm from the gasification and ash handling section was selected as the sizing basis for the process condensate treating unit. This sizing basis was derived from the overall water balance for the enriched air cases. Process condensate composition data are not adequate to finalize the flow which must be treated in a full-scale commercial GCC plant. Therefore, the size of the process condensate treating unit is assumed to be constant for all plants designed and costed in this study. The process itself is designed to remove formates, sulfides, ultrafine ash particles, and ammonia from the water using the following steps:

- Chemical addition with precipitation
- Settling
- Filtration
- Ammonia stripping with steam
- Biotreatment

The effluent water from the process condensate treating unit is suitable for disposal in a navigable body of water. By-products are a precipitate cake and a biotreater sludge both of which may be combined with the ash cake for disposal.

The ammonia stripped from the water is routed to the sulfur plant (Unit 23) furnace. The quantity of ammonia has been judged insufficient to consider its recovery for sale.

Storm water and utility waste waters are directed in underground sewers to the forebay of the storm water pond. Contaminated water from this pond is treated for oil removal in a CPI separator, processed in a deep-bed filtration unit, and then used as makeup water for the process cooling tower.

The cooling water system includes one process and two utility mechanical draft cooling towers for the saturated steam case. The superheated steam case cooling water system includes one process and one utility cooling tower. Blowdown from the utility cooling towers is softened by cold lime-soda treating before being added to the process cooling water system. Blowdown from the process cooling tower is combined with the treated process condensate prior to disposal.

Sanitary sewage streams are sent to the city sewer outside plant boundary limits.

<u>Flare System</u>

In conformance with accepted practice, a relief system is provided to protect the process equipment from overpressure. In the event of pressure release, relief lines will carry away the vented gases from the affected processing areas to two elevated flare stacks where ignition will occur. A continuous flare system pilot flame is maintained by a package LPG system comprised of an LPG tank, pumps, and a vaporizer. Separator drums are provided at the base of each flare stack to capture condensate which may be carried in the vented process gases. Sealing systems are provided in each flare stack to prevent air intrusion back into the relief system.

<u>Fire Water System</u>

A fire water loop encompassing the entire plant is provided. A motor-driven jockey pump keeps the system under pressure. All equipment and storage tanks are within range of hydrants and monitors in accordance with accepted practice. A total capacity of 5000 gpm is provided by two fire water pumps, one of which is motor-driven while the other is powered by a diesel engine. These pumps take suction either from a 30,000 bbl fire water storage tank or directly from the municipal water supply. The pumps are designed to start automatically when fire water loop pressure drops appreciably. The pumps are also designed to deliver water, at the design flow rate, to the hydrants at a pressure of 125 psig.

Buildings

The following is a list of the buildings included in the capital estimate:

- Substations
- Control Houses
- Operators Shelters
- Administration
- Laboratory
- Cafateria
- Change House and Guard House
- Fire House
- First Aid
- Maintenance
- Warehouse

Section 3

PROCESS DISCUSSION

The two base cases, involving the generation of high-pressure saturated steam (EXT-SS) and the generation of high-pressure superheated steam (EXT-SH) in high temperature gas cooling, are discussed here.

OVERALL SYSTEM PERFORMANCE

Table 3-1 summarizes the overall system performance for the base case designs. The table is organized into gasification, power system, and overall plant categories.

Both cases produce a gasifier effluent having a higher heating value of 275.8 Btu/SCF (dry). After the gas is cleaned to remove particulates and sulfur compounds, the higher heating value of gas available for export corresponds to 282.3 Btu/SCF (dry). The gas is available for export from the plant boundary limits at a pressure of 50 psia. If the gas is to be transported over longer distances, fuel gas reheat as well as gas expansion would be eliminated, and gas would be available at the plant boundary limits at a pressure of 536 psia. The high pressure steam generation rate would be increased due to the elimination of fuel gas reheater. However, the net effect, as expected, would be a reduction in the overall plant efficiency.

For the saturated steam case, 6664 x 10^6 Btu/hr of fuel gas and 142.4 MW of electrical power are available for export. For the superheated steam case, more fuel gas, 7637 x 10^6 Btu/hr, and less electrical power, 53.3 MW, are available for export. The energy recovery efficiency for the saturated and superheated steam cases is 77.11 percent and 78.77 percent respectively.

Table 3-1

SUMMARY OF SYSTEM PERFORMANCE TEXACO GASIFICATION
FUEL GAS PLANTS - BASE CASES [1]

Steam Cycle, psig/°F/°F	1450/900/900	1450/1000/1000
Steam Generated in Gas Coolers	Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F	2400	2400
Sulfur Removal, %	94.6	94.6
Oxidant Plant Compressor Drivers	Motors	Motors
Nominal Capacity of Gasifiers ST/day	1375	1375
CASE DESIGNATION	EXT-SS	EXT-SH
GASIFICATION		
Coal Feed Rate, lb/hr (dry)	806,666	806,666
Oxygen*/Coal Ratio, lb/lb m.f.	0.8921	0.8921
Oxidant Temperature. °F	300	300
Slurry Feed Solids Content, weight %	66.5	66.5
Gasification Section Average Pressure, psig	600	600
Raw Gasifier Effluent Temperature, °F	2,400	2,400
Raw Gasifier Effluent HHV (dry basis), Btu/SCF**	275.8	275.8
Cold Gas Efficiency (raw gas HHV/coal feed HHV x 100),	74.64	74.64
POWER SYSTEM		
Temperature of Fuel Gas to Gas Expander, °F	600	600
Gas Expander Exit Temperature, °F	195	195
Condenser Pressure, Inches Hg abs	2.5	2.5
Fired Heater Stack Temperature, °F	400	[illegible]
Gas Expander Power#, MW	61.75	61.75
Steam Turbine Power#, MW	258.97	166.85
Oxygen Plant Power##, MW	1.81	1.81
Power Consumed, MW	180.14	177.11
Net System Power, MW	142.40	53.30
OVERALL SYSTEM		
General Facilities Water Consumption, GPM	150	150
Land, acres	190	175
Ash Disposal Rate, Dry ST/D	1023	1023
Sulfur By-Product, ST/D	354	354
Process and Deaerator Makeup Water, GPM	467	450
Cooling Tower Makeup Water, GPM	4,370	3,212
Cooling Water Circulation Rate§§, 10^3 GPM	227	167
Cooling Tower Heat Rejection§§, % of coal HHV	22.11	16.26
Air Cooler Heat Rejections, % of coal HHV	2.10	3.04
Clean Fuel Gas Efficiency (exported clean gas HHV x 100/coal feed HHV), %	64.67	74.11
Energy Recovery Efficiency¶ (exported power + exported clean gas HHV)/ coal feed HHV x 100, %	77.11	78.77
Clean Fuel Gas HHV (dry basis), Btu/SCF	282.3	282.3
Net Clean Fuel Gas Product, 10^6 SCFD	566.53	649.25
10^3 SCF/ton DAF coal	65.03	74.52
10^6 Btu/hr	6664	7637

(1) This table is identical to Table S-1, page S-3
* Dry basis, 100 percent oxygen
** Excluding the HHV of H_2S, COS and NH_3
\# At generator terminals
\## From power recovery expander 11-ME-1
§§ Includes process and power plant portions
¶ Export power credited at 9,000 Btu/kWh.

GASIFIER MATERIAL BALANCE

A gasifier material balance for full capacity operation is given in Table 3-2.

The coal quantity and composition, the oxidant composition, the slurry concentration and reaction temperature were taken from EPRI report AP-1624. Fluor has estimated both the quantity of oxidant required and the effluent gas composition. The coal feed is 11,000 tons/day of Illinois No. 6, introduced to the gasifier in a 66.5 weight percent slurry. Texaco has indicated that for this particular coal, slurry concentrations in the range of 60 to possibly 70 percent solids could be achieved. It is important to bear in mind that slurrying characteristics of coals vary greatly and that it is not valid to extrapolate performance estimates presented in this report to other coals possessing different slurrying characteristics. Per-pass carbon conversion in the gasifier is assumed to be 95 percent.

Some nitrogen from the coal is converted to ammonia, but the extent of conversion cannot be predicted with precision at this time. This study has assumed that 25 percent of the coal's nitrogen forms ammonia. Downstream of the gasifier, ammonia has been assumed to be rapidly complexed as ammonium salts in the process condensate streams. These ammonia-bearing waters, with the exception of a blowdown stream, are eventually recycled to the gasifiers by way of the coal slurry. At gasification temperatures, the gasifiers are assumed capable of destroying recycled ammonia. The presence of ammonia in the process condensate has the beneficial effect of neutralizing dissolved carbon dioxide.

HIGH TEMPERATURE GAS COOLING/STEAM GENERATION

In both base cases the high temperature heat recovery equipment consists of a radiant boiler followed by convective heat recovery equipment. The temperature assumed for the gas exiting the radiant boiler is 1500°F. This temperature was chosen based on the assumption that the entrained ash in the gas will be sufficiently cool at 1500°F that it will minimize fouling in the convective exchangers.

The quantity of electric power generated by the steam raised in the hot gas cooling section of the plant depends on the type of service assumed for the high temperature gas coolers. For Case EXT-SS in which only saturated steam is raised in these coolers, all of the sensible heat in the fuel gas down to 640°F is used for steam raising. For Case EXT-SH, where superheating/reheating is assumed, the

Table 3-2

GASIFIER MATERIAL BALANCE

FEEDS

	T(°F)	lb/hr	lb mol/hr
Coal	140		
Moisture		110,000	6,105.9
Ash		80,667	
MAF Coal			
Carbon		560,909	46,699.8
Hydrogen		42,979	21,330.8
Oxygen		80,876	2,525.3
Nitrogen		10,091	360.2
Sulfur		31,145	972.3
TOTAL COAL		916,667	
Oxidant	300		
Oxygen		719,709	22,490.9
Argon		4,589	114.7
Nitrogen		9,638	344.2
TOTAL OXIDANT		733,936	22,949.8
Recycle	140		
Carbon		24,680	2,054.8
Ash		66,147	
Water		341,512	18,990.5
		432,339	
TOTAL FEEDS		2,082,942	

EFFLUENTS

	T(°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Gaseous/Vapor Effluent	2,300-2,600			
CH_4		3,533	220.2	0.24
H_2		53,102	26,340.1	28.35
CO		993,074	35,453.0	38.17
CO_2		465,615	10,579.5	11.39
H_2S		30,927	907.4	0.98
COS		3,835	63.8	0.07
N_2		17,213	614.4	0.66
Ar		4,584	114.7	0.12
H_2O		331,899	18,422.5	19.83
NH_3		3,067	180.1	0.19
TOTAL GASIFIER GASEOUS/VAPOR EFFLUENT		1,906,849	92,895.9	100.00
Gasifier Solid Effluent	2,300-2,600			
Carbon		29,279	2,437.7	
Ash		146,814		
TOTAL GASIFIER SOLID EFFLUENT		176,093		
TOTAL EFFLUENTS		2,082,942		

total quantity of steam raised is less than that in Case EXT-SS due to the additional duty required in the superheater/reheater. Therefore the total quantity of electric power generated in the saturated steam case (EXT-SS) is greater than that for the case where superheating/reheating capability is assumed (EXT-SH). This greater power production capability for Case EXT-SS is offset by a smaller net fuel gas make due to the consumption of approximately 13 percent of the clean fuel gas in the fired superheater/reheater.

POWER CONSUMPTION SUMMARY

Table 3-3 summarizes auxiliary power consumptions for the major plant sections under conditions of normal operation at 100 percent capacity factor.

Unit 10, requires 9176 kW of power at full design capacity. The coal grinding equipment (10-ME-4) consumes the greatest proportion of this power to the coal handling equipment. Since the power for grinding is very sensitive to both the type of equipment used and the coal properties, this estimate of power required for coal handling is preliminary, pending grinding tests.

Unit 11 power requirements represent by far the most substantial power consumption in the designs and are due solely to the air and oxygen compressor drivers (11-C-1-M and 11-C-2-M) which are synchronous, Type II induction motors. All intercoolers are water-cooled and the cooling tower pump power is included elsewhere.

Miscellaneous pumps use most of the power for Units 20 and 21.

In Unit 22, the Selexol® fluorocarbon refrigeration units (22-ME-1) and lean solvent pumps (22-P-1) consume most of the power. It should be noted that the hydraulic turbines (22-HT-1) recover approximately 40 percent of the power required by the lean solvent pumps.

In the sulfur recovery unit, the Claus plant air blowers (23-BL-1), the Stretford plant air blowers (24-BL-1), and the Stretford solution circulation pumps are the large power consumers.

The major power consumer in Unit 30 is the condensate transfer pump (30-P-4).

Table 3-3

POWER CONSUMPTION SUMMARY (kW at Full Design Capacity)

TEXACO GASIFICATION FUEL GAS PLANTS - BASE CASES

Steam Cycle, psig/°F/°F		1450/900/900	1450/1000/1000
Steam Generated in Gas Coolers		Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F		2,400	2,400
Sulfur Removal, %		94.6	94.6
Oxidant Plant Compressor Drivers		Motors	Motors
Nominal Capacity of Gasifiers, ST/day		1375	1375
Case Designation		EXT-SS	EXT-SH
Process Unit Number & Description		Power Consumption, kW	Power Consumption, kW
10	Coal Handling*	9,176	9,176
11	Oxidant Feed	143,439	143,439
20	Gasification	1,710	1,710
21	Gas Cooling	941	941
22	Acid Gas Removal	5,963	5,963
23	Sulfur Recovery	584	584
24	Tail Gas Treating	2,610	2,610
30	Raw Water Treating	323	222
32-45	General Facilities**	14,860	12,377
50	Fuel Gas Expander	0	0
51	Steam System	513	70
51	Surface Condenser#	16	10
	TOTAL PLANT POWER, kW	180,135	177,111
	Power Produced, kW	322,530	230,410
	Net Power Output, kW	142,400	53,300

*This power requirement is largely derived from the coal pulverizing system and is, therefore, preliminary pending grinding tests on Illinois No. 6 coal.

**About 70 percent of this power demand is attributable to the cooling water system pumps and fans.

#For mechanical vacuum pump.

In the general facilities section, the cooling water circulation pumps and the cooling tower fans require the most power. The entire cooling water system is responsible for 60 to 70 percent of the total power demand of the general facilities (Units 32 through 45).

Pumps consume the bulk of the power supplied to Units 50 and 51.

PROCESS ENERGY BALANCES

Tables 3-4 and 3-5 represent the overall process energy balances at 100 percent capacity operation for the two base cases. The entire plant, exclusive of the cooling tower but including the power demand of the pumps and fans, is encompassed by the boundary for each balance. Energy contents of streams crossing the boundary are expressed as the sum of the stream's higher heating value (HHV), sensible heat above 60°F, and the latent heat of water at 60°F. Electric power is converted to equivalent heat energy at 3413 Btu/kWh. The energy balances close within three-tenths of a percent. Discrepancies result from approximations applied in calculating enthalpies for some process streams.

Energy balances as a percent of coal HHV, depicted in Table 3-6, are derived from the previous tables. Coal charged at 11,000 short tons/day is equivalent to 10,304 x 10^6 Btu/hr HHV. The significant differences among the cases lie in the "Net Power," "Net Fuel Gas," and "Steam Turbine Condensers" categories. As noted previously, differences in the gas cooling process and the steam system have a substantial impact on the net power generation and fuel gas available for export.

Table 3-4

ENERGY BALANCE - BASE CASE EXT-SS

Gasifier: Texaco

Steam: 1450 psig, 900°F/900°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 2400°F

Sulfur Removal: 94.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	4	1	-	-	5
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	615	615
TOTAL	10,304	33	57	-	615	11,009
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	123	-	123
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	884	884
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	74	76	-	-	150
Fired Heater Radiation Losses	-	-	-	18	-	18
Steam Turbine Condensers	-	1,632	-	-	-	1,632
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	22	22
Motor and Mechanical Losses**	-	-	-	-	116	116
Net Fuel Gas Export	6,664	60	-	-	-	6,724
TOTAL	6,847	2,628	141	141	1,239	10,996

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.12\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 3-5

ENERGY BALANCE - BASE CASE EXT-SH

Gasifier: Texaco

Steam: 1450 psig, 1000°F/1000°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water is liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 2400°F

Sulfur Removal: 94.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	604	604
TOTAL	10,304	29	56	-	604	10,993
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	113	-	113
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	47	50	-	-	97
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	570	570
Air Separation Plant Power	-	-	-	-	6	6
Steam Turbine Condensers	-	1,034	-	-	-	1,034
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	16	16
Motor and Mechanical Losses**	-	-	-	-	105	105
Net Fuel Gas Export	7,637	69	-	-	-	7,708
TOTAL	7,820	2,012	115	113	908	10,968

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.23\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 3-6

ENERGY BALANCE AS PERCENT OF COAL HHV

TEXACO GASIFICATION FUEL GAS PLANT - BASE CASES

Steam Cycle, psig/°F/°F	1450/900/900	1450/900/900
Steam Generated in Gas Coolers	Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F	2,400	2,400
Sulfur Removal, %	94.6	94.6
Oxidant Plant Compressor Drivers	Motors	Motors
Nominal Capacity of Gasifiers, ST/day	1375	1375

CASE DESIGNATION	EXT-SS		EXT-SH	
	10^6 Btu/hr	%	10^6 Btu/hr	%
IN				
Coal HHV	10,304	100.00	10,304	100.00
OUT				
Oxidant Compressors Inter/After Cooling	468	4.54	468	4.54
Air Separation Plant Vent Gas, Sensible and Latent	42	0.41	42	0.41
Gasifier and WHB Heat Losses	123	1.19	113	1.10
Ash Cake	71	0.69	71	0.69
Selexol® and Tail Gas Treating Cooling	112	1.08	112	1.08
Sulfur By-product	117	1.13	117	1.13
Sulfur Plant Vent Gas, HHV, Sensible and Latent	8	0.07	8	0.07
Net Power	486	4.72	182	1.77
Fired Heater Flue Gases, Sensible and Latent	150	1.46	0	0
Fired Heater Radiation Losses	18	0.18	0	0
Steam Turbine Condensers	1,632	15.84	1,034	10.04
Power Block Losses	22	0.21	16	0.16
Motor Losses	116	1.13	105	1.02
Other (net) Sensible Losses	234	2.27	284	2.76
Other (net) Latent Losses	<32>	<0.31>	19	0.18
Net Fuel Gas Export, HHV and Sensible	6,724	65.26	7,008	74.80
TOTAL	10,291	99.87	10,279	99.76

*Values indicated within < > are negative numbers

Section 4

CAPITAL AND OPERATING COST ESTIMATES

PLANT FACILITIES INVESTMENT

A breakdown of the plant facilities investment for the major units of the base case designs, including initial catalysts and chemicals, is given in Tables 4-1 and 4-2. The estimates in these tables contain allowances for both project and process contingencies. The project contingency allowance is intended to cover additional or improved equipment that would result from a more detailed design of a definitive project at an actual site. An allowance of 15 percent of the sum of Process Plant Investment and the General Facilities Cost has been used for both base cases. The process contingency allowance, applied separately to each major plant section, is an attempt to account for unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. The different process contingency allowances used for each major subsection of the plant designs are shown in Table 4-3. These allowances were supplied by EPRI.

The absolute accuracy of the plant investment estimates is judged to be ±25 percent. Comparisons between cases should be much more accurate, perhaps ±5 percent, since the same inaccuracies are likely to occur in each case.

As anticipated, the cost of the gas cooling section varies widely for the two cases due to the different waste heat boiler options. Also showing significant variation are the sections associated with power generation such as the turbogenerator and the fired heater.

EPRI incremented by 27 percent the mid-1978 plant facilities investments for the two cases in order to escalate the investment estimate to mid-1980 dollars as shown in Table 4-4.

Table 4-1

PLANT FACILITIES INVESTMENT - BASE CASE EXT-SS
Mid-1978 Dollars

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/MM BTU/HR
Coal Handling	20,566	8,356	14,413	1,026	44,361	6.657	8.2	0	.6,654	51,015	7.657
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	23.417	29.0	0	23,407	179,453	26.930
Gasification and Ash Handling	24,062	9,810	16,702	1,171	51,745	7.765	9.7	5,133	7,762	64,640	9.700
Raw Gas Cooling	71,124	16,455	34,929	3,590	126,098	18.923	23.4	22,825	18,913	167,836	25.186
Acid Gas Removal	6,918	2,493	4,513	353	14,277	2.142	2.7	0	2,142	16,419	2.464
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.899	1.1	0	899	6,889	1.034
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.552	2.0	1,551	1,551	13,441	2.017
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	2.010	2.5	0	2,009	15,402	2.311
Steam, Condensate and BFW	970	351	426	19	1,766	.265	.3	0	265	2,031	.305
Steam Superheat/Reheat	7,044	2,090	4,070	363	13,567	2.036	2.5	0	2,035	15,602	2.340
Steam Turbine Generator	20,858	6,189	12,049	1,074	40,170	6.028	7.5	0	6,026	46,196	6.932
General Facilities	28,485	11,526	16,801	963	57,775	8.670	10.7	0	8,666	66,441	9.970
Initial Chemicals and Catalysts	-	-	-	-	2,358	0.354	0.4	-	-	2,358	0.354
TOTAL	279,337	93,220	151,854	11,112	537,885	80.718	100.0	29,509	80,329	647,723	97.200

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/MM BTU/HR**
Process Plant Investment and General Facilities	537,885	80.718
Process Contingency	29,509	4.428
Project Contingency	80,329	12.054
Total Plant Investment	647,723	97.200

*Mid-1978 dollars

**Based on 100 percent plant design power output of 6663.8 MM Btu/hr Fuel Gas

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision, and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 4-2

PLANT FACILITIES INVESTMENT - BASE CASE EXT-SH
Mid-1978 Dollars

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/MM BTU/HR
Coal Handling	20,566	8,356	14,413	1,026	44,361	5.809	8.5	0	6,654	51,015	6.681
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	20.433	29.9	0	23,407	179,453	23.499
Gasification and Ash Handling	24,062	9,810	16,702	1,171	51,745	6.776	9.9	5,133	7,762	64,640	8.464
Raw Gas Cooling	82,055	18,973	40,318	4,147	145,493	19.051	27.9	32,010	21,822	199,325	26.100
Acid Gas Removal	6,918	2,493	4,513	353	14,277	1.870	2.7	0	2,142	16,419	2.150
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.784	1.1	0	899	6,889	.902
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.354	2.0	1,551	1,551	13,441	1.760
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.754	2.6	0	2,009	15,402	2.017
Steam, Condensate and BFW	754	271	319	13	1,357	.178	.3	0	204	1,561	.204
Steam Turbine Generator	13,543	4,018	7,823	697	26,081	3.415	5.0	0	3,912	29,993	3.927
General Facilities	25,738	10,206	15,091	895	51,930	6.800	9.9	0	7,790	59,720	7.820
Initial Chemicals and Catalysts	-	-	-	-	1,277	0.167	0.2	-	-	1,277	0.167
TOTAL	272,946	90,077	147,134	10,855	522,289	68.391	100.0	38,694	78,152	639,135	83.691

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/MM BTU/HR**
Process Plant Investment and General Facilities	522,289	68.391
Process Contingency	38,694	5.066
Project Contingency	78,152	10.234
Total Plant Investment	639,135	83.691

*Mid-1978 dollars

**Based on 100 percent plant design power output of 7,636.8 MM Btu/hr Fuel Gas

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:

- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 4-3

PROCESS CONTINGENCIES
TEXACO-BASED GASIFICATION FUEL GAS PLANT
BASE CASES

Steam Cycle, psig/°F/°F	1450/900/900	1450/1000/1000
Steam Generated in Gas Coolers	Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F	2,400	2,400
Sulfur Removal, %	94.6	94.6
Oxidant Plant Compressor Drivers	Motors	Motors
Nominal Capacity of Gasifiers, ST/day	1,375	1,375
CASE DESIGNATION	EXT-SS	EXT-SH
PROCESS UNIT		
Coal Handling	0	0
Oxidant Feed	0	0
Gasification	15	15
Ash Handling	5	5
Gas Cooling		
Radiant WHB	20	20
Convective WHB	25	N/A
Convective Superheater or Reheater	N/A	40
Acid Gas Removal	0	0
Sulfur Recovery (Claus)	0	0
Tail Gas Treating	15	15
Process Condensate Treating	0	0
Steam, Condensate, and BFW	0	0
Fuel Gas Expansion	0	0
Steam Superheater/Reheater	0	N/A
Steam Turbine-Generator	0	0
General Facilities	0	0

Table 4-4

PLANT FACILITIES INVESTMENT - TEXACO BASED FUEL GAS BASE CASES

Mid-1980 Dollars

Steam Cycle, psig/°F/°F	1450/900/900	1450/1000/1000
Steam Generated in Gas Coolers	Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F	2,400	2,400
Sulfur Removal, %	94.6	94.6
Oxidant Plant Compressor Drivers	Motors	Motors
Size of Gasifiers, ST/d	1,375	1,375
Fuel Gas Production, 10^6 Btu/hr*	6,664	7,637
Net By-Product Power, MW*	142.40	53.30
CASE DESIGNATION	EXT-SS	EXT-SH
Plant Facilities Investment Mid-1978 ($1,000)	647,723	639,135
Plant Facilities Investment Mid-1980 ($1,000)	822,608	811,701

*Production at design capacity

General Facilities

As expected, the cost for the general facilities is higher for the saturated steam base case. Since more power is generated in this case, the cooling water demand increases, and since the cooling tower represents the major cost of the general facilities, the overall cost of this unit is substantially increased relative to the superheated steam base case.

TOTAL CAPITAL REQUIREMENT

The Total Capital Requirement for each case is defined as the sum of plant facilities investment, prepaid royalties, organization and startup costs, working capital, allowance for funds during construction and land costs. The plant facilities investment estimates (in mid-1978 dollars) for both cases studied have already been detailed in Table 4-1 and 4-2. The basis for estimating allowances for all of the other capital charges specified above is outlined in Table 4-5.

The mid-1980 Total Capital Requirements as defined above are shown in Table 4-6. The actual cost estimates were prepared in mid-1978 dollars. Mid-1980 dollar plant cost estimates were determined by increasing the mid-1978 dollar estimates by 27 percent. (The 27 percent represents experience with the escalation of installed costs for similar types of equipment in the two year period, from mid-1978 to mid-1980).

The constant dollar mid-1980 allowances for funds during construction (AFDC) were calculated as follows:

- Assuming construction to commence in January 1986 and end in December 1989 (a four year period), and assuming annual construction expenditures to be 15 percent, 25 percent, 35 percent, and 25 percent of the total funds required for the four years respectively, the actual current dollar capital outlays for each of the four years were calculated assuming an annual inflation rate of 10 percent.
- Interest charges on the current dollar capital outlays for each of the four years were calculated to December 1989 based on an interest rate of 12.25 percent per annum.
- The total current dollar investment, which includes escalated amounts for prepaid royalties, organization and startup costs, working capital, land and the AFDC charges, was de-escalated from December 1989 to July 1980 at the general inflation rate of 10 percent/year to generate the constant dollar, mid-1980 Total Capital Requirement shown in Table 4-6.

Table 4-5

BASIS FOR ESTIMATING CAPITAL CHARGES

Item	Basis
Prepaid Royalties	0.5 percent of the Plant Facilities Investment.
Organization and Startup Costs	The organization and startup costs are intended to cover operator training, equipment check-out, major changes in plant equipment, extra maintenance, and inefficient use of coal and other materials during plant startup. An allowance of 3 percent of the plant facilities investment should be made to cover organization and startup costs.
Working Capital	Working capital is the sum of the following: • Cost of a one month supply of coal at full capacity operation. • Three months of labor costs. • One month of all other operating costs (excluding coal) at full capacity operation. • A contingency of 25 percent of the total of the above three items.
Allowance for Funds During Construction (AFDC)	For a regulated utility company, the interest rate on debt (assumed to be 12.25 percent/annum for this study) is used to compute AFDC. For nonregulated companies, return on equity (assumed to be 20.00 percent/annum for this study) is used to calculate AFDC.
Land	Land costs have been estimated at $5,000/acre in mid-1980 dollars.

Table 4-6

TOTAL CAPITAL REQUIRMENT
FOR INVESTOR OWNED UTILITY PRODUCTION OF
TEXACO-BASED FUEL GAS
MID-1980 ($1000)

Steam Cycle, psig/°F/°F	1450/900/900	1450/1000/1000
Steam Generated in Gas Coolers	Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F	2,400	2,400
Sulfur Removal, %	94.6	94.6
Oxidant Plant Compressor Drivers	Motors	Motors
Size of Gasifiers, ST/day	1,375	1,375
Fuel Gas Production, 10^6 Btu/hr*	6,664	7,637
Net By-Products Power, MW*	142.40	53.30
CASE DESIGNATION	EXT-SS	EXT-SH
Plant Facilities Investment	822,608	811,701
Prepaid Royalities	4,113	4,059
Organization and Startup Costs	24,678	24,351
Working Capital	19,577	19,523
Land	950	875
AFDC**	31,044	30,623
Total Capital Requirement ($1,000)	902,970	891,132
Total Capital Requirement ($1,000/FOEB/D)#	27.70	26.76

* Production at design capacity.

** The mid-1980 AFDC was determined by the method described in the text, page 4-6 and 4-7.

\# This value was derived by dividing Total Capital Requirement (in $1,000) by the Fuel Oil Equivalent Barrel plant output per day, using a conversion factor of 5.85 x 10^8 Btu/FOEB, for fuel gas. Similarly, electricity production was converted to a FOEB/day equivalent assuming an energy value of 9,000 Btu/kWh.

Note: There will be some differences in Working Capital and AFDC for nonregulated company ownership.

- The mid-1980 AFDC allowance was then determined by subtracting the estimated constant dollar mid-1980 estimates for all other capital requirements from the Total Capital Requirement.

It is interesting to note that this procedure provides a rather low estimate for the constant dollar AFDC allowance, i.e., 3.8 percent of the plant facilities investment. If the current dollar AFDC charges are examined, it can be seen that they represent approximately 17.6 percent of the escalated plant facilities investment.

The reason for this apparent discrepancy is the following: As construction costs are paid, they are no longer subject to inflation. However, all construction costs are de-escalated from December 1989 to mid-1980, thereby creating the illusion that inflation is tending to reduce the constant mid-1980 plant facilities investment, i.e., Table 4-6 shows that for Case EXT-SS, the constant dollar mid-1980 plant facilities investment is $\$822.608 \times 10^6$.

However, if the escalated December 1989 investment of $\$1,721.440 \times 10^6$ is de-escalated at 10 percent/year for 9-1/2 years, an apparent mid-1980 investment of $\$696.084 \times 10^6$ results. This would appear to indicate that inflation is helping to reduce the constant dollar plant facilities investment. Lenders understand this problem and therefore index interest rates to inflation to handle this problem and offset the constant dollar principal loss. Therefore, the bulk of the current dollar AFDC charges is being employed to offset principal loss due to inflation. The AFDC allowance shown in Table 4-6, therefore, represents the "real" or "inflation free" interest required by the loan institution which, for this study, has been set at 2.045 percent/year.

OPERATING AND MAINTENANCE (O&M) COSTS

Operating and maintenance costs have been divided into two categories: Fixed Operating Costs (including operating labor, maintenance labor and materials, and administrative and support labor) and Variable Operating Costs (including raw water, catalyst and chemicals and ash disposal). The basis for calculating both fixed and variable O&M charges is delineated in Tables 4-7. Table 4-8 enumerates the factors applied to each plant section in order to assess the total plant maintenance cost. A summary of mid-1980 operating costs for both cases is shown in Table 4-9.

Table 4-7

BASIS FOR CALCULATING OPERATING AND MAINTENANCE COSTS

Item	Basis
Fixed Operating Costs	The fixed costs are essentially independent of the plant capacity factor and are composed of the following charges: • Operating Labor • Maintenance costs • Overhead charges These items are discussed below:
Operating Labor	The operating labor charge is computed using an average labor rate of $20.00/person hour (mid-1980$). This labor rate includes a 35 percent payroll burden.
Maintenance Costs	Annual maintenance costs are estimated as a percentage of the plant facilities investment (PFI), estimated on a section by section basis. The percentage of PFI to be used for each plant section is shown in Table 4-8. The maintenance costs are divided into maintenance labor and maintenance materials. A maintenance labor/materials ratio of 40/60 is used.
Overhead Charges	The only overhead charge to be included in the fixed costs for regulated utility producers is a charge for administrative and support (A & S) labor. This overhead charge is 30 percent of the sum of the operating and maintenance labor. For nonregulated company producers, an additional charge of 0.7 percent of the plant facilities investment should be included in the overhead. This additional cost is associated with general and administrative expenses.
Variable Operating Costs	The variable operating costs are dependent upon the plant capacity factor (CF) and are composed of the following charges: • Raw water • Catalysts and chemicals and other consumables • Ash and other waste disposal These items are discussed on the following page.

Table 4-7

BASIS FOR CALCULATING OPERATING AND MAINTENANCE COSTS
(Continued)

Item	Basis
Raw Water	The first year raw water acquisition cost is 50¢/1,000 gallons (mid-1980$). Treating costs and pumping costs are included in the operating and maintenance charges.
Catalysts and Chemicals and Other Consumables	The first year catalysts, chemicals and other consumable costs are to be determined by the contractor.
Ash and Other Waste Disposal	Solids disposal costs are to be estimated at $5.00/dry ton (mid-1980$). This charge is to be applied to nonhazardous wastes only.

Table 4-8

MAINTENANCE FACTORS

Process Unit	Maintenance % of Installed Plant Section Cost/Yr
Coal Handling	3.0
Oxidant Feed	2.0
Gasification and Ash Handling	4.5
Raw Gas Cooling	3.0
COS Hydrolysis	2.0
Acid Gas Removal	2.0
Sulfur Recovery and Tail Gas Treating	2.0
ZnO Treating	3.0
Fuel Gas Expansion	3.0
Steam, Condensate, and BFW	1.5
Steam Superheat/Reheat	2.0
Steam Turbine-Generator	1.5
General Facilities	1.5

Table 4-9

ANNUAL OPERATING AND MAINTENANCE COSTS - TEXACO-BASED FUEL GAS
INVESTOR OWNED UTILITY - MID-1980 ($1,000)

Steam Cycle, psig/°F/°F	1450/900/900	1450/1000/1000
Steam Generated in Gas Coolers	Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F	2,400	2,400
Sulfur Removal, %	94.6	94.6
Oxidant Plant Compressor Drivers	Motors	Motors
Size of Gasifiers, ST/day	1,375	1,375
Fuel Gas Production, 10^6 Btu/hr*	6,664	7,637
Net By-Products Power, MW*	142.40	53.30
CASE DESIGNATION	EXT-SS	EXT-SH
FIXED OPERATING COSTS		
Operating Labor	5,431	5,256
Maintenance Labor	8,277	8,420
Maintenance Materials	12,415	12,630
Administrative and Support Labor	4,112	4,103
Total Fixed Costs	30,235	30,408
VARIABLE OPERATING COSTS (100% Capacity Factor)		
Raw Water	1,311	1,002
Catalysts and Chemicals	1,464	1,164
Ash Disposal	1,867	1,867
Total Variable Costs	4,642	4,033

*Production at design capacity.

Operating labor requirements are a function of the number of trains. Requirements for the two plants under consideration are shown below on a per shift basis.

	Saturated Steam Base Case EXT-SS	Superheated Steam Base Case EXT-SH
Control Room Operators	5	5
Field Operators	20	19
Foremen	2	2
Lab and Instrument Technicians	4	4
	31	30

In determining labor requirements, modern computer assistance (cathode ray tube consoles) is assumed to be available.

The initial and annual catalyst and chemicals are presented in Table 4-10. The initial costs for the saturated steam base case (EXT-SS) are about 80 percent higher than those for the superheated steam base case. This is primarily due to the fuel oil requirement for the saturated steam case, a requirement which also increases the annual catalyst and chemicals costs by about 17 percent. In the saturated steam case fuel oil is required for startup of the fired steam superheater/reheater, an average of four times yearly. For each startup, eight hours is the allocated time of fuel oil firing. Other major costs for the two cases are associated with the corrosion inhibitor and the surfactant used exclusively in plant cooling water, and the chemicals required in the Texaco waste water treatment.

TABLE 4-10

CATALYST AND CHEMICAL SUMMARY - BASE CASES

ITEM	UNIT COST	TOTAL REQUIREMENTS				TOTAL COST*($1000)			
		EXT-SS		EXT-S		EXT-SS		EXT-SH	
		INITIAL	ANNUAL	INITIAL	ANNUAL	INITIAL	ANNUAL**	INITIAL	ANNUAL**
BFW DEMINERALIZER									
H_2SO_4 (93%)	$ 49.76/TON		53.3 TON		33.2 TON		2.7		1.6
NaOH (50%)	142.80/TON		5.5 TON		3.5 TON		0.8		0.5
BFW TREATING									
Na_2HPO_4	$0.22/LB		7,111 LB		4,466 LB		1.6		1.0
Hydrazine (35% Soln)	$1.00/LB		90,628 LB		56,917 LB		90.6		56.9
Morpholine	$0.86/LB		31,712 LB		19,920 LB		27.3		17.1
POLISHING									
H_2SO_4 (93%)	$ 49.76/TON		35.4 TON		20.7 TON		1.8		1.0
NaOH (50%)	142.80/TON		70.8 TON		41.5 TON		10.1		5.9
C.W. TREATING									
Lime	$42.50/TON		507.7 TON		301.2 TON		21.6		12.8
Soda Ash	$78.00/TON		604.1 TON		532.2 TON		47.1		41.5
H_2SO_4 (93%)	$49.76/TON	0.6 TON	795.6 TON	0.4 TON	636.3 TON	0.03	39.6	0.02	31.7
Corrosion Inhibitor	$ 0.80/LB	563 LB	198,600 LB	412 LB	196,100 LB	0.45	158.9	0.33	156.9
Surfactant	$ 0.60/LB	560 LB	198,600 LB	410 LB	196,100 LB	0.34	119.2	0.25	117.7
Chlorine	$135.00/TON		20.3 TON		14.8 TON		2.7		2.0
Biocide	$ 0.80/LB		22,510 LB		16,493 LB		18.0		13.2
SELEXOL® UNIT									
Solvent	$ 1.00/LB	908,660 LB	119,225 LB	908,660 LB	119,225 LB	908.7	119.2	908.7	119.2
CLAUS SULFUR PLANT									
Catalyst	$320.00/TON	153.8 TON	34.2 TON	153.8 TON	34.2 TON	49.2	10.9	49.2	10.9
STRETFORD PLANT									
Chemicals	$3504/TPD SULFUR $2.13/TON SULFUR					101.6	22.6	101.6	22.6
BEAVON UNIT									
Catalyst	$154/ft^3	1407 ft^3	187.6 ft^3	1407 ft^3	187.6 ft^3	216.7	28.9	216.7	28.9
MISCELLANEOUS OFFSITES									
Plant & Instrument Air Dryers			3,200 LB		3,200 LB		5.2		5.2
Flare	$12.22/BBL		4,500 BBL		4,500 BBL		55.0		55.0
Texaco Waste Water Treating							214.9		214.9
Fuel Oil System	$16.50/BBL	65,500 BBL	9,357 BBL			1,080.7	154.3	N/A	N/A
TOTALS, MID-1978 DOLLARS						2357.7	1153.0	1276.8	916.5
TOTALS, MID-1980 DOLLARS#						2994.3	1464.3	1621.5	1164.0

* Mid-1978 Basis, except where noted
** At 100 percent capacity
# These estimates were derived by escalating the mid-1978 costs at 27 percent.

Section 5

FINANCIAL ANALYSIS

Using the capital and operating expenses developed in the previous section, the cost of fuel gas from the two base case plant designs was determined. Cost estimating was performed by Fluor. The cost estimates thus derived were escalated by 27 percent in order to reach their equivalent value in mid-1980 dollars. This factor of 27 percent was chosen based on Fluor's experience with the escalation of similar equipment during the two year period between mid-1978 and mid-1980. All cost estimates are intended to represent those for mature technology plant designs. Because certain financial and technical parameters such as inflation rate, investment tax credits and overall plant efficiency are difficult to predict, a further objective of this section will be to analyze the sensitivity of the cost of fuel gas to changes in these parameters.

All evaluations in this section will be for the two base case plant designs, EXT-SS and EXT-SH. These designs have been described in previous sections. Briefly, the fundamental difference between these two designs is located in the high temperature gas cooling section in which saturated steam is generated in Case EXT-SS whereas superheated steam is generated in Case EXT-SH.

COST OF FUEL GAS

Calculations of the cost of fuel gas are based on the financial criteria given in Table 5-1. Costs are developed for two types of plant ownership, that of an investor owned utility and that of a nonregulated company. Those financial parameters which differ in these two cases of ownership are noted in Table 5-1. The high capacity factor used for the fuel gas plant designs reflects the high anticipated availability of the plants and the expectation that the fuel gas produced would be used to the limits that availability would allow.

The cost of fuel gas from the two base case plants is presented in Table 5-2. Costs are listed in both current dollars and mid-1980 dollars. The current

Table 5-1

FINANCIAL CRITERIA
FOR REVENUE REQUIREMENT CALCULATIONS [1]

Plant Location	• Southern Illinois
Post-1980 General Inflation Rate	• 10%/Year
Year of Plant Startup	• 1990
Design and Construction Period	• 4 Years
Project Book Life	• 30 Years for an Investor Owned Utility
	• 20 Years for a Nonregulated Company
Project Tax Life	• 13 years for Synfuels Plants
Tax Depreciation Method	• Sum-of-the-Year-Digits
Net Plant Salvage Value	• 10% of PFI
Delivered Coal Cost (Mid-1980S)	• $1.30 /$10^6$ Btu
Real Coal Price Escalation (Above General Inflation)	• 1%/Year
Property Tax Rate	• 2%/Year of Escalated PFI
Insurance Rate	• 1%/Year of Escalated PFI
Federal Income Tax Rate	• 46%
State Income Tax Rate	• 6%
Investment Tax Credit	• 10% of Escalated PFI. Normalized Over Period of Commercial Operation for Utility Ownership. Credited during construction period for non-regulated company ownership.
Project Financing	
Investor Owned Utility	
Common Equity	• 35% at 16%/Year After-Tax Return
Preferred Stock	• 15% at 12.75%/Year Dividend
Debt	• 50% at 12.25%/Year Interest
Nonregulated Company	
Common Equity	• 100% at 20%/Year After-Tax Return
Preferred Stock	• 0%
Debt	• 0%
Capacity Factor	• 90%
By-Product Electricity Credit	• 50 mills /kWh in Mid-1980S The Cost of Electricity is Allowed to Escalate at the General Inflation Rate

[1] This table is identical to Table S-2, page S-5.

dollar amounts represent the cost of fuel gas in the given year dollars, while the mid-1980 dollar amounts reflect the cost of fuel gas in the current year de-escalated at the general inflation rate of 10 percent to mid-1980. The fuel gas costs listed for nonregulated company ownership are based on a fixed requirement of 20 percent after tax return on equity.

Detailed calculations of the cost of fuel gas can be found in Appendices B and C. The eight tables in each appendix include a Capital Outlay Schedule, an Annual Capital Recovery Schedule, an Annual Revenue Requirements Schedule and an Annual Cash Flow Schedule for both investor owned utility ownership and nonregulated company ownership of the fuel gas plant designs EXT-SS and EXT-SH. For easy reference, a summary of the results from Table 5-2 follows:

	Investor Owned Utility		Nonregulated Company	
	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars
Case EXT-SS				
Levelized Cost of Fuel Gas, $\$/10^6$ Btu	3.27	19.11	5.32	24.51
Case EXT-SH				
Levelized Cost of Fuel Gas, $\$/10^6$ Btu	3.39	19.81	5.15	23.74

Three important conclusions can be drawn from these results:

- Texaco-based fuel gas produced by a regulated utility company has the potential to be considerably lower in cost than the selling price of No. 2 fuel oil.
- Utility produced fuel gas will be significantly less expensive than non-regulated company production.
- There appears to be no incentive for the development of high temperature heat recovery equipment capable of generating superheated steam for Texaco-based fuel gas plants.

A major utility application for fuel gas would be as a replacement for No. 2 fuel oil. Based on information from the Department of Energy publication, "Cost and Quality of Fuel for Electric Utility Plants," the mid-1980 cost of No. 2 fuel oil was $\$6.77/10^6$ Btu. Although there has been some softening in crude oil and

Table 5-2

FUEL GAS PRODUCTION COST AND SELLING PRICE ESTIMATES [(1)]

Case Designation	Investor Owned Utility Fuel Gas Plant				Nonregulated Company Fuel Gas Plant			
	EXT-SS		EXT-SH		EXT-SS		EXT-SH	
Steam Generated in Gas Coolers	Saturated		Superheated		Saturated		Superheated	
Net Fuel Gas Production, 10^6 Btu/Hr	6664		7637		6664		7637	
Net By-Product Power, MW*	142.40		53.30		142.40		53.30	
	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars
Total Capital Requirement For Startup in 1990, $/FOEB**/Day	68,503	27,700	66,178	26,760	68,404	27,660	66,080	26,720
Cost of Fuel Gas#, $/$10^6$ Btu								
First Year (1990)	11.63	4.27	11.56	4.25	14.48	5.32	14.01	5.15
Fifth Year (1994)	14.77	3.71	14.99	3.76	21.20	5.32	20.51	5.15
Tenth Year (1999)	20.67	3.22	21.47	3.35	34.14	5.32	33.03	5.15
Fifteenth Year (2004)	29.79	2.88	31.58	3.06	54.98	5.32	53.20	5.15
Twentieth Year (2009)	43.50	2.61	47.03	2.83	88.54	5.32	85.68	5.15
Levelized##	19.11	3.27	19.81	3.39	24.51	5.32	23.74	5.15

(1) This table is similar to Table S-3, page S-6.

* Production at 100 percent of design capacity.

** Barrels of distillate fuel oil (5.85×10^6 Btu/BBL) with higher heating value equivalent to fuel gas produced. Electricity credited at 9000 Btu/kWh.

\# End of year cost.

\## A levelized price is one which, if held constant, will yield the same return on common equity as the varying year-by-year values.

derived product prices from their mid-1980 values, the financial estimates presented in Table 5-2 indicate substantial potential for fuel cost savings based on utility ownership of mature technology Texaco-based fuel gas plants. It is critical to remember, however, that before such savings can be realized, Texaco's coal gasification technology must be demonstrated at the scale assumed as the basis for this study. It is also important to realize that the first few commercial scale Texaco-based fuel gas plants can be anticipated to cost appreciably more than the estimates presented for mature plants in this report. Therefore, it is not unreasonable to anticipate that the differential price between utility generated fuel gas and No. 2 fuel oil estimated in this study is substantially greater than that to be expected from the first few commercial fuel gas plants to be built.

The results of Table 5-2 clearly indicate that a substantial savings will be realized if a utility is producing its own fuel gas rather than buying the gas from nonregulated company producer. Because nonregulated companies require a higher return on equity and because a greater proportion of the nonregulated company capital, relative to that of an investor owned utility, is in the form of equity capital, it follows that the fuel gas cost is higher when produced by a nonregulated company.

However, unlike utilities which are limited to a certain maximum return on equity, nonregulated companies can and do maximize their returns by selling at the market price. For this reason the expected selling price of fuel gas from a nonregulated company would be the selling price, for No. 2 fuel oil ($6.77 in mid-1980 dollars). When sold at this market price, the discounted cash flow rate of return for the nonregulated producer is found to be the following:

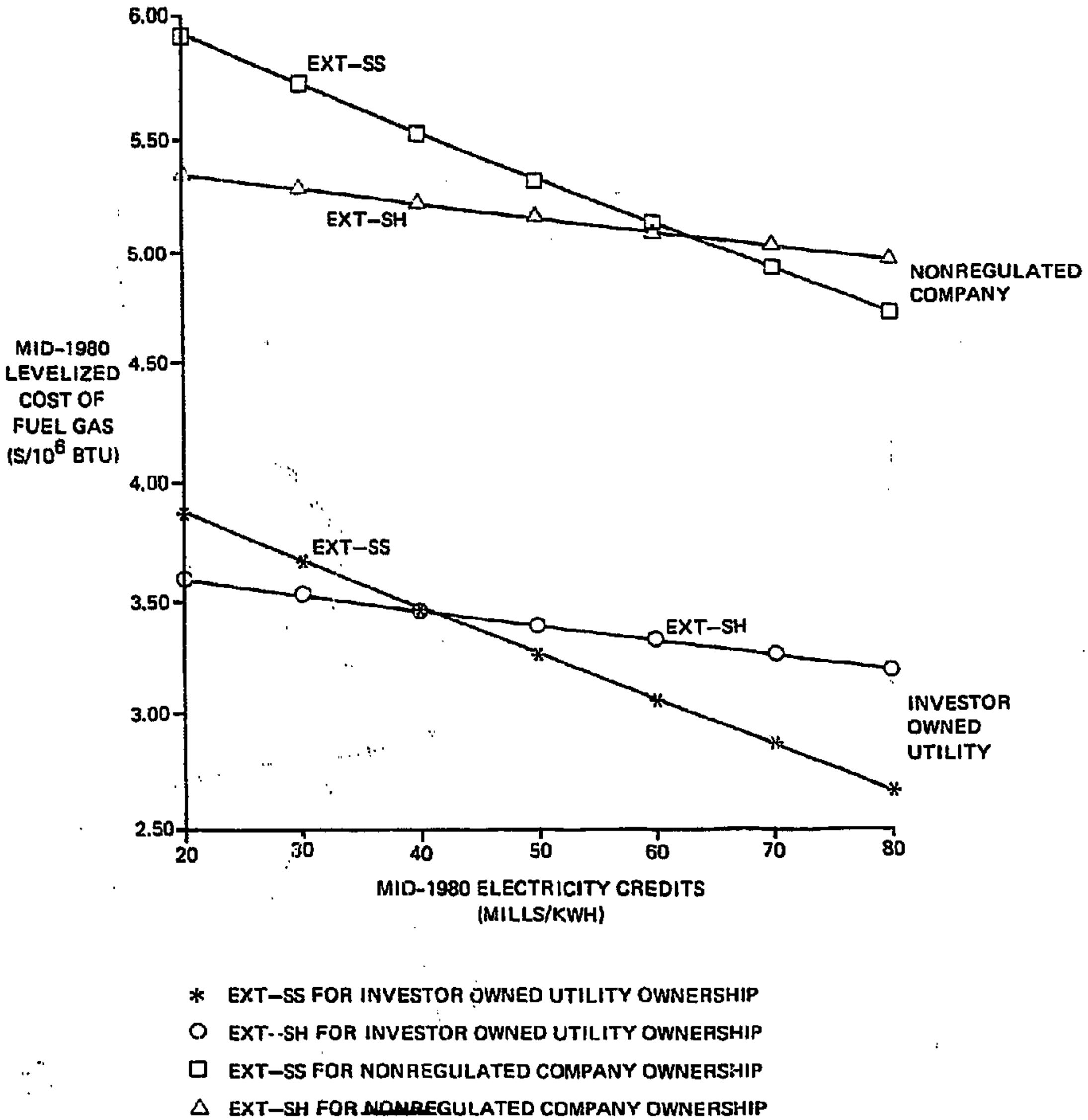
Case	Discounted Cash Flow Rate of Return on Common Equity
EXT-SS	30.41%
EXT-SH	32.55%

In order to understand why there is no economic incentive for superheating in the high temperature gas coolers, one must first know the method employed for crediting the by-product power generated by these fuel gas plants. The total revenue required is based on capital and operating costs. Revenue from the by-product electricity is then subtracted from the total revenue required to obtain a net revenue requirement. This net revenue must be recovered through the sale of the principal product, in this case, fuel gas. The by-product electricity revenue used to derive the net revenue is determined by multiplying the power generated by the electricity credit. For the results of Table 5-2 the electricity credit was assumed to be 50 mills/kWh (in mid-1980). However, when the sensitivity of the cost of fuel gas to the electricity credit is measured, an interesting phenomenon results. Figure 5-1 shows the derived cost of fuel gas when the by-product electricity is credited at rates between 20 mills/kWh and 80 mills/kWh. Since the saturated steam case (EXT-SS) generates almost three times as much electricity as does the superheated steam case (EXT-SH), its consequent cost of fuel gas is more sensitive to electricity crediting than is the fuel gas cost for the superheated case. As a result the slopes of the lines in Figure 5-1 for Case EXT-SS, both regulated and nonregulated ownership, are more negative than the slopes of the lines for Case EXT-SH. Ultimately, as the electricity credit increases for a given type of plant ownership, the lines for EXT-SS and EXT-SH intersect. The electricity credit at the point of intersection of these lines will be henceforth referred to as the "limiting by-product credit."

Figure 5-1 reveals that, for an investor owned utility, when electricity is credited at a rate greater than or equal to 41 mills/kWh (the limiting by-product credit), fuel gas produced by design EXT-SS is less expensive than that by design EXT-SH. However, for a nonregulated company with a rate of return fixed at 20 percent, electricity credits must be greater than or equal to 63 mills/kWh before fuel gas from design EXT-SS is less costly than that from design EXT-SH. Because utilities would be expected to credit the by-product electricity at a marginal rate above 41 mills/kWh in 1980, the EXT-SS design produces less costly fuel gas than the superheating case. Similarly, since a nonregulated company would probably be able to sell its by-product electricity at an avoided cost greater than 63 mills/kWh, again the saturated steam design would produce less expensive fuel gas than the superheated steam case. In conclusion, whether fuel gas is produced by an investor owned utility or a nonregulated company (allowing only a 20 percent return on equity), there appears to be no incentive for developing the high temperature heat recovery equipment capable of generating superheated steam.

FIGURE 5–1

LEVELIZED COST OF FUEL GAS AT VARYING RATES OF ELECTRICITY CREDIT

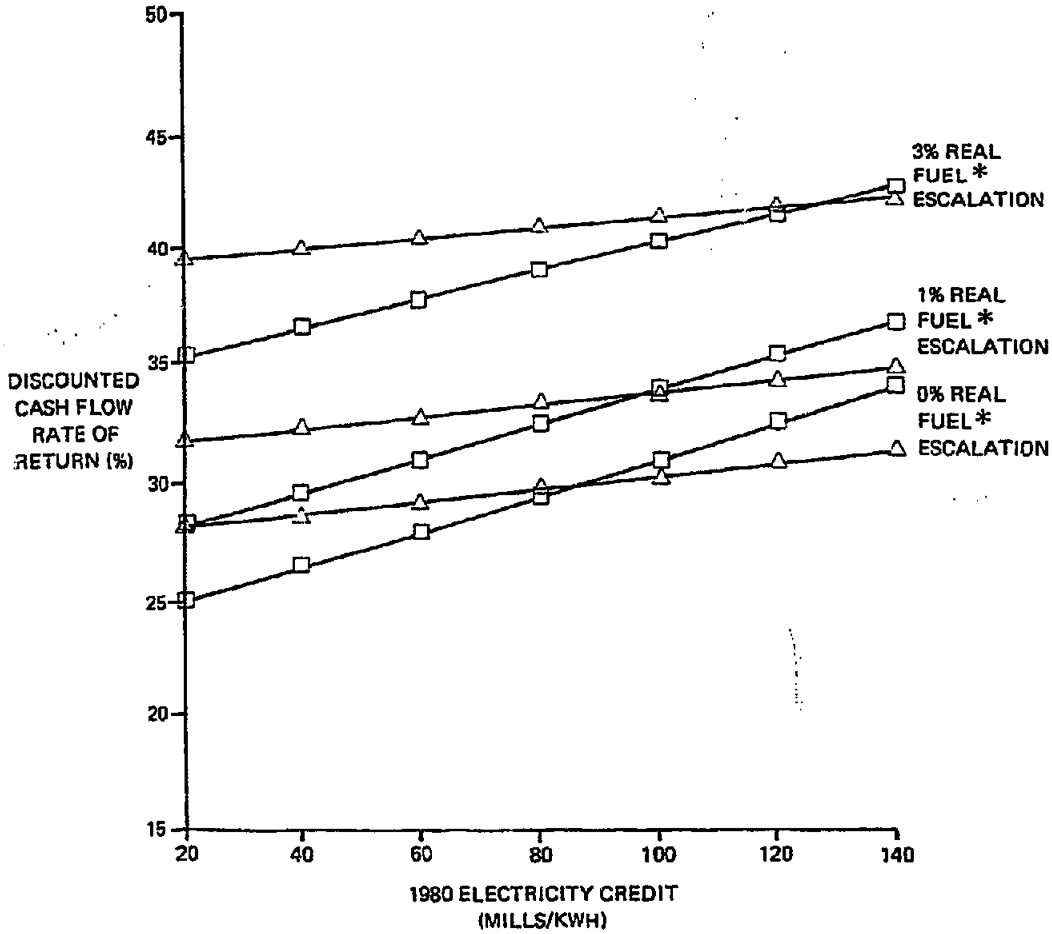


Because a nonregulated company would be expected to strive for the highest discounted cash flow (DCF) rate of return, it will be informative to consider the relative DCF returns at the various rates of by-product electricity crediting. Since the expected market price of fuel gas is higher than the cost of fuel gas derived for the case where the rate of return is fixed at 20 percent, the relative value of fuel gas to electric power increases when the nonregulated company sells fuel gas at the market. Accordingly, when this gas is sold at the market price, the electricity credit rate, above which plant design EXT-SS is more economical than EXT-SH, would be higher. This occurs because the saturated steam case generates more electricity and less fuel gas than does the superheated steam case. Figure 5-1 indicates that, at or above the electricity credit of 63 mills/kWh, saturated steam production is the best design alternative for a nonregulated company with a fixed 20 percent return on equity. However, Figure 5-2 shows that, at the same coal real escalation rate of 1 percent above inflation, an electricity credit of 98 mills/kWh is the by-product credit rate above which the saturated steam design yields the highest DCF return for a nonregulated producer selling fuel gas at the market price. Thus the results confirm the expectation that selling fuel gas at the market price increases the limiting by-product credit. Above this limiting by-product credit the most economic design choice is that of saturated steam generation in the high temperature gas coolers.

Figure 5-2 also shows the impact of different real escalation rates for fuel, both coal and fuel gas. In this analysis, electricity escalation is fixed at the general inflation rate and coal and fuel gas escalate at rates equal to or greater than the inflation rate. This diagram makes evident the conclusion that the limiting by-product credit varies in the same direction as the rate of fuel escalation. Higher rates of escalation result in higher limiting by-product credits. At the high escalation rate of 3 percent real, or 13.3 percent including inflation, the limiting by-product credit for nonregulated company ownership is 128 mills/kWh. Although it is possible that the nonregulated producer could obtain as much as 128 mills/kWh for its net by-product power, even if this were not to be the case, the small expected gain in DCF return for chosing design EXT-SH over EXT-SS at electricity credit rates below the limiting by-product credit may not offset the technical risks and costs associated with the highly developmental high temperature gas coolers which can superheat steam.

FIGURE 5–2

DISCOUNTED CASH FLOW RATE OF RETURN AS A FUNCTION OF THE ELECTRICITY CREDIT RATE GIVEN FIXED REAL RATES OF FUEL ESCALATION



□ EXT–SS NONREGULATED COMPANY OWNERSHIP

△ EXT–SH NONREGULATED COMPANY OWNERSHIP

* BOTH COAL AND FUEL GAS

In summary, there appears to be no economic advantage to superheating steam in the gas coolers in the case of investor owned utility production of fuel gas. At best, under certain economic conditions there might be a marginal advantage to superheating in the case of nonregulated company ownership.

Finally, results from these Texaco fuel gas cost estimates are compared in Table 5-3 with updated cost estimates for oxygen-blown, Lurgi-based fuel gas production from Illinois No. 6 coal. The Lurgi cost estimates were first detailed in EPRI report AF-244. These costs were updated to mid-1978 dollars in report AP-1725. With these estimates the current financial analysis of a Lurgi-based fuel gas plant was performed in the same manner as the estimates for the present Texaco-based designs. Mid-1978 dollars were updated to mid-1980 dollars using a 27 percent escalation factor and the remainder of the analysis is based on the financial factors given in Table 5-1. The ammonia by-product credit assumed is $120.00/short ton and the liquid hydrocarbons are credited both at $3.00 /10^6 Btu and at $0/10^6 Btu. It is important to realize that the original Lurgi fuel gas plant cost estimates were prepared over four years ago. It is therefore likely that these Lurgi estimates are not strictly consistent with the Texaco estimates as they do not incorporate any advances in Lurgi gasification technology that have probably resulted from the extensive experience gained at SASOL within recent years. These older Lurgi estimates have been included simply to provide a frame of reference for the newer Texaco projections. From Table 5-3 it appears that even when all three by-products of the Lurgi process (electricity, ammonia, and liquid hydrocarbons) are credited, fuel gas from a Lurgi-based plant utilizing Illinois No. 6 coal will be more costly than that from a Texaco-based plant.

ECONOMIC SENSITIVITY ANALYSIS

Because certain financial and technical parameters are difficult to project for a 1990 plant startup date, knowledge of the sensitivity of fuel gas cost to changes in certain key parameters is critical. Tables 5-4 and 5-5 illustrate the sensitivity to design and financial factors of first year and levelized fuel gas costs, respectively.

Considering the first year fuel gas cost sensitivities of Table 5-4, several interesting observations can be made:

- A reduction in the operating capacity factor from 90 to 70 percent causes an 18 to 21 percent increase in the first year cost of fuel gas.

Table 5-3

SUMMARY OF LURGI- AND TEXACO-BASED FUEL GAS

Type of Ownership	Investor Owned Utility			
Gasifier	Texaco, 1375 ST/day		Lurgi, Oxygen-Blown	
By-Products Credited	Electricity	Electricity	Electricity Ammonia Hydrocarbons	Electricity Ammonia
Case Designation	EXT-SS	EXT-SH	MX	MX
Total Capital Requirements* (Mid-1980, $1000)	902,970	891,132	887,527	887,527
Total Capital Requirement (Mid-1980, $1000/FOEB/day)**	27.70	26.76	30.99	35.65
Cost of Fuel Gas* ($/10^6 Btu)				
1st Year in Mid-1980 Dollars	4.27	4.25	5.28	5.76
1st Year in Current Dollars	11.63	11.56	14.36	15.66
Levelized in Mid-1980 Dollars	3.27	3.39	4.08	4.56
Levelized in Current Dollars	19.11	19.81	23.85	26.64

Type of Ownership	Nonregulated Ownership			
Gasifier	Texaco, 1375 ST/day		Lurgi, Oxygen-Blown	
By-Products Credited	Electricity	Electricity	Electricity Ammonia Hydrocarbons	Electricity Ammonia
Case Designation	EXT-SS	EXT-SH	MX	MX
Total Capital Requirement* (Mid-1980, $1000)	901,491	889,653	886,097	886,097
Total Capital Requirement (Mid-1980, $1000/FOEB/day)**	27.66	26.72	30.94	35.59
Cost of Fuel Gas* ($10^6 Btu)				
1st Year in Mid-1980 Dollars	5.32	5.15	6.48	6.96
1st Year in Current Dollars	14.48	14.01	17.63	18.93
Levelized in Mid-1980 Dollars	5.32	5.15	6.48	6.96
Levelized in Current Dollars	24.51	23.74	29.87	32.07

*Plant startup at beginning of 1990.

**This value was derived by dividing total capital requirement in $1000, by the Fuel Oil Equivalent plant output per day, using conversion factor of 5.85×10^6 Btu/FOEB for fuel gas. Similarly, electricity production was converted to a FOEB/Day equivalent assuming an energy value of 9,000 Btu/kWh. Hydrocarbons were credited in a like manner where noted.

- A real coal price escalation of 3 percent causes a 9 to 17 percent increase in the first year cost of fuel gas relative to that when the real coal escalation rate is taken to be 1 percent. The more substantial cost increases are associated with nonregulated company ownership.

- A 35 percent increase in the plant facilities investment causes a 17 to 21 percent increase in the cost of fuel gas.

- A lower (5%) annual inflation rate significantly reduces the first year cost of fuel gas produced by an investor owned utility, but it has little impact on the cost of nonregulated company produced gas. It is important to note that the impact of inflation on the constant dollar levelized gas cost for an investor owned utility is negligibly small.

- A 10 percent decrease in clean fuel gas efficiency results in more expensive fuel gas by up to 14 percent while a 10 percent increase in this efficiency can yield a savings of up to 11 percent.

- Governmental assistance in the form of a 25 percent Investment Tax Credit achieves approximately an 11 percent savings in the cost of fuel gas. Assistance in the form of a 75 percent loan guarantee yields a savings of between 28 and 31 percent.

- When the electricity credit is fixed at 50 mills/kWh and when a 30 percent return on common equity is required for a nonregulated owner, the most dramatic impact on fuel gas cost is attained.

- With the exception of this last change which has been analyzed in greater detail in previous discussions, all the changes in financial and technical parameters listed in Table 5-4 result in first year fuel gas costs which remain competitive with the costs of No. 2 fuel oil.

Table 5-4

FIRST YEAR (1990) FUEL GAS COST SENSITIVITY TO DESIGN AND FINANCIAL FACTORS

CONSTANT MID-1980 DOLLARS

	Investor Owned Utility Ownership Of Case EXT-SS Cost Of Fuel Gas		Investor Owned Utility Ownership Of Case EXT-SH Cost Of Fuel Gas		Nonregulated Company Ownership Of Case EXT-SS Cost Of Fuel Gas		Nonregulated Company Ownership Of Case EXT-SH Cost Of Fuel Gas	
	First Year $/10^6 Btu	% Change From Base	First Year $/10^6 Btu	% Change From Base	First Year $/10^6 Btu	% Change From Base	First Year $/10^6 Btu	% Change From Base
Base Case Results*	4.27	Base	4.25	Base	5.32	Base	5.15	Base
70% Capacity Factor Operation	5.16	+ 20.8	5.02	+ 18.1	6.46	+ 21.4	6.14	+ 19.2
3% Escalation in Real Coal Price[δ]	4.73	+ 10.8	4.65	+ 9.4	6.24	+ 17.3	5.95	+ 15.5
35% Increase in Plant Facilities Investment	5.12	+ 19.9	4.98	+ 17.2	6.44	+ 21.1	6.12	+ 18.8
5% Annual Inflation Rate[≠]	3.20	- 25.1	3.34	- 21.4	5.25	- 1.3	5.11	- 0.8
Clean Fuel Gas Efficiency Decrease by 10%[#]	4.86	+ 13.8	4.76	+ 12.0	6.03	+ 13.3	5.76	+ 11.8
Clean Fuel Gas Efficiency Increase by 10%[#]	3.79	- 11.2	3.83	- 9.9	4.75	- 10.7	4.66	- 9.5
25% Investment Tax Credit	--		--		4.73	- 11.1	4.65	- 9.7
Loan Guarantee (75% Debt at 12.25%/Year)	--		--		3.65	- 31.4	3.71	- 28.0
30% Return on Common Equity	--		--		8.80	+ 65.4	8.15	+ 58.3

See Table 5-6 for explanatory notes.

Table 5-5

LEVELIZED FUEL GAS COST SENSITIVITY TO DESIGN AND FINANCIAL FACTORS

CONSTANT MID-1980 DOLLARS

	Investor Owned Utility Ownership Of Case EXT-SS Cost Of Fuel Gas		Investor Owned Utility Ownership Of Case EXT-SH Cost Of Fuel Gas		Nonregulated Company Ownership Of Case EXT-SS Cost Of Fuel Gas		Nonregulated Company Ownership Of Case EXT-SH Cost Of Fuel Gas	
	First Year $\$/10^6$ Btu	% Change From Base	First Year $\$/10^6$ Btu	% Change From Base	First Year $\$/10^6$ Btu	% Change From Base	First Year $\$/10^6$ Btu	% Change From Base
Base Case Results*	3.27	Base	3.39	Base	5.32	Base	5.15	Base
70% Capacity Factor Operation	3.80	+ 16.2	3.84	+ 13.3	6.46	+ 21.4	6.14	+ 19.2
3% Escalation in Real Coal Price[δ]	4.55	+ 39.1	4.50	+ 32.7	6.24	+ 17.3	5.95	+ 15.5
35% Increase in Plant Facilities Investment	3.70	+ 13.1	3.76	+ 10.9	6.44	+ 21.1	6.12	+ 18.8
5% Annual Inflation Rate≠	3.28	+ 0.3	3.41	+ 0.6	5.25	- 1.3	5.11	- 0.8
Clean Fuel Gas Efficiency Decrease by 10%[#]	3.74	+ 14.4	3.80	+ 12.1	6.03	+ 13.3	5.76	+ 11.8
Clean Fuel Gas Efficiency Increase by 10%[#]	2.88	- 11.9	3.05	- 10.0	4.75	- 10.7	4.66	- 9.5
25% Investment Tax Credit	--		--		4.73	- 11.1	4.65	- 9.7
Loan Guarantee (75% Debt at 12.25%/Year)	--		--		3.65	- 31.4	3.71	- 28.0
30% Return on Common Equity	--		--		8.80	+ 65.4	8.15	+ 58.3

See Table 5-6 for explanatory notes.

Table 5-6

NOTES FOR TABLES 5-4 AND 5-5

*The Base Case financial parameters used are given in Table 5-1.

$^{\delta}$A 3 percent inflation free coal price escalation rate is equivalent to a 13.3 percent actual escalation rate when the general inflation rate is 10 percent/ year.

≠For this 5 percent annual inflation rate case, the following financial parameters were used:

	Investor Owned Utility	Nonregulated Company
Annual Return on Common Equity	10.73%	14.55%
Annual Preferred Stock Dividends	7.63%	--
Annual Interest on Debt	7.15%	--
Annual Coal Price Escalation	6.05%	6.05%

#The clean fuel gas efficiencies used were the following:

	Case EXT-SS	Case EXT-SH
Clean Fuel Gas Efficiency When Increased by 10%	71.14%	81.53%
Clean Fuel Gas Efficiency When Decreased by 10%	58.20%	66.71%

(Clean Fuel Gas Efficiency is the exported clean gas HHV x 100/coal feed HHV)

Section 6

DESIGN SENSITIVITY STUDIES

GENERAL

The primary objective of this study has been both to assess the cost of intermediate-Btu fuel gas produced by a Texaco-based coal gasification plant and to determine whether a savings in the cost of fuel gas could be realized by a design which provides for superheating of steam in the raw gas cooling section. However, four secondary objectives have been addressed in this study, and some interesting conclusions can be drawn from the results.

Each sensitivity study is composed of the evaluation of plant designs which process the same quantity of coal, but which differ from the base case designs in one or more ways. Each was performed with the same degree of accuracy as were the base cases. Results from these analyses are tabulated at the end of this section. The tables include summaries of operating results, summaries of power consumptions, energy balances, and lists of plant facilities investments. Also, included at the end of this section are summaries of total capital requirements, annual operating and maintenance costs, catalyst and chemical costs, and overall block flow diagrams. Two base cases and twelve substudy cases, or fourteen cases in total, were evaluated with the objective of assessing the impact of the four design changes on major performance parameters and on the cost of fuel gas.

The design sensitivities analyzed are:

- The use of a gas recycle to quench the hot gasifier effluent instead of the use of a radiant gas cooler
- Variations in the degree of sulfur removal
- The use of larger capacity gasifiers
- Variations in steam cycle temperature and pressure

The rationale behind each of these design changes, except that regarding the extent of sulfur removal, is that because of their association with expensive equipment items, their optimization has potential for reducing the cost of fuel gas produced by the resulting modified plant designs. The studies evaluating variations in the depth of sulfur removal allow for an understanding of the costs associated with meeting different levels of environmental control requirements.

Tables 6-1 and 6-2 summarize the results of these substudy cases for both investor owned utility and nonregulated company ownership of the fuel gas plants. The base cases which have been discussed throughout this report have been included for comparison purposes. The table headings for each of the fourteen designs indicate the major design changes associated with each plant study. The reader may find these tables a convenient reference for distinguishing the major characteristics of each design.

GAS RECYLE DESIGN OPTION

The raw gas leaves the gasifier at a temperature in the neighborhood of 2400°F. It is difficult to recover heat from such high temperature, ash laden gas in convective type heat transfer equipment due to the "stickiness" and fouling characteristics of the ash at these temperatures. In the base case designs considered for this study, the hot gas is first cooled in radiant type heat exchangers to a temperature of approximately 1500°F, at which point it is possible to continue cooling the gas in convective type equipment. This radiant/convective heat transfer design represents the system that is currently being tested at the 150 ton/day scale in the Ruhrchemie plant as well as the design that has been chosen for the 1000 ton/day Cool Water plant. The radiant boiler used in these designs is an extremely large and costly vessel. It is therefore possible that if the radiant boiler could be eliminated, the plant capital cost could be reduced which would in turn lead to a reduction in the cost of fuel gas. One way to eliminate the radiant unit would be to recycle cold fuel gas from the particulate scrubber to quench the high temperature gasifier effluent gas to 1500°F. The mixed gases could then be further cooled in convective type heat transfer equipment.

The saturated steam base case (EXT-SS) and the saturated steam case with gas recycle (EXT-SS1), shown schematically in the overall block flow diagrams EXT-SS and EXT-SS1, represent the two alternative approaches to gas cooling. Both

Table 6-1

PRODUCTION COSTS AND SELLING PRICE ESTIMATES

FOR TEXACO-BASED FUEL GAS [1]

(INVESTOR OWNED UTILITY, MID-1980 DOLLARS)

	Base Case						
Steam Cycle, psig/°F/°F	1450/900/900	1450/900/900	1450/900/900	736/900	1450/800/800	1450/1000/1000	1450/900/900
Steam Generated in Gas Coolers	←	←	←	Saturated	→	→	→
Gas Temperature Entering Heat Recovery, °F	←	2400	←	1500	→	→	2400
Sulfur Removal, %	←	←	←	94.6	→	→	+99.9¶
Oxidant Plant Compressor Drivers	←	←	←	Motors	→	→	→
Nominal Capacity of Gasifiers, ST/day	1375	2200	←	1375	→	→	→
Fuel Gas Production, 10^6 Btu/hr#	6664	6664	6802	7163	6942	6658	6626
Net By-Product Power, MW#	142.40	142.40	105.96	58.28	88.92	125.32	136.61
Case Designation	EXT-SS	EXT-SS*	EXT-SS1	EXT-SS2	EXT-SS3	EXT-SS4	EXT-SS5
Total Capital Requirement for 1990 Startup ($1000)	902,970	835,959	860,481	830,730	849,053	880,614	923,423
Total Capital Requirement ($1000/FOEB/day##)	27.70	25.65	27.04	26.34	26.73	27.57	28.65
Fuel Gas Price When Electricity is Credited at 50 Mills/kWh δ							
First Year (1990) $/10^6$ Btu	4.27	4.04	4.32	4.33	4.31	4.33	4.43
Twentieth Year (2009) $/10^6$ Btu	2.61	2.53	2.79	2.94	2.83	2.72	2.71
Levelized $/10^6$ Btu	3.27	3.12	3.39	3.49	3.41	3.36	3.39

	Base Case						
Steam Cycle, psig/°F/°F	1450/1000/1000	1450/1000/1000	1450/900/900	736/900	736/900	1450/1000/1000	1450/1000/1000
Steam Generated in Gas Coolers	←	←	←	Superheated	→	→	→
Gas Temperature Entering Heat Recovery, °F	2400	→	1500	→	2400	→	→
Sulfur Removal, %	←	←	94.6	→	→	83.6	+99.9¶
Oxidant Plant Compressor Drivers	←	←	←	Motors	Steam Turbines	Motors	→
Nominal Capacity of Gasifiers, ST/day	1375	2200	1375	→	→	→	→
Fuel Gas Production, 10^6 Btu/hr#	7637	7637	7402	7637	7637	7668	7621
Net By-Product Power, MW#	53.30	53.30	46.82	15.84	36.57	54.73	47.20
Case Designation	EXT-SH	EXT-SH*	EXT-SH1	EXT-SH2	EXT-SH3	EXT-SH4	EXT-SH5
Total Capital Requirement for 1990 Startup ($1000)	891,132	819,814	800,543	811,730	832,277	881,790	902,610
Total Capital Requirement ($1000/FOEB/day##)	26.76	24.62	24.94	25.43	25.47	26.34	27.34
Fuel Gas Price When Electricity is Credited at 50 Mills/kWh δ							
First Year (1990) $/10^6$ Btu	4.25	4.03	4.17	4.26	4.19	4.20	4.35
Twentieth Year (2009) $/10^6$ Btu	2.83	2.75	2.89	3.00	2.88	2.80	2.90
Levelized $/10^6$ Btu	3.39	3.25	3.39	3.49	3.40	3.35	3.47

(1) This table is identical to Table S-6, page S-12.

* Case EXT-SS or EXT-SH with 2200 ton/day gasifiers.

¶ One ppm total sulfur in the product gas on a mole basis.

\# Production at design capacity.

\## This value was derived by dividing Total Capital Requirement (in $1000) by the Fuel Oil Equivalent Barrel output per day. Using a conversion factor of 5.85×10^6 Btu/FOEB for fuel gas. Similarly, electricity production was converted to a FOEB/day equivalent assuming an energy value of 9,000 Btu/kWh.

δ Electricity is credited at 50 mills/kWh in Mid-1980 Dollars and is allowed to escalate at the general inflation rate.

Table 6-2

PRODUCTION COSTS AND SELLING PRICE ESTIMATES
FOR TEXACO-BASED FUEL GAS
(NONREGULATED COMPANY, MID-1980 DOLLARS)

	Base Case						
Steam Cycle, psig/°F/°F	1450/900/900	1450/900/900	1450/900/900	736/900	1450/800/800	1450/1000/1000	1450/900/900
Steam Generated in Gas Coolers	←	←	←	Saturated	→	→	→
Gas Temperature Entering Heat Recovery, °F	←	2400	←	1500	→	→	2400
Sulfur Removal, %	←	←	←	94.6	→	→	+99.9¶
Oxidant Plant Compressor Drivers	←	←	←	Motors	→	→	→
Nominal Capacity of Gasifiers, ST/day	1375	2200	←	1375	→	→	→
Fuel Gas Production, 10^6 Btu/hr#	6664	6664	6802	7163	6942	6658	6626
Net By-Product Power, MW#	142.40	142.40	105.96	58.28	88.92	125.32	136.61
Case Designation	EXT-SS	EXT-SS*	EXT-SS1	EXT-SS2	EXT-SS3	EXT-SS4	EXT-SS5
Total Capital Requirement for 1990 Startup ($1000)	901,491	834,613	859,089	829,398	847,685	879,179	921,904
Total Capital Requirement ($1000/FOEB/day##)	27.66	25.60	27.00	26.30	26.69	27.52	28.61
Fuel Gas Price When Electricity is Credited at 50 Mills/kWh δ First Year (1990) $/$10^6$ Btu	5.32	5.02	5.30	5.24	5.26	5.36	5.50

	Base Case						
Steam Cycle, psig/°F/°F	1450/1000/1000	1450/1000/1000	1450/900/900	736/900	736/900	1450/1000/1000	1450/1000/1000
Steam Generated in Gas Coolers	←	←	←	Superheated	→	→	→
Gas Temperature Entering Heat Recovery, °F	2400	→	1500	→	2400	→	→
Sulfur Removal, %	←	←	94.6	→	→	83.6	+99.9
Oxidant Plant Compressor Drivers	←	←	←	Motors	Steam Turbines	Motors	→
Nominal Capacity of Gasifiers, ST/day	1375	2200	1375	→	→	→	→
Fuel Gas Production, 10^6 Btu/hr#	7637	7637	7402	7637	7637	7668	7621
Net By-Product Power, MW#	53.30	53.30	46.82	15.84	36.57	54.73	47.20
Case Designation	EXT-SH	EXT-SH*	EXT-SH1	EXT-SH2	EXT-SH3	EXT-SH4	EXT-SH5
Total Capital Requirement for 1990 Startup ($1000)	889,653	818,476	799,263	810,411	830,917	880,330	901,108
Total Capital Requirement ($1000/FOEB/day##)	26.72	24.58	24.90	25.39	25.42	26.29	27.30
Fuel Gas Price When Electricity is Credited at 50 Mills/kWh δ First Year (1990) $/$10^6$ Btu	5.15	4.87	5.02	5.10	5.04	5.09	5.26

*Case EXT-SS or EXT-SH with 2200 ton/day gasifiers.
¶One ppm total sulfur in the product gas on a mole basis.
#Production at design capacity.
##This value was derived by dividing Total Capital Requirement (in $1000) by the Fuel Oil Equivalent Barrel output per day. Using a conversion factor of 5.85 x 10^6 Btu/FOEB for fuel gas. Similarly, electricity production was converted to a FOEB/day equivalent assuming an energy value of 9,000 Btu/kWh.
δElectricity is credited at 50 mills/kWh in Mid-1980 Dollars and is allowed to escalate at the general inflation rate.

cases process 11,000 tons/day of Illinois No. 6 coal. They have identical steam cycles, and the degree of sulfur removal is 94.6 percent in each case.

The option of using a gas recycle to quench the gasifier effluent reduces not only the quality but also the quantity of thermal energy recovered. With no quench gas recycle, Case EXT-SS has thermal energy available for recovery at a higher temperatures (the gas enters the heat recovery section at 2400°F) and produces 24 MW more power than the corresponding recycle Case EXT-SS1, in which gas enters the heat recovery section at the lower temperature of 1500°F. In the recycle case, steam generation is reduced to avoid the lower fuel gas reheater approach temperatures that would be required in order to transfer the same quantity of heat as is transferred in the nonrecycle design. Thus less heat is recovered, and the rejection of more heat to an air-cooled exchanger is a necessary consequence. Table 6-1 indicates that the fuel gas production in the EXT-SS case is 6664×10^6 Btu/hr and is slightly lower than the 6802×10^6 Btu/hr for the EXT-SS1 case. This result follows from the fact that EXT-SS produces more high-pressure steam and, therefore, more fuel gas is required in the fired heater for superheating and reheating. The energy recovery efficiency for the EXT-SS and EXT-SS1 cases is 77.11 and 75.27 percent respectively. As shown in Table 6-4, the power consumption summary, the major difference in power requirements is in the gas cooling section. For the EXT-SS case it is 941 kW, whereas for the EXT-SS1 case it is 6592 kW. This is due to the power requirements of the recycle gas compressors.

Tables 6-1 and 6-2 give a summary of the economic results for the gas recycle design option. By eliminating the radiant heat recovery unit, mid-1980 capital cost savings of 43 million dollars would be realized by an investor owned utility. However, the reduction in capital cost and the slight increase in net fuel gas output together do not offset the cost associated with both the reduction in efficiency and the increased power consumption of the quench gas recycle design. Consequently, the net impact on the mid-1980 levelized cost of fuel gas from an investor owned utility is a $\$0.12/10^6$ Btu increase when the recycle option is employed. This increase is less than 4 percent of the fuel gas cost, and given the accuracy of the study, this differential is very small. This is a most encouraging result as it indicates the existance of an alternate approach to high temperature gas cooling to the radiant/convective configuration currently being considered. If scale-up of this approach is not successful, the alternate recycle configuration would be applied at essentially no cost penalty.

SULFUR REMOVAL DESIGN OPTIONS

In a Texaco-based GCC system, the cost of the acid gas removal, sulfur recovery and tail gas treating units is a relatively small fraction (approximately four percent) of the total plant investment. These same units constitute a much larger fraction (six to nine percent) of the total plant investment for a clean fuel gas plant. The impact of increased sulfur removal standards on fuel gas plant economics could be significant in terms of final product cost. Hence, sub-studies evaluating the impact of the depth of sulfur removal were performed.

The block flow diagrams EXT-SS and EXT-SS5 depict schematically the high-pressure saturated steam cases with 94.6 percent and +99.9 percent* sulfur removal respectively. The block flow diagrams EXT-SH, EXT-SH4, and EXT-SH5 represent the schematic block flow diagrams for the high-pressure superheated steam cases, with 94.6 percent, 83.6 percent, and +99.9 percent* sulfur removal respectively. The deep sulfur removal cases corresponding to 1 ppm total sulfur on a volume basis in the fuel gas require a COS hydrolysis step and zinc oxide treatment.

<u>COS Hydrolysis</u>. The substudy cases EXT-SS5 and EXT-SH5 are the only cases that include a COS hydrolysis step. The following overall reaction occurs in the presence of Haldor Topsoe, Inc. catalyst Type CKA:

$$COS + H_2O = CO_2 + H_2S \qquad (6\text{-}1)$$

Shifting the reaction equilibria to the right is favored by low temperatures and a high ratio of steam to dry gas. The operating temperature is maintained 50°F above the dew point of the gas to avoid pore condensation. This is accomplished by reheating the saturated gas leaving the particulate scrubber from 320°F to 370°F. COS concentration is reduced to 6.5 ppmv in the gas leaving the catalyst bed.

<u>Acid Gas Removal</u>. All cases utilize the Selexol® process for acid gas removal. In Cases EXT-SS5 and EXT-SH5 where sulfur is removed to the extent that only 5 ppmv sulfur on a mole basis remain in the gas leaving the acid gas removal unit, use of the Selexol® unit without the COS hydrolysis step would only be

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*Corresponding to 1 ppm total sulfur in the product fuel gas possible through the use of uneconomical high solvent circulation and steam rates. In these cases, one ppm (mole basis) specification on the clean gas is made possible by passing reheated gas at 600°F through ZnO beds.

In general, acid gas removal processes tend to absorb both the hydrogen sulfide (H_2S) and the carbon dioxide (CO_2) present in the feed gas. Absorption of all of the CO_2 would increase the required solvent circulation rates, equipment sizes and energy demands. The design and size of the downstream sulfur recovery units are affected adversely due to the decrease in H_2S concentration in the acid gas. However, the Selexol® process removes H_2S in preference to CO_2. The process has been selected for use in these cases because it accomplishes selective H_2S removal and its economics compare favorably with other similar processes. In the Selexol® process applied here, the H_2S concentration is over 20 percent in the acid gas feed to the sulfur recovery unit. At H_2S concentrations in this range, a "split-flow" sulfur plant design is employed to avoid the use of fuel gas in the sulfur furnace. In the "spilt-flow" design, the flame in the sulfur furnace can be sustained by burning only acid gas.

Zinc Oxide Treatment. The Deep Sulfur Removal Cases (EXT-SS5 and EXT-SH5) with one ppm sulfur in the fuel gas are the only cases that include a ZnO treating unit. This involves passing the reheated fuel gas at 600°F from which most of the sulfur is removed through beds of ZnO (Type HTZ-3 by Haldor Topsoe, Inc.). Both COS and H_2S are removed forming ZnS. Sizing of the units provides a one-year bed life, after which the spent bed is replaced with fresh ZnO.

The increased CO_2 absorption with greater depths of sulfur removal results in slightly reduced gas flow to the gas expander. Table 6-3, Summary of Operating Results, shows that less power production in the expander is a consequence. The COS hydrolysis process and the ZnO unit together produce additional pressure drops which further contribute to the reduced power generation by way of decreasing the inlet working pressure at the gas expander. Greater depths of sulfur removal are also associated with increased steam demands in the Selexol® unit, and as a result, less power is generated in the turbogenerator. Net system powers for the EXT-SS and EXT-SS5 cases are 142 and 137 MW respectively and for the EXT-SH, EXT-SH4 and EXT-SH5 cases are 53, 55 and 47 MW respectively.

The fuel gas production rates for the EXT-SS and EXT-SS5 cases are 6664 x 10^6 Btu/hr and 6626 x 10^6 Btu/hr respectively with the energy recovery efficiencies at 77.11 and 76.24 percent for the two respective cases. For the EXT-SH, EXT-SH4, and EXT-SH5 cases, the fuel gas production rates are 7637 x 10^6 Btu/hr, 7668 x 10^6 Btu/hr and 7621 x 10^6 Btu/hr respectively with energy efficiencies of [illegible] 77, 79.20, and 78.08 percent for the three respective cases. As the depth of sulfur removal is increased, the fuel gas available for export is decreased slightly due to increased gas absorption in the Selexol® unit.

The initial and annual catalyst and chemicals for Cases EXT-SS5, EXT-SH4, and EXT-SH5 are presented in Table 6-34. Comparing these to the costs presented in Table 4-10 for the base cases, it may be seen that the costs associated with the Selexol® unit, Claus Sulfur Plant, and Stretford Plant increase as the level of sulfur removal is increased.

The mid-1980 capital cost associated with increasing sulfur removal from 94.6 percent to greater than 99.9 percent is 20.5 million dollars in cases where saturated steam is generated in the raw gas coolers. A smaller cost of 11.5 million is associated with the same change in extent of sulfur removal in the cases where superheated steam is generated in the gas coolers. The net impact of increased capital costs and decreased product and by-product output is an increase in investor owned utility constant dollar levelized cost of fuel gas from \$3.27/10^6 Btu (EXT-SS) to \$3.39/10^6 Btu (EXT-SS5). Table 6-1 reveals that a similar increase in depth of sulfur removal results in an increased constant dollar levelized cost of fuel gas as produced from a superheat design by \$.08/10^6 Btu. Figure 6-1 displays, in graphical form, the levelized cost of fuel gas as a function of the depth of sulfur removal. Discontinuities are shown in the plot since the COS hydrolysis step is included at the +99.9 percent level and not at the 94.6 percent level.

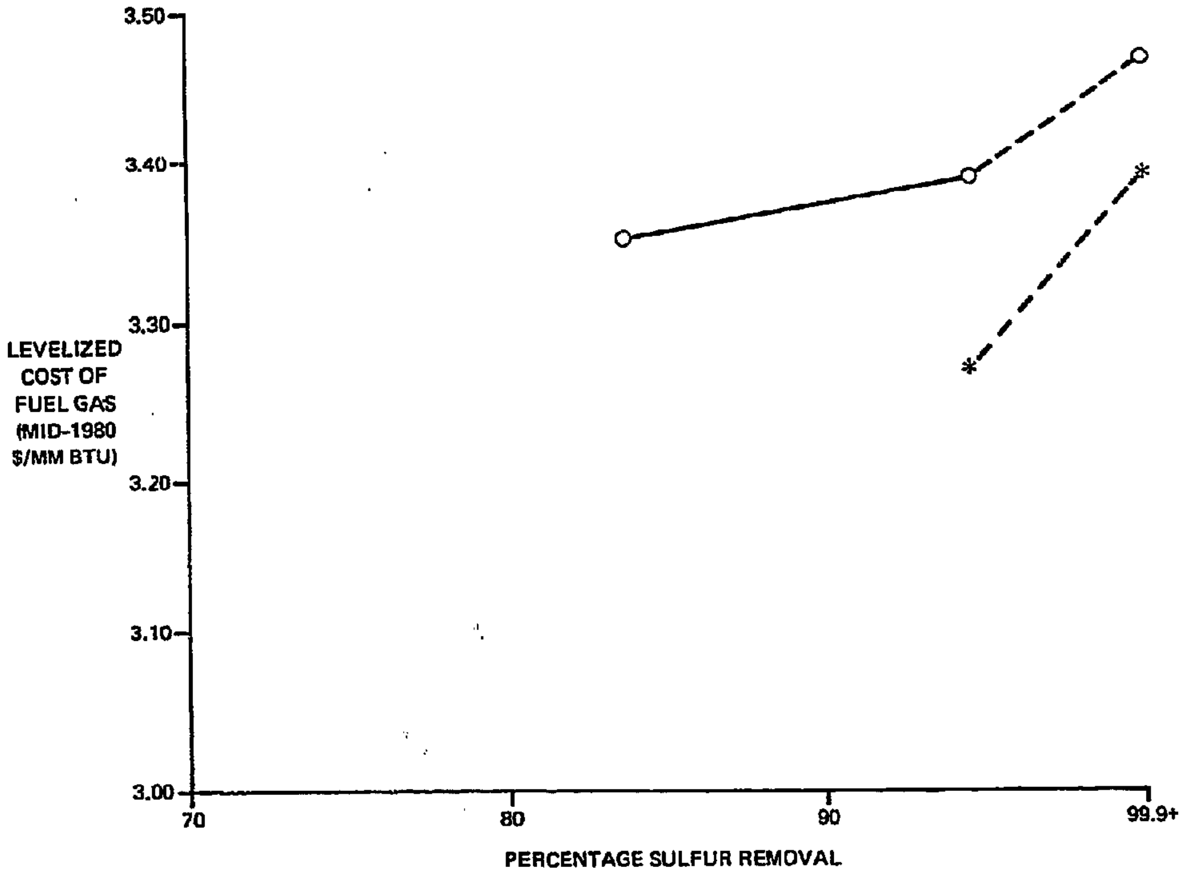
As a percent of gas cost, almost complete sulfur removal costs only 3.7 percent more than does 94.6 percent sulfur removal.

GASIFIER CAPACITY DESIGN OPTION

The most expensive sections in a fuel gas plant apart from the oxidant feed system are the gasification/gas cooling systems. Decreases in capital requirement and cost of fuel gas may be realized through design changes in these two units. One of the options available is to increase the individual capacities of trains (reducing the number of trains) and to thus realize economies of scale.

FIGURE 6–1

SENSITIVITY OF FUEL GAS COSTS TO PERCENTAGE OF SULFUR REMOVAL



INVESTOR OWNED UTILITY PRODUCTION

* SATURATED STEAM FUEL GAS PLANT DESIGN WITH A 1450 PSIG/900°F/900°F STEAM CYCLE, NO GAS RECYCLE AND 1375 TON/DAY CAPACITY GASIFIERS.

○ SUPERHEATED STEAM FUEL GAS PLANT DESIGN WITH A 1450 PSIG/1000°F/1000°F STEAM CYCLE, NO GAS RECYCLE AND 1375 TON/DAY CAPACITY GASIFIERS.

Two gasifier sizes were studied, one having a nominal coal processing capacity of 1375 tons/day and the other with a nominal capacity of 2200 tons/day. The smaller capacity gasifier used for the base case designs and cost estimates presented in this report is representative of the first generation commercial-scale Texaco gasifiers to be employed in the Cool Water demonstration plant. The larger, 2200 ton/day gasifiers represent a potential future extension of the Texaco technology after the smaller first generation systems have been successfully demonstrated. The objective of this substudy was to estimate the economic incentives to be associated with development of larger capacity Texaco gasifiers.

With the larger gasifiers, the number of gasification and gas cooling trains were reduced from eight operating and two spare to five operating and one spare, or from 25 to 20 percent spare capacity. Table 6-3 shows that the system performance remains essentially the same when larger gasifiers are used. However, the overall system availability would be somewhat reduced due to the reduction in spare capacity.

The total capital requirement for the four cases (EXT-SS, EXT-SS*, EXT-SH, and EXT-SH*) are presented in Table 6-1. As expected, the gasification and gas cooling unit mid-1980 costs are lower for the 2200 ton/day capacity gasifier cases. The effect of the larger gasifiers on the Total Plant Facilities Investment is a decrease of 7 to 8 percent. Detailed Capital Requirements and Operating and Maintenance Costs are shown in Tables 6-32 and 6-33. The 30-year levelized costs of fuel gas in mid-1980 dollars are $\$3.27/10^6$ Btu and $\$3.39/10^6$ Btu for the base cases EXT-SS and EXT-SH. These costs for the corresponding cases with the larger gasifiers are $\$3.12/10^6$ Btu and $\$3.25/10^6$ Btu respectively. Therefore, a small but significant savings could be realized through the development of these larger scale gasifiers.

STEAM CYCLE DESIGN OPTIONS

The oxygen-blown Texaco coal gasification process has significant steam raising capability in the raw gas cooling section. This allows for the development of Texaco-based clean fuel gas plant designs which can export as much as 150 MW of by-product electric power. Choices of steam cycle operating conditions can significantly impact the quantity of this by-product power generation which, together with an associated change in fuel gas output, has the potential for reducing the cost of gas produced.

Steam Pressure. The two steam cycle pressure options studied here include the base case design pressure of 1450 psig and an alternate design of a pressure of 736 psig. The low-pressure designs produce less by-product power and more fuel gas. As there is no reheat in the low-pressure case, power production will be reduced, relative to the high-pressure reheat design. Futhermore, the absence of a reheat results in less internal consumption of the product fuel gas. Material costs are also reduced when equipment operates at lower pressure. Therefore, three major changes result from reducing the steam cycle pressure from the base case condition (1450 psig) to 736 psig. By-product power generation is reduced; fuel gas production is increased; and material costs are lowered.

Results from two sets of designs allow us to evaluate the net impact of these changes on the costs of fuel gas. (Flow diagrams for these designs can be found under the appropriate names at the end of this section.) Table 6-1 shows that designs EXT-SS1 and EXT-SS2 are identical in all operating parameters except steam cycle conditions. While Case EXT-SS1 has a cycle which operates at 1450 psig and superheats and reheats the steam to 900°F, Case EXT-SS2 operates a steam cycle at 736 psig and superheats to 900°F but does not reheat. As anticipated, the low-pressure, non-reheat case generates less power by 47.68 MW, and more fuel gas, by 361 x 10^6 Btu/hr. The low-pressure design requires a total capital outlay that is almost 30 million (mid-1980) dollars less than that for the high-pressure case.

Cases EXT-SH1 and EXT-SH2 allow for the evaluation of a low-pressure steam cycle in designs which superheat in the raw gas cooling section. Again, Table 6-1 shows that all operating parameters, except steam cycle pressure, are identical. Just as in the previous comparison, the low-pressure design, EXT-SH2, results in reduced by-product power, increased fuel gas production and a decreased total capital requirement.

Once again assessment of the economic incentives to design Texaco-based fuel gas plants with high-pressure, reheat steam cycles is complicated by the assignment of a value to the by-product electric power. All financial results shown in Table 6-1 for fuel gas costs generated by regulated utilities have been calculated on the basis of an electricity value of 50 mills/kWh in mid-1980 dollars. The following table summarizes pertinent fuel gas costs from Table 6-1 and presents equivalent fuel gas costs assuming electricity were to be credited at 100 mills/kWh in mid-1980 dollars.

Fuel Gas Production Costs For An
Investor Owned Utility, Constant Mid-1980 Dollars

Case Designation	EXT-SS1	EXT-SS2	EXT-SH1	EXT-SH2
Steam Generated in Gas Coolers	Saturated	Saturated	Superheated	Superheated
Steam Cycle, psig/°F/°F	1450/900/900	736/900	1450/900/900	736/900
Fuel Gas Production, 10^6 Btu/hr	6802	7163	7402	7637
Net By-product Power, MW	105.96	58.28	46.82	15.84
Levelized Fuel Gas Cost, \$10^6 BTU				
Electricity Credited at 50 mills/kWh	3.39	3.49	3.39	3.49
Electricity Credited at 100 mills/kWh	2.64	3.10	3.09	3.39

It is evident from the above results that at an electricity credit of 50 mills/kWh, the choice of steam cycle conditions has minor impact (~ 3%) on the cost of fuel gas. However, if the electricity credit is increased to 100 mills/kWh, going to the high-pressure, reheat steam cycle will have major impact (10 to 17%) on the cost of fuel gas produced.

Steam Superheat/Reheat Temperature. Three superheat and reheat temperatures with 1450 psig steam cycle designs are evaluated here. Higher steam temperatures increase the thermodynamic efficiency of the cycle. They also necessitate increased fuel gas consumption in the fired heater where superheating and reheating of steam occurs in designs with saturated steam production in the raw gas coolers. In summary, increasing steam cycle temperature yields increased power production and decreased net fuel gas output. Increasing temperature also increases demands on materials, and therefore total capital requirement changes in the same direction as the temperature.

The three temperatures evaluated in this study are 800°F, 900°F, and 1000°F. Designs EXT-SS3, EXT-SS1 and EXT-SS4 allow for the comparision of these different steam cycle temperatures since, in all other respects, their operating condition are the same. That is, they all generate only saturated steam in the gas coolers; they all employ a quench gas recycle; and their depth of sulfur removal is 94.6 percent. As with all these sensitivity studies, the performance parameters, process flow diagrams, and detailed cost estimates are tabulated at the end of this section. Table 6-1 provides a summary of this informaton.

As in the previous case of steam cycle pressure, the impact of steam superheat/reheat temperatures on the cost of fuel gas is critically impacted by the value assigned to the by-product electric power. The financial results presented in Table 6-1 for fuel gas generated by a regulated utility have been calculated on the basis of an electricity value of 50 mills/kWh in mid-1980 dollars. The following table summarizes pertinent fuel gas costs from Table 6-1 and presents equivalent fuel gas costs assuming an electricity credit of 100 mills/kWh in mid-1980 dollars.

Fuel Gas Production Costs For An
Investor Owned Utility, Constant Mid-1980 Dollars

Case Designation	EXT-SS3	EXT-SS1	EXT-SS4
Steam Generated in Gas Coolers	Saturated	Saturated	Saturated
Steam Cycle, psig/°F/°F	1450/800/800	1450/900/900	1450/1000/1000
Fuel Gas Production, 10^6 Btu/hr	6942	6802	6658
Net By-product Power, MW	88.92	105.96	125.32
Levelized Fuel Gas Cost, $/$10^6$ BTU			
Electricity Credited at 50 mills/kWh	3.41	3.39	3.36
Electricity Credited at 100 mills/kWh	2.80	2.64	2.46

As with the previous case, it can be seen that at an electricity credit of 50 mills/kWh, the impact of steam superheat/reheat temperatures on the cost of fuel gas in negligible (1%). However, at an electricity value of 100 mills/kWh, increasing superheat/reheat temperatures from 800°F to 1000°F will reduce the cost of fuel gas by 12 percent.

Table 6-3

SUMMARY OF OPERATING RESULTS - TEXACO GASIFICATION FUEL GAS PLANTS - ALL CASES

	Base Case						
Steam Cycle, psig/°F/°F	←	1450/900/900	1450/900/900	736/900	1450/800/800	1450/1000/1000	1450/900/900
Steam Generated in Gas Coolers	←	←	←	Saturated	→	→	→
Gas Temperature Entering Heat Recovery, °F	←	2400	←	1500	→	→	2400
Sulfur Removal, %	←	←	←	94.6	→	→	+99.9¶
Oxidant Plant Compressor Drivers	←	←	←	Motors	→	→	→
Nominal Capacity of Gasifier, ST/day	1375	2200	←	1375	→	→	→
Case Designation	EXT-SS	EXT-SS*	EXT-SS1	EXT-SS2	EXT-SS3	EXT-SS4	EXT-SS5
GASIFICATION							
Coal Feed Rate, lb/hr m.f.	←	←	←	806,666	→	→	→
Oxygenδ/Coal Ratio, lb/lb m.f.	←	←	←	0.8921	→	→	→
Oxidant Temperature, °F	←	←	←	300	→	→	→
Slurry Water/Solids Feed Ratio, lb/lb m.f.	←	←	←	0.50308	→	→	→
Gasification Section Average Pressure, psig	←	←	←	600	→	→	→
Raw Gasifier Effluent Temperature, °F	←	←	←	2,400	→	→	→
Raw Gasifier Effluent HHV (dry basis), Btu/SCF**	←	←	←	275.8	→	→	→
Cold Gas Efficiency (raw gas HHV/coal feed HHV x 100), %	←	←	←	74.64	→	→	→
POWER SYSTEM							
Temperature of Fuel Gas to Gas Expander, °F	←	←	←	600	→	→	→
Gas Expander Exit Temperature, °F	←	←	←	195	→	→	198
Condenser Pressure, Inches Hg abs	←	←	←	2.5	→	→	→
Fired Heater Stack Temperature, °F	←	←	←	400	→	→	→
Gas Expander Power#, MW	←	←	←	61.75	→	→	60.81
Steam Turbine Power#, MW	←	258.97	227.98	179.43	206.91	247.61	255.93
Oxygen Plant Power##, MW	←	←	←	1.81	→	→	→
Power Consumed, MW	←	180.14	185.58	184.71	181.55	185.65	181.94
Net System Power, MW	←	142.40	105.96	58.28	88.92	125.32	136.61
OVERALL SYSTEM							
General Facilities Water Consumption, GPM	←	←	←	150	→	→	→
Land, Acres	190	188	←	190	→	→	→
Ash Disposal Rate, Dry, ST/day	←	←	←	1023	→	→	→
Sulfur By-Product, ST/day	←	←	←	354	→	→	373
Process & Deaerator Makeup Water, gal/ton DAF coal	←	77.19	76.53	77.02	76.36	76.53	77.19
Cooling Tower Makeup Water, gal/ton DAF coal	←	722.31	668.93	618.35	649.75	686.12	716.69
Cooling Water Circulation Rate§§, mgal/ton DAF coal	←	37.48	34.71	32.50	33.75	35.62	37.15
Cooling Tower Heat Rejection§§, % of coal HHV	←	22.11	20.48	18.98	19.90	21.00	22.00
Air Cooler Heat Rejections, % of coal HHV	←	2.10	3.78	3.90	3.78	3.75	2.51
Clean Fuel Gas Efficiency (exported clean gas HHV x 100/coal feed HHV), %	←	64.67	66.01	69.51	67.37	64.61	64.30
Energy Recovery Efficiency (exported power¶¶ + exported clean gas HHV)/coal feed HHV x 100, %	←	77.11	75.27	74.61	75.14	75.56	76.24
Clean Fuel Gas HHV (dry basis), Btu/SCF	←	←	←	282.3	→	→	283.8
Net Clean Fuel Gas Product, 10^6 SCFD	←	566.63	578.25	608.95	590.16	566.04	560.35
10^3 SCF/ton DAF coal	←	65.03	66.37	69.90	67.74	64.97	64.32
10^6 Btu/hr	←	6664	6802	7163	6942	6658	6626

*Case EXT-SS with 2200 ST/day gasifiers
δDry basis, 100 percent oxygen
**Excluding the HHV of H_2S, COS, and NH_3

#At generator terminals
##From power recovery expander 11-ME-1
§§Includes process and power plant portions

¶One ppm total sulfur in product gas on a mole basis
¶¶Power credited at 9000 Btu/kWh

Table 6-3 (Continued)

SUMMARY OF OPERATING RESULTS - TEXACO GASIFICATION FUEL GAS PLANTS - ALL CASES

	Base Case	1450/1000/1000	1450/900/900	736/900	736/900	1450/1000/1000	1450/1000/1000
Steam Cycle, psig/°F/°F	←						
Steam Generated in Gas Coolers	←	←	←	SUPERHEATED	→	→	→
Gas Temperature Entering Heat Recovery, °F	←	2400	←	1500	2400	→	→
Sulfur Removal, %	←	←	←	94.6	→	83.6	+99.9¶
Oxidant Plant Compressor Drivers	←	←	←	MOTORS	STEAM TURBINES	MOTORS	→
Nominal Capacity of Gasifier, ST/day	1375	2200	←	1375	→	→	→
Case Designation	EXT-SH	EXT-SH*	EXT-SH1	EXT-SH2	EXT-SH3	EXT-SH4	EXT-SH5
GASIFICATION							
Coal Feed Rate, lb/hr m.f.	←	←	←	806,666	→	→	→
Oxygenδ/Coal Ratio, lb/lb m.f.	←	←	←	0.8921	→	→	→
Oxidant Temperature, °F	←	←	←	300	→	→	→
Slurry Water/Solids Feed Ratio, lb/lb m.f.	←	←	←	0.50308	→	→	→
Gasification Section Average Pressure, psig	←	←	←	600	→	→	→
Raw Gasifier Effluent Temperature, °F	←	←	←	2,400	→	→	→
Raw Gasifier Effluent HHV (dry basis), Btu/SCF**	←	←	←	275.8	→	→	→
Cold Gas Efficiency (raw gas HHV/coal feed HHV x 100), %	←	←	←	74.64	→	→	→
POWER SYSTEM							
Temperature of Fuel Gas to Gas Expander, °F	←	←	←	600	→	→	→
Gas Expander Exit Temperature, °F	←	←	←	195	→	196	198
Condenser Pressure, Inches Hg abs	←	←	←	2.5	→	→	→
Fired Heater Stack Temperature, °F	-	-	400	-	-	-	-
Gas Expander Power##, MW	←	←	←	61.75	→	62.25	60.81
Steam Turbine Power#, MW		166.85	166.93	135.39	8.05	166.73	163.51
Oxygen Plant Power##, MW	←	←	←	1.81	→	→	→
Power Consumed, MW	←	177.11	183.67	183.11	35.04	176.06	178.93
Net System Power, MW	←	53.30	46.82	15.84	36.57	54.73	47.20
OVERALL SYSTEM							
General Facilities Water Consumption, GPM	←	←	←	150	→	→	→
Land, Acres	175	173	185	175	→	→	→
Ash Disposal Rate, Dry, ST/day	←	←	←	1023	→	→	→
Sulfur By-Product, ST/day	←	←	←	354	→	312	373
Process and Deaerator Makeup Water, gal/ton DAF coal	←	74.38	74.88	75.37	76.03	74.38	74.38
Cooling Tower Makeup Water, gal/ton DAF coal	←	530.91	551.74	518.18	608.26	527.27	524.30
Cooling Water Circulation Rate§§, mgal/ton DAF coal	←	27.52	28.58	26.98	29.14	27.34	27.15
Cooling Tower Heat Rejection§§, % of coal HHV	←	16.26	16.88	15.93	17.19	16.15	16.04
Air Cooler Heat Rejections, % of coal HHV	←	3.04	4.45	4.48	2.74	2.92	3.48
Clean Fuel Gas Efficiency (exported clean gas HHV x 100/coal feed HHV), %	←	74.11	71.83	74.11	74.11	74.42	73.96
Energy Recovery Efficiency (exported power¶¶ + exported clean gas HHV)/coal feed HHV x 100, %	←	78.77	75.92	75.50	77.31	79.20	78.08
Clean Fuel Gas HHV (dry basis), Btu/SCF	←	←	←	282.3	→	281.1	283.8
Net Clean Fuel Gas Product, 10^6 SCFD	←	649.25	629.28	649.25	649.25	654.68	644.49
10^3 SCF/ton DAF coal	←	74.52	72.23	74.52	74.52	75.15	73.98
10^6 Btu/hr	←	7637	7402	7637	7637	7668	7621

*Case EXT-SH with 2200 ST/day gasifiers
δDry basis, 100 percent oxygen
**Excluding the HHV of H_2S, COS, and NH_3
#At generator terminals
##From power recovery expander 11-ME-1
§§Includes process and power plant portions
¶One ppm total sulfur in product gas on a mole basis
¶¶Power credited at 9000 Btu/kWh

Table 6-4

POWER CONSUMPTION SUMMARY (kW at Full Design Capacity)
TEXACO GASIFICATION FUEL GAS PLANTS

		Base Case						
Steam Cycle, psig/°F/°F		1450/900/900	1450/900/900	1450/900/900	736/900	1450/800/800	1450/1000/1000	1450/900/900
Steam Generated in Gas Coolers					SATURATED			
Gas Temperature Entering Heat Recovery, °F		2400		1500				2400
Sulfur Removal, %					94.6			+99.9
Oxidant Plant Compressor Drivers					MOTORS			
Nominal Capacity of Gasifiers		1375	2200	1375				
Case Designation		EXT-SS	EXT-SS*	EXT-SS1	EXT-SS2	EXT-SS3	EXT-SS4	EXT-SS5
Plant Power, kW								
Unit	Unit Operation							
10	Coal Handling#	9,176	9,176	9,176	9,176	9,176	9,176	9,176
11	Oxidant Feed	143,439	143,439	143,439	143,439	143,439	143,439	143,439
20	Gasification	1,710	1,710	2,294	2,294	2,294	2,294	1,710
21	Gas Cooling	941	941	6,592	6,533	6,581	6,584	941
22	Acid Gas Removal	5,963	5,963	5,963	5,963	5,963	5,963	7,487
23	Sulfur Recovery	584	584	584	584	584	584	627
24	Tail Gas Treating	2,610	2,610	2,610	2,610	2,610	2,610	2,947
30	Raw Water Treating**	323	323	291	273	282	286	317
32-45	General Facilities	14,860	14,860	14,155	13,528	10,213	14,377	14,768
50	Fuel Gas Expander	0	0	0	0	0	0	0
51	Steam System	513	513	464	298	395	516	513
51	Surface Condenser##	16	16	14	13	14	15	16
	TOTAL PLANT POWER, kW	180,135	180,135	185,583	184,711	181,551	185,846	181,941
	Power Produced, kW	322,530	322,530	291,540	242,990	270,470	311,170	318,550
	New Power Output, kW	142,400	142,400	105,960	58,280	88,920	125,320	136,610

* Case EXT-SS with 2200 ST/day gasifiers.
** About 70 percent of this power demand is attributable to the cooling water system pumps and fans
\# This power requirement is largely derived from the coal pulverizing system and is, therefore, preliminary pending grinding tests on Illinois No. 6 coal
\## For mechanical vacuum pump

Table 6-4 (Continued)

POWER CONSUMPTION SUMMARY (kW at Full Design Capacity)
TEXACO GASIFICATION FUEL GAS PLANTS

		Base Case							
Steam Cycle, psig/°F/°F		1450/1000/1000	1450/1000/1000	1450/900/900	736/900		736/900	1450/1000/1000	1450/1000/1000
Steam Generated in Gas Coolers		←			SATURATED		→		→
Gas Temperature Entering Heat Recovery, °F		2400	→	1500	→		2400	→	→
Sulfur Removal, %		94.6	→				→	83.6	+99.9
Oxidant Plant Compressor Drivers		MOTORS	→				STEAM TURBINES	MOTORS	→
Nominal Capacity of Gasifiers		1375	2200	1375	→				→
Case Designation		EXT-SS	EXT-SS*	EXT-SS1	EXT-SS2		EXT-SS3	EXT-SS4	EXT-SS5
Plant Power, kW									
Unit	Unit Operation								
10	Coal Handling#	9,176	9,176	9,176	9,176	9,176	9,176	9,176	
11	Oxidant Feed	143,439	143,439	143,439	143,439	73	143,439	143,439	
20	Gasification	1,710	1,710	2,294	2,294	1,710	1,710	1,710	
21	Gas Cooling	941	941	6,546	6,508	924	947	952	
22	Acid Gas Removal	5,963	5,963	5,963	5,963	5,963	5,492	7,487	
23	Sulfur Recovery	584	584	584	584	584	486	627	
24	Tail Gas Treating	2,610	2,610	2,610	2,610	2,610	2,181	2,947	
30	Raw Water Treating	222	222	218	209	230	220	217	
32-45	General Facilities**	12,377	12,377	12,649	12,239	13,694	12,328	12,293	
50	Fuel Gas Expander	0	0	0	0	0	0	0	
51	Steam System	70	70	182	80	79	68	69	
51	Surface Condenser##	10	10	11	10	12	10	10	
	TOTAL PLANT POWER, kW	177,111	177,111	183,672	183,113	35,044	176,057	178,927	
	Power Produced, kW	230,410	230,410	230,490	198,950		71,610	230,790	226,130
	New Power Output, kW	53,300	53,300	46,820	15,840		36,570	54,730	47,200

** About 70 percent of this power demand is attributable to the cooling water system pumps and fans
* Case EXT-SH with 2200 ST/day gasifiers.
\# This power requirement is largely derived from the coal pulverizing system and is, therefore, preliminary pending grinding tests on Illinois No. 6 coal
\## For mechanical vacuum pump

Table 6-5

ENERGY BALANCE - BASE CASE EXT-SS
AND CASE EXT-SS*

Gasifier: Texaco

Steam: 1450 psig, 900°F/900°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 2400°F

Sulfur Removal: 94.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	4	1	-	-	5
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	615	615
TOTAL	10,304	33	57	-	615	11,009
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	123	-	123
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses#	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	884	884
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	74	76	-	-	150
Fired Heater Radiation Losses	-	-	-	18	-	18
Steam Turbine Condensers	-	1,632	-	-	-	1,632
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	22	22
Motor and Mechanical Losses**	-	-	-	-	116	116
Net Fuel Gas Export	6,664	60	-	-	-	6,724
TOTAL	6,847	2,628	141	141	1,239	10,996

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.12\%$

#Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent
*Case EXT-SS with 2200 ST/day gasifiers
**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-6

ENERGY BALANCE - CASE EXT-SSI

Gasifier: Texaco

Steam: 1450 psig, 900°F/900°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 1500°F

Sulfur Removal: 94.6 percent

Basis: 60°F, water as liquid, 3,413 Btu/kWh

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	4	1	-	-	5
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	633	633
TOTAL	10,304	33	57	-	633	11,027
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	202	-	-	-	202
Gasifier and WHB Heat Losses	-	-	-	115	-	115
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	81	117	-	-	198
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	778	778
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	64	65	-	-	129
Fired Heater Radiation Losses	-	-	-	15	-	15
Steam Turbine Condensers	-	1,466	-	-	-	1,466
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	20	20
Motor and Mechanical Losses**	-	-	-	-	115	115
Net Fuel Gas Export	6,802	63	-	-	-	6,865
TOTAL	6,985	2,511	247	130	1,130	11,003

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.22\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-7

ENERGY BALANCE - CASE EXT-SS2

Gasifier: Texaco

Steam: 736 psig, 900°F superheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 1500°F

Sulfur Removal: 94.6%

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	3	-	-	-	3
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	630	630
TOTAL	10,304	32	56	-	630	11,022
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from OxidantPlant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	202	-	-	-	202
Gasifier & WHB Heat Losses	-	-	-	117	-	117
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	85	124	-	-	209
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	612	612
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	36	37	-	-	73
Fired Heater Radiation Losses	-	-	-	9	-	9
Steam Turbine Condensers	-	1,314	-	-	-	1,314
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	17	17
Motor and Mechanical Losses**	-	-	-	-	112	112
Net Fuel Gas Export	7,163	64	-	-	-	7,227
TOTAL	7,346	2,336	236	126	958	10,992

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.27\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-8

ENERGY BALANCE - CASE EXT-SS3

Gasifier: Texaco

Steam: 1450 psig, 800°F/800°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 1500°F

Sulfur Removal: 94.6 percent

Basis: 60°F, water as liquid, 3,413 Btu/kWh

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	4	1	-	-	5
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	620	620
TOTAL	10,304	33	57	-	620	11,014
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	202	-	-	-	202
Gasifier and WHB Heat Losses	-	-	-	115	-	115
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	81	117	-	-	198
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	706	706
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	53	54	-	-	107
Fired Heater Radiation Losses	-	-	-	13	-	13
Steam Turbine Condensers	-	1,407	-	-	-	1,407
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	19	19
Motor and Mechanical Losses**	-	-	-	-	101	101
Net Fuel Gas Export	6,942	62	-	-	-	7,004
TOTAL	7,125	2,440	236	128	1,043	10,972

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.38\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-9

ENERGY BALANCE - CASE EXT-SS4

Gasifier: Texaco

Steam: 1450 psig, 1000°F/1000°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 1500°F

Sulfur Removal: 94.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	4	1	-	-	5
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	634	634
TOTAL	10,304	33	57	-	634	11,028
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	202	-	-	-	202
Gasifier and WHB Heat Losses	-	-	-	115	-	116
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	79	116	-	-	195
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	845	845
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	74	76	-	-	150
Fired Heater Radiation Losses	-	-	-	18	-	18
Steam Turbine Condensers	-	1,518	-	-	-	1,518
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	22	22
Motor and Mechanical Losses**	-	-	-	-	116	116
Net Fuel Gas Export	6,658	60	-	-	-	6,718
TOTAL	6,841	2,568	257	134	1,200	11,000

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.25\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-10

ENERGY BALANCE - CASE EXT-SS5

Gasifier: Texaco

Oxidant: 98 mole percent oxygen

Steam: 1450 psig, 900°F/900°F superheat/reheat

Gas Temperature Entering Heat Recovery: 2400°F

Oxidant Plant Compressor Drivers: Motors

Sulfur Removal: +99.9 percent

Basis: 60°F, water as liquid, 3,413 Btu/kWh

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	4	1	-	-	5
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	621	621
TOTAL	10,304	33	57	-	621	11,015
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	118	-	118
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	72	-	-	-	72
Selexol Overhead Condenser	-	60	-	-	-	60
Tail Gas Treating Unit Cooling	-	23	12	-	-	35
Sulfur By-product (31,117 lbs/hr)	121	2	-	-	-	123
Tail Gas Unit Vent Gas	4	1	9	-	-	14
Steam Heat Losses*	-	3	20	-	-	23
Fuel Gas Expander Power	-	-	-	-	208	208
Steam Turbine Power	-	-	-	-	873	873
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	74	76	-	-	150
Fired Heater Radiation Losses	-	-	-	18	-	18
Steam Turbine Condensers	-	1,598	-	-	-	1,598
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	22	22
Motor and Mechanical Losses**	-	-	-	-	116	116
Net Fuel Gas Export	6,626	60	-	-	-	6,686
TOTAL	6,816	2,651	145	136	1,225	10,973

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.38\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-11

ENERGY BALANCE - BASE CASE EXT-SH
AND CASE EXT-SH*

Gasifier: Texaco

Steam: 1450 psig, 1000°F/1000°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 2400°F

Sulfur Removal: 94.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	604	604
TOTAL	10,304	29	56	-	604	10,993
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	113	-	113
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	47	50	-	-	97
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses#	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	570	570
Air Separation Plant Power	-	-	-	-	6	6
Steam Turbine Condensers	-	1,034	-	-	-	1,034
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	16	16
Motor and Mechanical Losses**	-	-	-	-	105	105
Net Fuel Gas Export	7,637	69	-	-	-	7,708
TOTAL	7,820	2,012	115	113	908	10,968

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.23\%$

#Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent
*Case EXT-SH with 2200 ST/day gasifiers
**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-12

ENERGY BALANCE - CASE EXT-SH1

Gasifier: Texaco

Steam: 1450 psig, 900°F/900°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/KWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 1500°F

Sulfur Removal: 94.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Air to Fired Heater	-	1	-	-	-	1
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	627	627
TOTAL	10,304	30	56	-	627	11,017
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	202	-	-	-	202
Gasifier and WHB Heat Losses	-	-	-	116	-	116
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	109	157	-	-	266
Raw Gas Trim Cooler	-	19	5	-	-	24
elexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	570	570
Air Separation Plant Power	-	-	-	-	6	6
Fired Heater Flue Gas	-	18	20	-	-	38
Fired Heater Radiation Loss	-	-	-	4	-	4
Steam Turbine Condensers	-	1,098	-	-	-	1,098
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	16	16
Motor and Mechanical Losses**	-	-	-	-	108	108
Net Fuel Gas Export	7,402	67	-	-	-	7,469
TOTAL	7,585	2,129	242	120	911	10,987

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.27\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-13

ENERGY BALANCE - CASE EXT-SH2

Gasifier: Texaco

Steam: 736 psig, 900°F superheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 1500°F

Sulfur Removal: 94.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	625	625
TOTAL	10,304	29	56	-	625	11,014
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	202	-	-	-	202
Gasifier and WHB Heat Losses	-	-	-	118	-	118
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	111	159	-	-	270
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	462	462
Air Separation Plant Power	-	-	-	-	6	6
Steam Turbine Condensers	-	1,001	-	-	-	1,001
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	14	14
Motor and Mechanical Losses**	-	-	-	-	107	107
Net Fuel Gas Export	7,637	69	-	-	-	7,706
TOTAL	7,820	2,018	224	118	800	10,980

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.31\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-14

ENERGY BALANCE - CASE EXT-SH3

Gasifier: Texaco

Oxidant: 98 mole percent oxygen

Steam: 736 psig, 900°F superheat

Gas Temperature Entering Heat Recovery: 2400°F

Oxidant Plant Compressor Drivers: Steam Turbines

Sulfur Removal: 94.6 percent

Basis: 60°F, water as liquid, 3,413 Btu/kWh

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	120	120
TOTAL	10,304	29	56	-	120	10,509
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	120	-	120
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	32	34	-	-	66
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	59	-	-	-	59
Selexol Overhead Condenser	-	19	-	-	-	19
Tail Gas Treating Unit Cooling	-	21	13	-	-	34
Sulfur By-product (29,490 lbs/hr)	115	2	-	-	-	117
Tail Gas Unit Vent Gas	3	1	4	-	-	8
Steam Heat Losses*	-	2	20	-	-	22
Fuel Gas Expander Power	-	-	-	-	211	211
Steam Turbine Power	-	-	-	-	28	28
Air Separation Plant Power	-	-	-	-	6	6
Steam Turbine Condensers	-	1,138	-	-	-	1,138
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	5	5
Motor and Mechanical Losses**	-	-	-	-	97	97
Net Fuel Gas Export	7,637	69	-	-	-	7,706
TOTAL	7,820	2,101	99	120	347	10,487

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.21\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-15

ENERGY BALANCE - CASE EXT-SH4

Gasifier: Texaco

Steam: 1450 psig, 1000°F/1000°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 2400°F

Sulfur Removal: 83.6 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	-	1	-	-	1
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	601	601
TOTAL	10,304	28	56	-	601	10,989
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	121	-	121
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	46	50	-	-	96
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	57	-	-	-	57
Selexol Overhead Condenser	-	10	-	-	-	10
Tail Gas Treating Unit Cooling	-	16	15	-	-	31
Sulfur By-product (26,020 lbs/hr)	101	2	-	-	-	103
Tail Gas Unit Vent Gas	10	1	6	-	-	17
Steam Heat Losses*	-	1	20	-	-	21
Fuel Gas Expander Power	-	-	-	-	212	212
Steam Turbine Power	-	-	-	-	569	569
Air Separation Plant Power	-	-	-	-	6	6
Steam Turbine Condensers	-	1,024	-	-	-	1,024
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	16	16
Motor and Mechanical Losses**	-	-	-	-	104	104
Net Fuel Gas Export	7,668	69	-	-	-	7,737
TOTAL	7,844	1,984	119	121	907	10,975

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.13\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-16

ENERGY BALANCE - CASE EXT-SH5

Gasifier: Texaco

Steam: 1450 psig, 1000°F/1000°F superheat/reheat

Oxidant Plant Compressor Drivers: Motors

Basis: 60°F, water as liquid, 3,413 Btu/kWh

Oxidant: 98 mole percent oxygen

Gas Temperature Entering Heat Recovery: 2400°F

Sulfur Removal: +99.9 percent

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
IN						
Coal	10,304	3	-	-	-	10,307
Air to Oxidant Feed System	-	23	55	-	-	78
Air to Sulfur Recovery	-	1	1	-	-	2
Demineralized and Raw Water	-	2	-	-	-	2
Auxiliary Power Inputs	-	-	-	-	611	611
TOTAL	10,304	29	56	-	611	11,000
OUT						
Oxidant Compressors Inter/After Cooling	-	468	-	-	-	468
Condensate from Oxidant Plant	-	1	-	-	-	1
Air Separation Plant Vent Gas	-	19	23	-	-	42
Process Water Cooling	-	227	-	-	-	227
Gasifier and WHB Heat Losses	-	-	-	121	-	121
Ash Cake	65	6	-	-	-	71
Process Wastewater	-	18	-	-	-	18
Raw Gas Air Cooler	-	49	51	-	-	100
Raw Gas Trim Cooler	-	19	5	-	-	24
Selexol Refrigeration Cooler	-	72	-	-	-	72
Selexol Overhead Condenser	-	60	-	-	-	60
Tail Gas Treating Unit Cooling	-	23	12	-	-	35
Sulfur By-product (31,117 lbs/hr)	121	2	-	-	-	123
Tail Gas Unit Vent Gas	4	1	9	-	-	14
Steam Heat Losses*	-	3	20	-	-	23
Fuel Gas Expander Power	-	-	-	-	208	208
Steam Turbine Power	-	-	-	-	558	558
Air Separation Plant Power	-	-	-	-	6	6
Steam Turbine Condensers	-	988	-	-	-	988
Steam Turbine and Fuel Gas Expander Losses	-	-	-	-	16	16
Motor and Mechanical Losses**	-	-	-	-	106	106
Net Fuel Gas Export	7,621	69	-	-	-	7,690
TOTAL	7,811	2,025	120	121	894	10,971

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.26\%$

*Includes tracing, miscellaneous process users, sulfur melting steam and deaerator vent

**Includes all pump and turbine mechanical losses, power to waste water treatment, Unit 32 and other offsites

Table 6-17

PLANT FACILITIES INVESTMENT - BASE CASE EXT-SS
MID-1978 DOLLARS

Steam Cycle: 1450 psig/900°/900°F

Gas Temperature Entering Gas Cooling: 2400°F

Oxidant Plant Compressor Drivers: Motors

Sulfur Removal: 94.6 percent

Product Fuel Gas: $6,664 \times 10^6$ Btu/hr**

By-Product Power: 142.40 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/MM BTU/HR
Coal Handling	20,566	8,356	14,413	1,026	44,361	6.657	8.2	0	6,654	51,015	7.657
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	23.417	29.0	0	23,407	179,453	26.930
Gasification and Ash Handling	24,062	9,810	16,702	1,171	51,745	7.765	9.7	5,133	7,762	64,640	9.700
Raw Gas Cooling	71,124	16,455	34,929	3,590	126,098	18.923	23.4	22,825	18,913	167,836	25.186
Acid Gas Removal	6,918	2,493	4,513	353	14,277	2.142	2.7	0	2,142	16,419	2.464
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.899	1.1	0	899	6,889	1.034
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.552	2.0	1,551	1,551	13,441	2.017
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	2.010	2.5	0	2,009	15,402	2.311
Steam, Condensate and BFW	970	351	426	19	1,766	.265	.3	0	265	2,031	.305
Steam Superheat/Reheat	7,044	2,090	4,070	363	13,567	2.036	2.5	0	2,035	15,602	2.340
Steam Turbine Generator	20,858	6,189	12,049	1,074	40,170	6.028	7.5	0	6,026	46,196	6.932
General Facilities	28,485	11,526	16,801	963	57,775	8.670	10.7	0	8,666	66,441	9.970
Initial Chemicals and Catalysts	-	-	-	-	2,358	0.354	0.4	-	-	2,358	0.354
TOTAL	279,337	93,220	151,854	11,112	537,885	80.718	100.0	29,509	80,329	647,723	97.200

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/MM BTU/HR**
Process Plant Investment and General Facilities	537,885	80.718
Process Contingency	29,509	4.428
Project Contingency	80,329	12.054
Total Plant Investment	647,723	97.200

*Mid-1978 dollars

**Based on 100 percent plant design power output of 6663.8 MM Btu/hr Fuel Gas

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:

- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-18

PLANT FACILITIES INVESTMENT - BASE CASE EXT-SS WITH 2200 ST/D GASIFIER
MID-1978 DOLLARS

Steam Cycle: 1450 psig/900°/900°F — Oxidant Plant Compressor Drivers: Motors — Product Fuel Gas: 6,664 x 10^6 Btu/hr**

Gas Temperature Entering Gas Cooling: 2400°F — Sulfur Removal: 94.6 percent — By-Product Power: 142.40 MW

	Cost Breakdown Without Contingencies							Contingencies		Total Plant Investment	
Plant Section	Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Process $1000*	Project $1000*	$1000*	$/MM BTU/HR
Coal Handling	19,807	8,051	13,907	991	42,756	6.416	8.5	0	6,413	49,169	7.380
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	23.417	31.2	0	23,407	179,453	26.930
Gasification and Ash Handling	20,344	8,185	14,008	993	43,530	6.532	8.7	4,105	6,530	54,165	8.128
Raw Gas Cooling	55,866	12,925	27,434	2,820	99,045	14.864	19.8	18,028	14,855	131,928	19.797
Acid Gas Removal	6,918	2,493	4,513	353	14,277	2.142	2.8	0	2,142	16,419	2.464
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.899	1.2	0	899	6,889	1.034
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.552	2.1	1,551	1,551	13,441	2.017
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	2.010	2.7	0	2,009	15,402	2.311
Steam, Condensate and BFW	970	351	426	19	1,766	.265	.4	0	265	2,031	.305
Steam Superheat/Reheat	7,044	2,090	4,070	363	13,567	2.036	2.7	0	2,035	15,602	2.340
Steam Turbine Generator	20,858	6,189	12,049	1,074	40,170	6.028	8.0	0	6,026	46,196	6.932
General Facilities	28,210	11,434	16,724	963	57,331	8.603	11.4	0	8,600	65,931	9.894
Initial Chemicals and Catalysts	-	-	-	-	2,358	0.354	0.5	-	-	2,358	0.354
TOTAL	254,327	87,668	141,086	10,129	500,568	75.118	100.0	23,684	74,732	598,984	89.886

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/MM BTU/HR**
Process Plant Investment and General Facilities	500,568	75.118
Process Contingency	23,684	3.554
Project Contingency	74,732	11.214
Total Plant Investment	598,984	89.886

*Mid-1978 dollars

**Based on 100 percent plant design power output of 6663.8 MM Btu/hr Fuel Gas

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-19

TOTAL PLANT INVESTMENT - CASE EXT-SS1
MID-1978 DOLLARS

Steam Cycle: 1450 psig/900°/900°F

Gas Temperature Entering Gas Cooling: 1500°F

Oxidant Plant Compressor Drivers: Motors

Sulfur Removal: 94.6 percent

Product Fuel Gas: 6,802 x 10^6 Btu/hr**

By-Product Power: 105.96 MW

	Cost Breakdown Without Contingencies							Contingencies		Total Plant Investment	
Plant Section	Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Process $1000*	Project $1000*	$1000*	$/10^6 Btu/hr**
Coal Handling	20,566	8,356	14,413	1,026	44,361	6.522	8.6	0	6,654	51,015	7.500
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	22.942	30.1	0	23,407	179,453	26.384
Gasification & Ash Handling	36,916	13,511	23,907	1,820	76,154	11.196	14.6	8,613	11,423	96,190	14.142
Raw Gas Cooling	50,022	11,567	24,580	2,528	88,697	13.044	17.1	10,303	13,304	112,304	16.511
Acid Gas Removal	6,918	2,493	4,513	353	14,277	2.099	2.8	0	2,142	16,419	2.414
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.881	1.2	0	899	6,889	1.013
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.520	2.0	1,551	1,551	13,441	1.976
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.969	2.6	0	2,009	15,402	2.264
Steam Condensate & BFW	920	332	402	18	1,672	.246	.3	0	251	1,923	.283
Steam Superheat/Reheat	6,426	1,907	3,712	331	12,376	1.819	2.4	0	1,856	14,232	2.092
Steam Turbine Generator	18,776	5,571	10,848	967	36,162	5.317	7.0	0	5,424	41,586	6.114
General Facilities	28,149	11,315	16,580	962	57,006	8.381	10.9	0	8,551	65,557	9.639
Initial Catalysts and Chemicals	-	-	-	-	2,204	0.324	.4	-	-	2,204	0.324
TOTAL	268,003	91,002	146,910	10,558	518,677	76.257	100.0	20,467	77,471	616,615	90.656

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/10^6 Btu/hr**
Process Plant Investment and General Facilities	518,677	76.257
Process Contingency	20,467	3.009
Project Contingency	77,471	11.390
Total Plant Investment	616,615	90.656

*Mid-1978 dollars

**Based on HHV of net fuel gas product at 100 percent capacity factor

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-20

TOTAL PLANT INVESTMENT - CASE EXT-SS2
MID-1978 DOLLARS

Steam Cycle: 736 psig/900°F

Gas Temperature Entering Gas Cooling: 1500°F

Oxidant Plant Compressor Drivers: Motors

Sulfur Removal: 94.6 percent

Product Fuel Gas: $7,163 \times 10^6$ Btu/hr**

By-Product Power: 58.28 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/10^6 Btu/hr**
Coal Handling	20,566	8,356	14,413	1,026	44,361	6.193	8.9	0	6,654	51,015	7.122
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	21.786	31.2	0	23,407	179,453	25.053
Gasification & Ash Handling	36,916	13,511	23,907	1,820	76,154	10.632	15.2	8,613	11,423	96,190	13.429
Raw Gas Coaling	48,405	11,193	23,785	2,447	85,830	11.983	17.2	10,408	12,873	109,111	15.233
Acid Gas Removal	6,918	2,493	4,513	353	14,277	1.993	2.9	0	2,142	16,419	2.292
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.836	1.2	0	899	6,889	.962
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.443	2.1	1,551	1,551	13,441	1.877
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.870	2.7	0	2,009	15,402	2.150
Steam Condensate & BFW	872	315	383	17	1,587	.222	0.3	0	238	1,825	.255
Steam Superheat/Reheat	2,266	673	1,310	117	4,366	.610	0.9	0	655	5,021	.701
Steam Turbine Generator	15,601	4,629	9,015	803	30,048	4.195	6.0	0	4,507	34,555	4.824
General Facilities	27,390	10,957	16,127	944	55,418	7.737	11.0	0	8,313	63,731	8.898
Initial Catalysts and Chemicals	-	-	-	-	1,803	0.252	0.4	-	-	1,803	0.252
TOTAL	258,244	88,077	141,408	10,080	499,612	69.752	100.0	20,572	74,671	594,855	83.048

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/10^6 Btu/hr**
Process Plant Investment and General Facilities	499,612	69.752
Process Contingency	20,572	2.872
Project Contingency	74,671	10.424
Total Plant Investment	594,855	83.048

*Mid-1978 dollars

**Based on HHV of net fuel gas product at 100 percent capacity factor

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:

- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-21

TOTAL PLANT INVESTMENT - CASE EXT-SS3
MID-1978 DOLLARS

Steam Cycle: 1450 psig/800°/800°F

Gas Temperature Entering Gas Cooling: 1500°F

Oxidant Plant Compressor Drivers: Motors

Sulfur Removal: 94.6 percent

Product Fuel Gas: 6,942 x 10^6 Btu/hr**

By-Product Power: 88.92 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/$10^6$ Btu/hr**	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/$10^6$ Btu/hr**
Coal Handling	20,566	8,356	14,413	1,026	44,361	6.390	8.7	0	6,654	51,015	7.349
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	22.479	30.5	0	23,407	179,453	25.851
Gasification & Ash Handling	36,916	13,511	23,907	1,820	76,154	10.971	14.9	8,613	11,423	96,190	13.857
Raw Gas Cooling	48,891	11,304	24,022	2,471	86,688	12.488	16.9	9,883	13,003	109,574	15.785
Acid Gas Removal	6,918	2,493	4,513	353	14,277	2.057	2.8	0	2,142	16,419	2.365
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.863	1.2	0	899	6,889	.992
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.489	2.0	1,551	1,551	13,441	1.936
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.929	2.6	0	2,009	15,402	2.219
Steam Condensate & BFW	895	324	392	17	1,628	.235	0.3	0	244	1,872	.270
Steam Superheat/Reheat	4,738	1,406	2,739	244	9,128	1.315	1.7	0	1,369	10,497	1.512
Steam Turbine Generator	18,509	5,491	10,692	953	35,645	5.134	7.0	0	5,347	40,992	5.905
General Facilities	27,664	11,137	16,327	947	56,075	8.078	11.0	0	8,411	64,486	9.289
Initial Catalysts and Chemicals	-	-	-	-	2,049	0.295	.4	-	-	2,049	.295
TOTAL	264,408	89,972	144,960	10,384	511,773	73.723	100.0	20,047	76,459	608,279	87.625

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/$10^6$ Btu/hr**
Process Plant Investment and General Facilities	511,773	73.723
Process Contingency	20,047	2.888
Project Contingency	76,459	11.014
Total Plant Investment	608,279	87.625

*Mid-1978 dollars

**Based on HHV of net fuel gas product at 100 percent capacity factor

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:

- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-22

TOTAL PLANT INVESTMENT - CASE EXT-SS4
MID-1978 DOLLARS

Steam Cycle: 1450 psig/1000°/1000°F — Oxidant Plant Compressor Drivers: Motors — Product Fuel Gas: 6,658 x 10^6 Btu/hr**

Gas Temperature Entering Gas Cooling: 1500°F — Sulfur Removal: 94.6 percent — By-Product Power: 125.32 MW

	Cost Breakdown Without Contingencies							Contingencies		Total Plant Investment	
Plant Section	Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Process $1000*	Project $1000*	$1000*	$/10^6 Btu/hr**
Coal Handling	20,566	8,356	14,413	1,026	44,361	6.663	8.4	0	6,654	51,015	7.662
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	23.438	29.3	0	23,407	179,453	26.953
Gasification & Ash Handling	36,916	13,511	23,907	1,820	76,154	11.438	14.4	8,613	11,423	96,190	14.447
Raw Gas Cooling	50,574	11,694	24,849	2,556	89,673	13.468	16.9	10,510	13,449	113,632	17.067
Acid Gas Removal	6,918	2,493	4,513	353	14,277	2.144	2.7	0	2,142	16,419	2.466
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.900	1.1	0	899	6,889	1.035
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.553	2.0	1,551	1,551	13,441	2.019
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	2.012	2.5	0	2,009	15,402	2.313
Steam Condensate & BFW	927	334	406	18	1,685	.253	.3	0	253	1,938	.291
Steam Superheat/Reheat	9,320	2,766	5,384	480	17,950	2.696	3.4	0	2,693	20,643	3.101
Steam Turbine Generator	21,608	6,411	12,483	1,112	41,614	6.250	7.8	0	6,242	47,856	7.188
General Facilities	28,369	11,425	16,717	967	57,478	8.633	10.8	0	8,622	66,100	9.928
Initial Catalysts and Chemicals	-	-	-	-	2,364	0.355	0.4	-	-	2,364	0.355
TOTAL	274,508	92,940	150,627	10,885	531,324	79.803	100.0	20,674	79,344	631,342	94.825

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/10^6 Btu/hr**
Process Plant Investment and General Facilities	531,324	79.803
Process Contingency	20,674	3.105
Project Contingency	79,344	11.917
Total Plant Investment	631,342	94.825

*Mid-1978 dollars

**Based on HHV of net fuel gas product at 100 percent capacity factor

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-23

TOTAL PLANT INVESTMENT - CASE EXT-SS5
MID-1978 DOLLARS

Steam Cycle: 1450 psig/900°/900°F   **Oxidant Plant Compressor Drivers:** Motors   **Product Fuel Gas:** 6,626 x 10^6 Btu/hr**

Gas Temperature Entering Gas Cooling: 2400°F   **Sulfur Removal:** +99.9 percent   **By-Product Power:** 136.61 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/MM Btu/hr	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/MM Btu/hr
Coal Handling	20,566	8,356	14,413	1,026	44,361	6.695	8.1	0	6,654	51,015	7.699
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	23.550	28.4	0	23,407	179,453	27.083
Gasification & Ash Handling	24,062	9,810	16,702	1,171	51,745	7.810	9.4	5,133	7,762	64,640	9.755
Raw Gas Cooling	72,376	16,744	35,542	3,653	128,315	19.364	23.4	25,021	19,246	172,582	26.046
COS Hydrolysis	447	246	400	23	1,116	.168	.2	0	167	1,283	.194
Acid Gas Removal	7,821	2,818	5,102	399	16,140	2.436	3.0	0	2,421	18,561	2.801
Sulfur Recovery	2,753	1,293	2,177	143	6,366	.961	1.2	0	955	7,321	1.105
Tail Gas Treating	4,899	2,240	3,638	230	11,007	1.661	2.0	1,651	1,651	14,309	2.160
ZnO Bed	1,812	326	775	91	3,004	.453	0.5	0	451	3,455	.521
Fuel Gas Expansion	6,900	2,047	3,985	355	13,287	2.005	2.4	0	1,993	15,280	2.306
Steam Condensate & BFW	962	347	421	18	1,748	.264	0.3	0	262	2,010	.303
Steam Superheat/Reheat	7,051	2,092	4,072	363	13,578	2.049	2.5	0	2,037	15,615	2.357
Steam Turbine Generator	20,642	6,125	11,927	1,063	39,757	6.000	7.2	0	5,964	45,721	6.900
General Facilities	28,521	11,526	16,817	966	57,830	8.728	10.5	0	8,675	66,505	10.037
Initial Chemicals and Catalyst	-	-	-	-	4,813	0.726	0.9	-	-	4,813	0.726
TOTAL	284,026	94,538	154,396	11,340	549,113	82.870	100.0	31,805	81,645	662,563	99.992

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/MM Btu/hr**
Process Plant Investment and General Facilities	549,113	82.870
Process Contingency	31,805	4.800
Project Contingency	81,645	12.322
Total Plant Investment	662,563	99.992

*Mid-1978 dollars

**Based on 100 percent plant design power output of 6626.1 MM Btu/hr fuel gas

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-24

PLANT FACILITIES INVESTMENT - BASE CASE EXT-SH
MID-1978 DOLLARS

Steam Cycle: 1450 psig/1000°/1000°F   **Oxidant Plant Compressor Drivers:** Motors   **Product Fuel Gas:** 7,637 x 10^6 Btu/hr**

Gas Temperature Entering Gas Cooling: 2400°F   **Sulfur Removal:** 94.6 percent   **By-Product Power:** 53.30 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Percent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/MM BTU/HR
Coal Handling	20,566	8,356	14,413	1,026	44,361	5.809	8.5	0	6,654	51,015	6.681
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	20.433	29.9	0	23,407	179,453	23.499
Gasification and Ash Handling	24,062	9,810	16,702	1,171	51,745	6.776	9.9	5,133	7,762	64,640	8.464
Raw Gas Cooling	82,055	18,973	40,318	4,147	145,493	19.051	27.9	32,010	21,822	199,325	26.100
Acid Gas Removal	6,918	2,493	4,513	353	14,277	1.870	2.7	0	2,142	16,419	2.150
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.784	1.1	0	859	6,889	.902
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.354	2.0	1,551	1,551	13,441	1.760
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.754	2.6	0	2,009	15,402	2.017
Steam, Condensate and LFW	754	271	319	13	1,357	.178	.3	0	204	1,561	.204
Steam Turbine Generator	13,543	4,018	7,823	697	26,081	3.415	5.0	0	3,912	29,993	3.927
General Facilities	25,738	10,206	15,091	895	51,930	6.800	9.9	0	7,790	59,720	7.820
Initial Chemicals and Catalysts	-	-	-	-	1,277	0.167	0.2	-	-	1,277	0.167
TOTAL	272,946	90,077	147,134	10,855	522,289	68.391	100.0	38,694	78,152	639,135	83.691

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/MM BTU/HR**
Process Plant Investment and General Facilities	522,289	68.391
Process Contingency	38,694	5.066
Project Contingency	78,152	10.234
Total Plant Investment	639,135	83.691

*Mid-1978 dollars

**Based on 100 percent plant design power output of 7,636.8 MM Btu/hr Fuel Gas

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-25

PLANT FACILITIES INVESTMENT - BASE CASE EXT-SH WITH 2200 ST/D GASIFIERS
MID-1978 DOLLARS

Steam Cycle: 1450 psig/1000°/1000°F | Oxidant Plant Compressor Drivers: Motors | Product Fuel Gas: 7,637 x 10^6 Btu/hr**

Gas Temperature Entering Gas Cooling: 2400°F | Sulfur Removal: 94.6 percent | By-Product Power: 53.30 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/MM Btu/hr	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/MM Btu/hr
Coal Handling	19,807	8,051	13,907	991	42,756	5.599	8.8	0	6,413	49,169	6.439
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	20.433	32.3	0	23,407	179,453	23.499
Gasification and Ash Handling	20,344	8,185	14,008	993	43,530	5.700	9.0	4,106	6,529	54,165	7.093
Raw Gas Cooling	65,879	15,234	32,372	3,330	116,815	15.296	24.2	25,933	17,522	160,270	20.986
Acid Gas Removal	6,918	2,493	4,513	353	14,277	1.870	3.0	0	2,142	16,419	2.150
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.784	1.2	0	899	6,889	.902
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.354	2.1	1,551	1,551	1	1.760
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.754	2.8	0	2,009	15,402	2.017
Steam, Condensate and BFW	754	271	319	13	1,357	.178	.3	0	204	1,561	.204
Steam Turbine-Generator	13,543	4,018	7,823	697	26,081	3.415	5.4	0	3,912	29,993	3.927
General Facilities	25,463	10,114	15,015	894	51,486	6.742	10.6	0	7,723	59,209	7.753
Initial Chemicals and Catalysts	-	-	-	-	-	-	-	-	-	1,277	.167
TOTAL	252,018	84,316	135,912	9,824	483,367	63.292	100.0	31,590	72,311	585,971	76.897

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/MM Btu/hr**
Process Plant Investment and General Facilities	483,347	63.292
Process Contingency	31,590	4.136
Project Contingency	72,311	9.469
Total Plant Investment	587,248	76.897

*Mid-1978 dollars

**Based on 100 percent 7,636.8 MM Btu/hr fuel gas

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-26

TOTAL PLANT INVESTMENT - CASE EXT-SH1
MID-1978 DOLLARS

<u>Steam Cycle</u>: 1450 psig/900°/900°F | <u>Oxidant Plant Compressor Drivers</u>: Motors | <u>Product Fuel Gas</u>: 7,402 x 10^6 Btu/hr**

<u>Gas Temperature Entering Gas Cooling</u>: 1500°F | <u>Sulfur Removal</u>: 94.6 percent | <u>By-Product Power</u>: 46.82 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Percent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/10^6 Btu/hr**
Coal Handling	20,566	8,356	14,413	1,026	44,361	5.993	9.2	0	6,654	51,015	6.892
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	21.082	32.4	0	23,407	179,453	24.244
Gasification and Ash Handling	36,916	13,511	23,907	1,820	76,154	10.288	15.8	8,613	11,423	96,190	12.995
Raw Gas Cooling	42,983	9,939	21,121	2,173	76,216	10.297	15.9	8,947	11,432	96,595	13.050
Acid Gas Removal	6,918	2,493	4,513	353	14,277	1.929	3.0	0	2,142	16,419	2.218
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.809	1.2	0	899	6,889	.931
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.397	2.1	1,551	1,551	13,441	1.816
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.809	2.8	0	2,009	15,402	2.081
Steam, Condensate and BFW	857	307	362	15	1,541	.208	.3	0	231	1,772	.239
Steam Superheat/Reheat	1,219	362	705	63	2,349	.317	.5	0	352	2,701	.365
Steam Turbine-Generator	13,616	4,040	7,865	701	26,222	3.545	5.5	0	3,933	30,155	4.074
General Facilities	26,343	10,495	15,541	921	53,300	7.201	11.0	0	7,995	61,295	8.281
Initial Catalysts and Chemicals	-	-	-	-	1,538	.208	.3	-	-	1,538	.208
TOTAL	248,728	85,453	136,382	9,625	481,726	65.081	100.0	19,111	72,028	572,865	77.394

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/10^6 Btu/hr**
Process Plant Investment and General Facilities	481,726	65.081
Process Contingency	19,111	2.582
Project Contingency	72,028	9.731
Total Plant Investment	572,865	77.394

*Mid-1978 dollars

**Based on HHV of net fuel gas product at 100 percent capacity factor

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee

Table 6-27

TOTAL PLANT INVESTMENT - CASE EXT-SH2
MID-1978 DOLLARS

Steam Cycle: 736 psig/900°F | Oxidant Plant Compressor Drivers: Motors | Product Fuel Gas: 7.637 x 10^6 Btu/hr**

Gas Temperature Entering Gas Cooling: 1500°F | Sulfur Removal: ..6 percent | By-Product Power: 15.84 MW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/10^6 Btu/hr**	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/10^6 Btu/hr**
Coal Handling	20,566	8,356	14,413	1,026	44,361	5.809	9.2	0	6,654	51,015	6.680
Oxidant Feed	85,214	30,568	38,425	1,839	156,046	20.433	32.1	0	23,407	179,453	23.498
Gasification and Ash Handling	36,916	13,511	23,907	1,820	76,154	9.972	15.7	8,613	11,423	96,190	12.596
Raw Gas Cooling	49,705	11,493	24,422	2,512	88,132	11.540	18.1	12,287	13,219	113,638	14.880
Acid Gas Removal	6,918	2,493	4,513	353	14,277	1.870	2.9	0	2,142	16,419	2.150
Sulfur Recovery	2,590	1,217	2,049	134	5,990	.784	1.2	0	899	6,889	.902
Tail Gas Treating	4,552	2,102	3,463	222	10,339	1.354	2.1	1,551	1,551	13,441	1.760
Fuel Gas Expansion	6,954	2,063	4,018	358	13,393	1.754	2.8	0	2,009	15,402	2.017
Steam, Condensate and BFW	804	289	338	14	1,445	.189	.3	0	217	1,662	.218
Steam Turbine-Generator	11,579	3,435	6,690	596	22,300	2.920	4.6	0	3,345	25,645	3.358
General Facilities	25,939	10,226	15,163	906	52,234	6.840	10.7	0	7,835	60,069	7.866
Initial Catalysts and Chemicals	-	-	-	-	1,277	.167	.3	-	-	1,277	.167
TOTAL	251,737	85,753	137,401	9,780	485,948	63.632	100.0	22,451	72,701	581,100	76.092

TOTAL PLANT INVESTMENT SUMMARY

	$1000*	$/10^6 Btu/hr**
Process Plant Investment and General Facilities	485,948	63.632
Process Contingency	22,451	2.940
Project Contingency	72,701	9.520
Total Plant Investment	581,100	76.092

*Mid-1978 dollars

**Based on HHV of net fuel gas product at 100 percent capacity factor

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes:
- Indirect field costs including all labor, supervision and expense required to support field construction
- Home office costs including all salaries and expenses required for engineering design and procurement
- Contractor's fee