Section 4

PLANT DESCRIPTIONS - CASE B2
LOW-CONVERSION METHANOL PLANT WITH COS HYDROLYSIS UNIT/1500 PSIG SATURATED STEAM AND 2000°F GAS TURBINES

GENERAL

A grass roots plant for electric power generation based on single-stage entrained oxygen-blown gasifiers of the Texaco type, integrated with current state-of-the-art combined-cycle generating equipment, is shown schematically on Block Flow Diagram EXTC(ME-B2)-1-1 for Case B2. Each block indicates the area and unit numbering, as well as the number of operating trains in each unit. The plant consumes 10,000 short tons per day of Illinois No. 6 coal, fed to the gasifiers in a water slurry containing 66.5 weight percent solids.

The main plant consists of coal pulverization and slurry preparation, oxidant feed, gasification, gas cooling, acid gas removal, zinc oxide treatment and methanol units, together with the combined-cycle power system. Coal receiving, storage, and conveying are accomplished in a single train to minimize space and operating labor requirements. Coal pulverization requires two parallel trains containing equipment of the largest sizes now available. The oxidant feed unit has five parallel operating trains. The gasification unit has three parallel operating trains and one spare train. One train ash handling system (without spare) serves all of the gasification units. The gas cooling comprising of COS hydrolysis unit and an acid gas removal unit, has two operating parallel trains. The zinc oxide treatment unit has three parallel trains. The methanol plant consists of five reactor trains, and one gas cooling and methanol flash train. There are five parallel gas turbines, ten heat recovery steam generators, and a single primary steam turbine.

In addition to the main processing trains, the plant includes necessary utility, environmental, and support facilities. Environmental safeguards have been considered by recovering elemental sulfur from the hydrogen sulfide in the acid gas.

Besides the two 50 percent operating trains, the sulfur recovery and tail gas treating units each have one 50 percent spare train to protect the environment in the event of equipment failure. Most of the process condensate is recycled to slurry preparation, while a small purge stream is treated before disposal. The plant storm water and utility waste water are collected and treated. The utility systems supporting the plant operation consist of a raw water treating unit, cooling towers, and a condensate collection and deaeration system. Additional support facilities provided are plant and instrument air, potable water, fuel gas flare, fire water, buildings, loading docks, and electrical distribution.

In the flow diagram numbering scheme, EXTC is an acronym for Entrained oXygen-blown Texaco gasifier, with a Combined-cycle power plant. ME designates a MEthanol coproduction study, and A2 and B2 refer to the two cases studied as described by the flow diagram titles. The numbers refer to the unit number and then the flow diagram number for each unit.

Table 4-1 shows the number of operating and spare trains for major sections of Case B2.

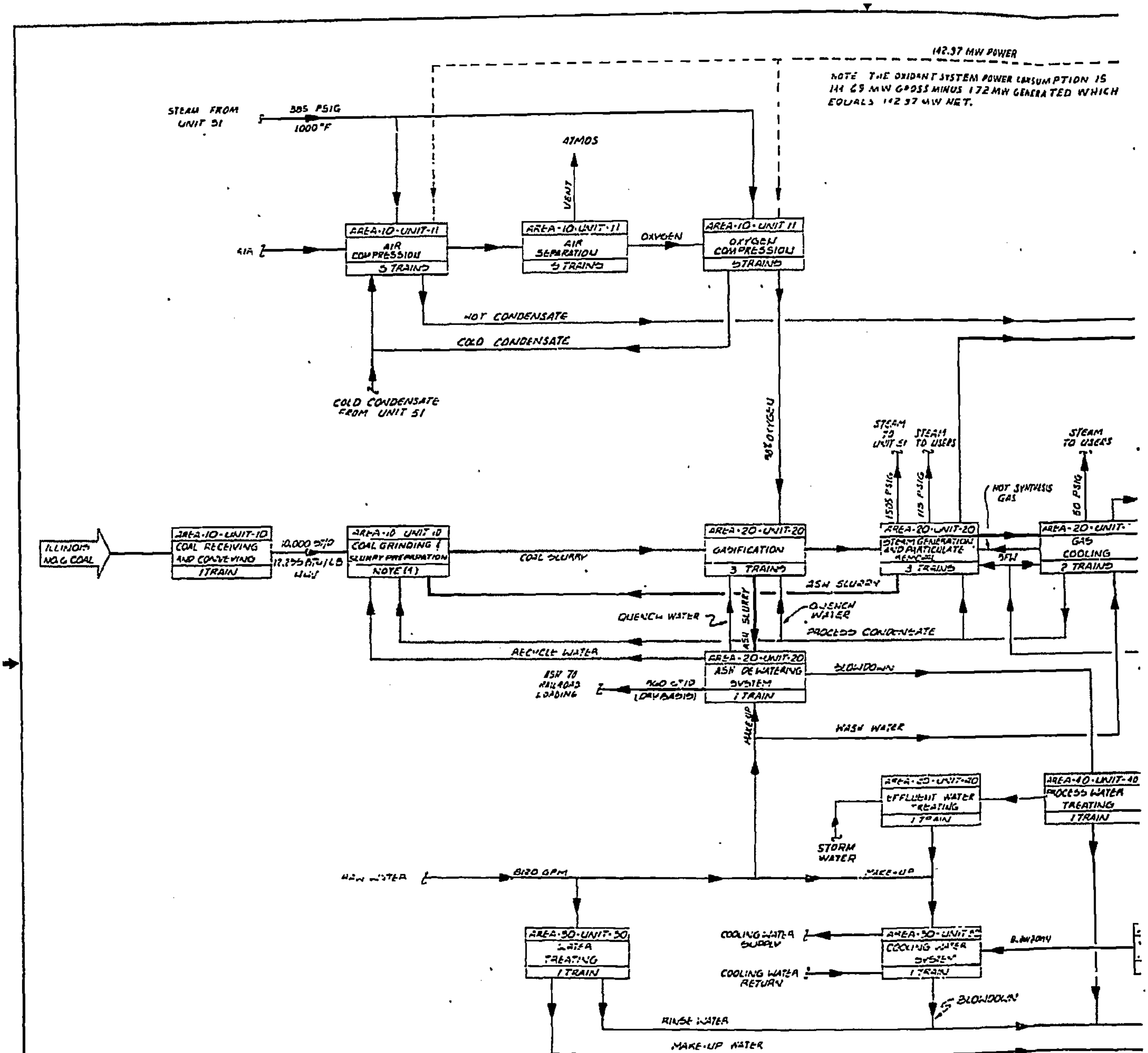
Table 4-1

TRAINS OF EQUIPMENT IN MAJOR PLANT SECTIONS - CASE B2

Unit No.	Unit Name	Operating	Spare
10	Coal Handling	1	0
10	Coal Grinding	2	0
10	Slurry Preparation	1	0
11	Oxidant Feed	5	0
20	Gasification	3	1
20	High-Temperature Gas Cooling and Gas Scrubbing	3	1
20	Ash Handling	1	0
21	Gas Cooling	2	0
21	COS Hydrolysis	2	0
22	Acid Gas Removal	2	0
23	Sulfur Recovery	2	1
24	Tail Gas Treating	2	1
25	Zinc Oxide Treating	3	0
25	Methanol Plant	*	0
30	Steam, BFW and Condensate System		
	• Condensate Collection and and Deaeration	1	0
	• Water Treating	1	0
32	Cooling Water System	1**	0
40	Effluent Water Treating	1	0
40	Process Condensate Treating	1	0
50	Gas Turbine/Generator	5	0
51	Heat Recovery Steam Generator	10	0
51	Steam Turbine/Generator	1	0

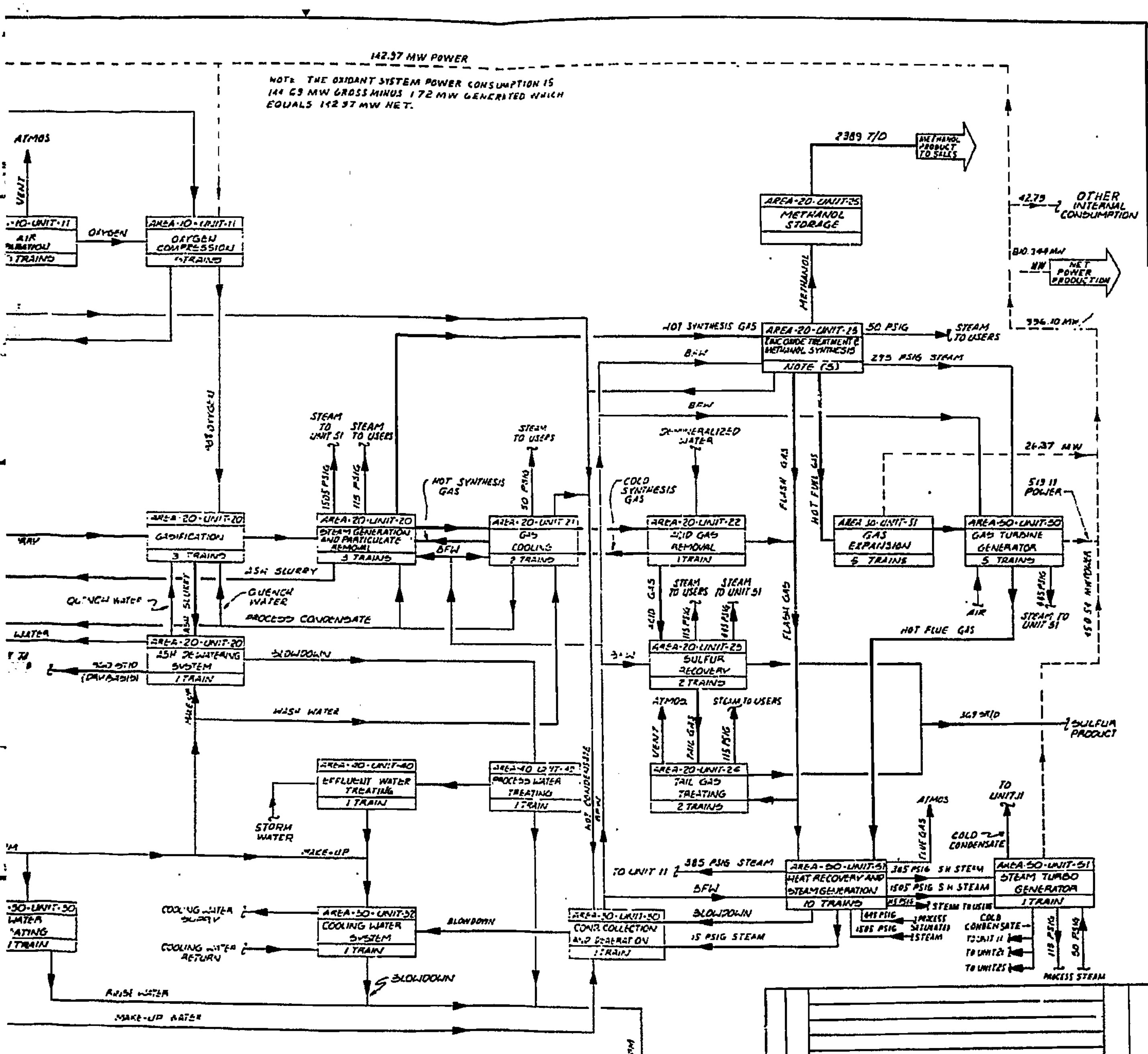
*Five reactor trains and one gas cooling and methanol flash train

**The cooling tower dedicated to the process plant sections is separate from the towers dedicated to the steam turbo-generator condenser



NOTES:

1) OPERATING TRAINS ONLY ARE SHOWN.
2) FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3) EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
4) TWO TRAINS OF COAL GRINDING & ONE TRAIN OF SLURRY PREPARATION ARE PROVIDED.
5) THREE ZINC OXIDE TREATMENT TRAINS, FIVE METHANOL REACTOR TRAINS & ONE EFFLUENT GAS COOLDOWN TRAIN ARE PROVIDED.



NOTES:

1) OPERATING TRAINS ONLY ARE SHOWN.
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3) EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
4) TWO TRAINS OF COAL GRINDING & ONE TRAIN OF SLURRY PREPARATION ARE PROVIDED.
5) THREE ZINC OXIDE TREATMENT TRAINS, FIVE METHANOL REACTOR TRAINS & ONE EFFLUENT GAS COOLDOWN TRAIN ARE PROVIDED.

V. FRIAS	OVERALL BLOCK FLOW DIAGRAM CASE B2-MAXIMUM METHANOL CO-PRODUCTION (1500 PSIG STM)
G. BELMITO	EPRI — PALO ALTO, CALIF.
SHELSER / EPRI	SCALE: NONE — DRAWING NUMBER: 448334-EXTC (ME-B2-1-1 — 04

OXIDANT FEED

Process Flow Diagram EXTC(ME-B2)-11-1 shows the oxidant feed system design used for Case B2. There are five parallel trains each consisting of one air compression system, one air separation plant and one oxygen compression system. No spare train is provided in this section.

Atmospheric air at 14.4 psia, 88°F is compressed to 95 psia in two-stage axial-centrifugal machines 11-1-C-1. The heat of compression is rejected to vacuum condensate water in intercooler 11-1-E-1 and to cooling water in intercooler 11-1-E-2 and aftercooler 11-1-E-3.

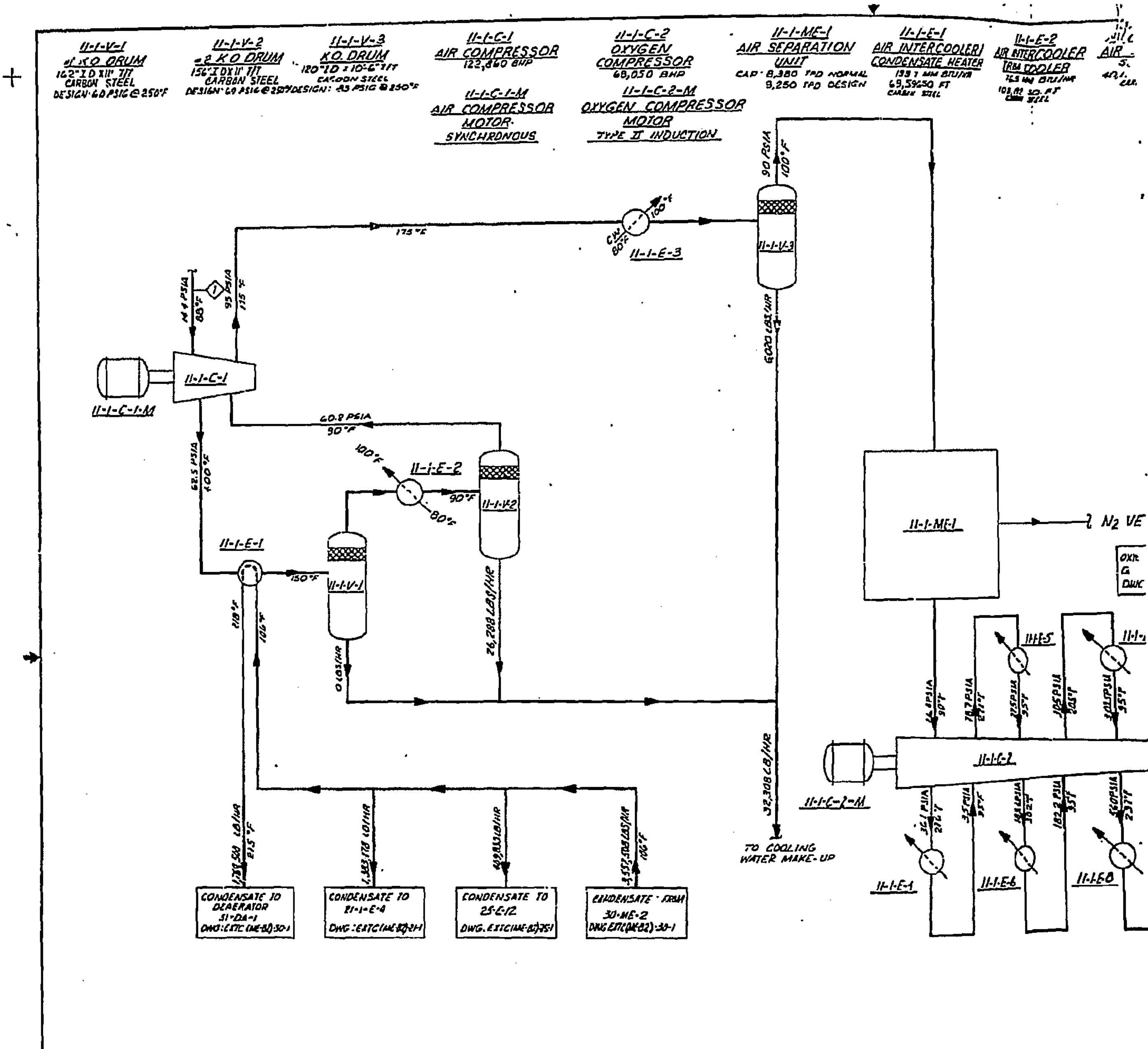
The 122,900 total hp required by the air compressors is supplied by electric motors. The compressed air at 90 psia, 100°F is processed in air separation unit 11-1-ME-1 to produce a total of 8380 tons per day (100 percent O_2 basis) of 98 volume percent oxygen. The air separation unit operating parameters are typical of those for reversing exchanger plant design, which uses turboexpanders for refrigeration. These turboexpanders produce 1.72 MW of power for in-plant consumption.

The 98 mole percent oxygen product at 2 psig, 90°F is compressed to 1120 psig in six stages, prior to being fed to the gasifiers. The interstage heat of compression is rejected to cooling water in interstage coolers 11-1-E-4 through 11-1-E-8. The final discharge temperature is 287°F which is judged to be within design limits for commercial equipment.

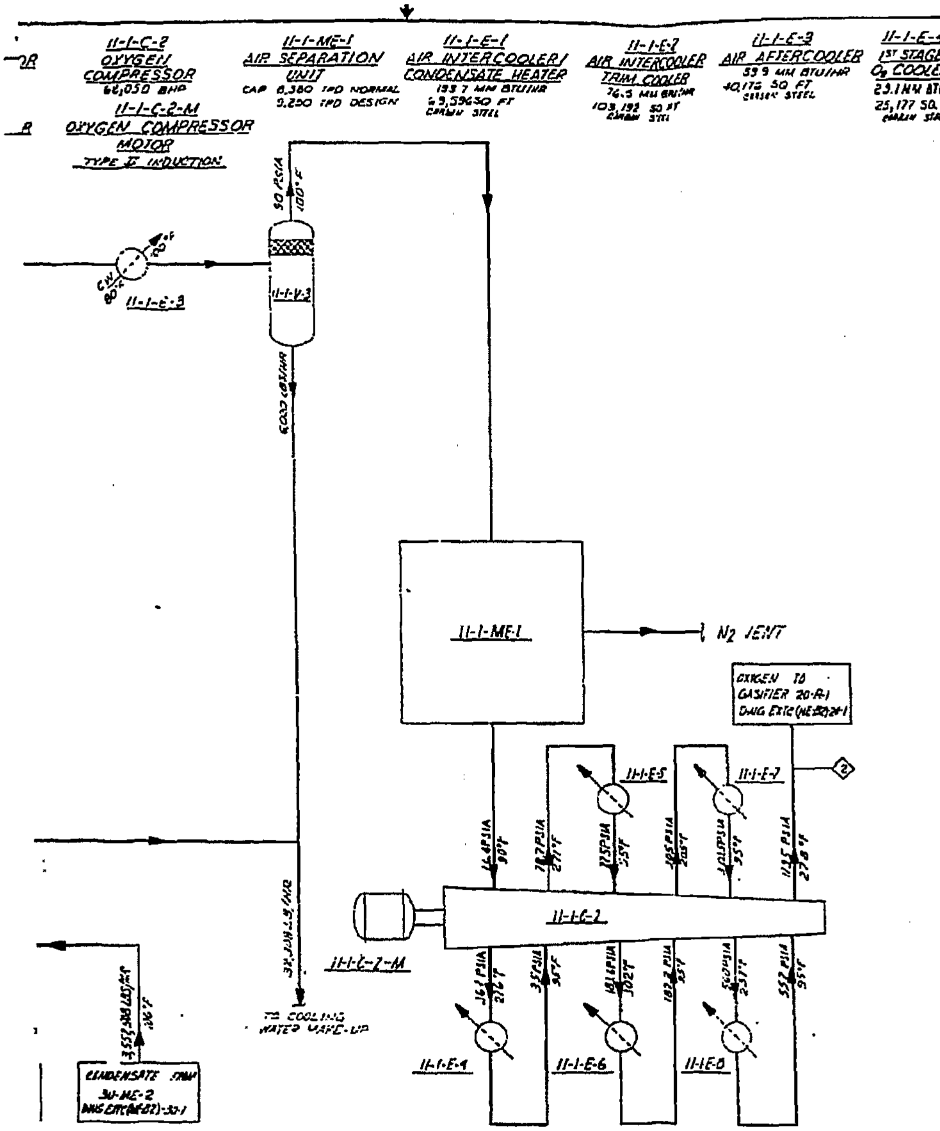
The 68,000 total hp oxidant compression requirement is supplied by electric motors. The startup of the coal gasification unit will be greatly simplified by using electric motors, rather than steam turbines as drivers in the oxidant feed system. Additionally, the steam distribution and condensate collection systems are simplified by concentrating the higher pressure steam usages in the combined-cycle section of the plant.

Equipment Notes

The air compressor and cryogenic air separation plant are commercially available. The oxygen compressor with 1120 psig discharge pressure, is an extension of the commercially-demonstrated centrifugal machine with 950 psig. Attainment of designs based on 1120 psig discharge pressure with current technology is judged to be commercially available. The use of water-cooled oxygen compressors to obtain a 95°F interstage temperature lowers the required compression horsepower. Many of the previous oxidant feed system designs in EPRI studies used air-cooled exchangers for this service. Minimizing power demand is an important consideration, since the oxidant feed system is the largest internal consumer of electric power in the GCC plant. Power requirements may be reduced further, through process optimization, by air separation plant suppliers.



MATERIAL BALANCE				
	PROCESS STREAM			
	① AIR		② OXIDANT	
COMPONENT	MPH	MOL %	MPH	MOL %
O_2	23,714	20.48	21,397.5	98.00
N_2	84,310	76.07	3300	1.50
Ar	1087	0.98	108.5	0.50
H_2O	2,799	2.47	—	—
TOTAL MPH	110,910	100.00	21,836.0	100.00
LB/HR	3,183,198		698,250	
MOL WT	28.70		31.98	



NOTES:
1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FIVE PARALLEL TRAINS
2. ALL FLOW RATES, DUTIES, AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

DOC. NO.	REFERENCE DRAWING	DWG. NO.	REFERENCE DRAWING

C. ABESAMIS

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
CASE B2-MAXIMUM METHANOL CO-PRODUCT/MAXIMUM SNG

EPRI — PALO ALTO, CALIFORNIA

SCALE: NONE — DRAWING NUMBER: 448334-EXTC(ME-B2)-11-1 — REVISION: 01

GASIFICATION AND ASH HANDLING

Process Flow Diagram EXTC(ME-B2)-20-1 shows the gasification, raw gas cooling, and particulate removal steps for Case B2. Three operating trains and one spare train are provided. The ash handling system is a single 100 percent capacity train. The 20-ME-2 box represents proprietary sections of the Texaco coal gasification process containing many units of equipment.

The Texaco gasifier is a vertical cylindrical vessel with a low alloy steel shell. The reaction section of the gasifier, the effluent gas line, and the slag separator are refractory lined.

Coal slurry and oxygen combine at the gasifier burners. Each burner is oriented downward from the top head of the gasifier. The burners have circulating, tempered water-cooling coils.

The gasification section 20-1-R-1 operates at an average pressure of 1000 psig and at temperatures in the range of 2300°F to 2600°F. The ash melts to form slag. The gasification temperature must be sufficiently above the ash flow point to ensure free-flowing molten slag. Most of the coal ash is converted to molten slag and falls into a water quench at the bottom of the gasifier. Part of the coal burns with oxygen to produce a hot flue gas. This combustion reaction provides heat for the endothermic steam/carbon and carbon/CO_2 reactions. The hydrogen and carbon in the coal react to form CO, CO_2, H_2 and a small amount of CH_4. Most of the sulfur is converted to H_2S and COS. Nitrogen in the coal transforms to free nitrogen and a small quantity of ammonia. At the high temperatures prevailing in the gasifier, some of the ammonia in the recycled water is eliminated by dissociation and combustion reactions in the gasifier.

The crude gas product formed in the gasification zone separates from most of the molten ash, leaves the gasifier, and is then quenched with cool, scrubbed, recycle gas below the ash softening point. The amount of this recycle gas is related to ash properties. We have selected an amount which reduces the gas temperature sufficiently below the cool softening temperature to assume it is solidified. If more recycle is actually required to reduce ash fouling, the overall plant efficiency would not be altered significantly. However, due to the higher throughput, the capital costs would change for the affected exchangers and recycle gas compressors. The mixing with recycle gas takes place in a gas quench vessel attached to the gasifier. Both the gasifier and gas quench vessel are vertical cylindrical chambers that are refractory lined to shield the low-alloy vessel shell from high temperatures.

Solids entrained in the exit gas are captured in gas scrubber 20-1-V-4, combined with the slag from all operating gasifiers and processed in a single ash dewatering system 20-ME-2. The resulting ash cake, assumed to contain 40 weight percent water, is transported to a landfill disposal by railroad cars. Overflow from the slag dewatering unit is recycled to the coal slurry and slag quench areas. A slip stream of 107 gpm of reclaimed process water is purged to a proprietary Texaco water treating process for removal of ultrafine slag and soot particles, dissolved metals, formates, sulfides, and ammonia. This water treating unit is included in the general facilities section.

Energy Recovery

Hot crude gas with entrained ash particles enters 20-1-E-1, where 1500 psig saturated steam is generated by recovery of high-level sensible heat. For this feasibility study, the capital cost of these units is based on a horizontal fire-tube-type design. It is recognized that the exchanger configuration ultimately adopted for commercial plants may not be the same as that used in this case. Final designs of the commercial units must accommodate the ash fouling characteristics at high pressure in a reducing environment. These conditions are severe ones, for which more operating experience is required. In the design adopted for this study, the boiler inlet channel is refractory lined and the tubes are constructed of low-alloy steel to resist the temperature and hydrogen content of the crude gas. This heat transfer equipment includes special proprietary features to effectively prevent ash buildup. Soot blowers or other special solids removal systems are not provided. A process contingency of 20 percent has been applied to the estimated installed cost of this unit to reflect the uncertainty associated with this design.

Raw gas leaving the high-pressure saturated steam generator is further cooled by heating methanol synthesis gas from Unit 21 in 20-1-E-2, before being routed to the methanol plant and by heating gas scrubber overhead in 20-1-E-4, before being routed to the COS hydrolysis unit. Saturated medium-pressure (MP) steam at 115 psig is generated by cooling the raw gas further in 20-1-E-3. The ash containing raw gas flows on the tube side to reduce solids deposition. Hot boiler feedwater at HP steam saturation temperature (598°F), and boiler feedwater streams at 347°F are supplied from heat recovery steam generation (HRSG) units located in

Unit 51. Exchanger 20-1-E-3 is a kettle-type boiler with the boiler feedwater fed to the shell side.

Particulate Removal

The particulate bearing raw gas leaves the cooling unit and flows to the gas scrubber 20-1-V-4. Ammonia absorber bottoms and hot process condensate from the gas-cooling area (Flow Diagram EXTC(ME-B2)-21-1) are used for gas scrubbing. Water from 20-1-V-4 is recycled to 20-ME-2. The solids-free raw gas from 20-1-V-4 is reheated to 50°F, above its dew point in 20-1-E-4, and flows to the gas-cooling section Unit 21. In subsequent sections of this report dealing with economics, the reader's attention is called to the fact that costs for equipment included in the proprietary gas cooling and scrubbing units are included in the gas cooling system (Unit 21) costs.

Equipment Notes

The Texaco gasifier is commercially proven for the gasification of liquid hydrocarbons. Commercial experience with coal gasification is limited. One Texaco coal gasifier has been operating for over two years in Germany at about 560 psig. This gasifier handles only six tons per hour of coal, about four percent of the design throughput of the gasifiers used in this study. Another installation for TVA which feeds eight tons per hour at a similar pressure is ready for startup. A gasifier of the size used in this study, but air blown at a lower pressure, is being readied for startup for a confidential U.S. company. The Texaco coal gasification research facility at Montebello, California is presently testing coals in a gasifier which operates at over 1000 psig.

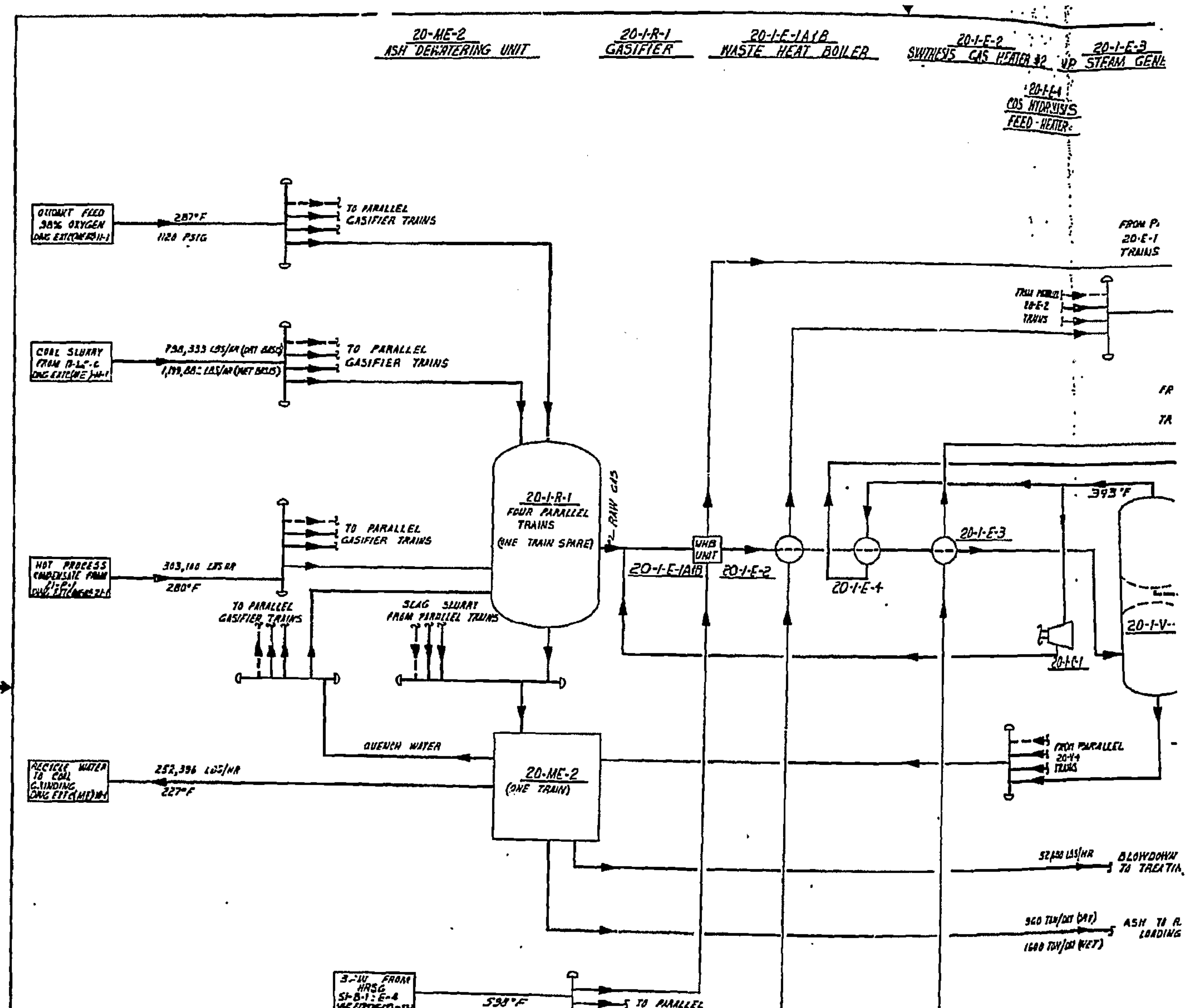
A coal gasifier having approximately one-half of the capacity of the gasifiers used for this study (when corrected for pressure effects) is currently in the final engineering design stages. This gasifier, to be constructed as part of the Cool Water Coal Gasification-Combined Cycle Demonstration Plant by Southern California Edison Company, Texaco, Inc., EPRI, General Electric Company, Bechtel and others, is scheduled to commence operation in 1984. Therefore the gasifiers employed in this study should be considered to be an extension of existing technology, even after the Cool Water plant has operated. The intent of this study is to project equipment performanace and costs for "mature" technology systems, i.e., systems that could exist after approximately five large scale commercial plant have been built and successfully operated.

The slag dewatering system is composed of commercially-proven equipment.

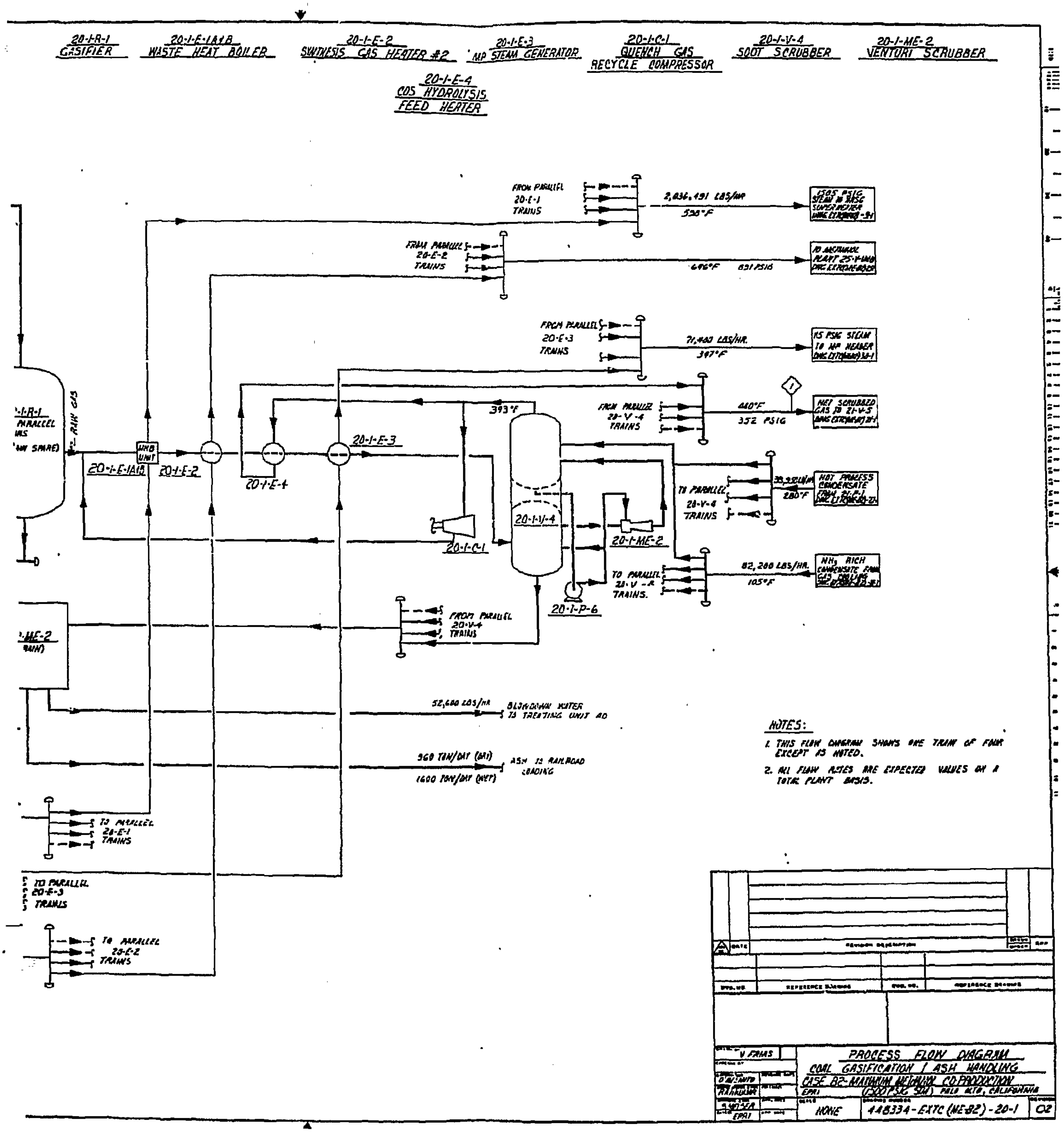
The gas scrubbing unit equipment is commercially available.

The key features in these designs center on the heat transfer equipment used for high-level sensible heat recovery. 1500 psig saturated steam is generated in an unconventional fire-tube boiler. Successful designs of similar items, which process gas containing no entrained solids, have been developed by Steinmuller and by Siegener, both of West Germany. The designs and cost estimates adopted in this study were developed by a major waste heat boiler manufacturer. It is also important to realize that the gas cooler designs for this study are different from those being designed for the Cool Water Demonstration plant.

The gasifier and dry-gas equipment metallurgies are well-defined based on the liquid hydrocarbon partial oxidation experience. Materials of construction for equipment in contact with recovered process condensate are difficult to specify at this stage of development. Actual materials for commercial units will likely be highly specific to the feed coal. The purge rate of process condensate to treating is one parameter which will affect the choice of metallurgies in commercial systems. A detailed study of the cost/benefit relationship between purge rate and material costs is beyond the scope of the present work.



MATERIAL BALANCE		
COMPONENT	PROCESS STREAM NO. 1 WET SCRUBBED GAS MPH	MOL %
CH_4	72.2	0.07
H_2	25974.2	26.89
CO	38216.4	39.56
CO_2	7844.4	8.12
H_2S	904.9	0.94
COS	54.2	0.06
N_2	597.1	0.62
Ar	105.3	0.11
NH_3	178.1	0.18
H_2O	22633.3	23.43
TOTAL	96605.6	100.00
LBS/HR	1935,750	
MOL WT	20.04	


20-1-R-1 GASIFIER
20-1-E-1A1B WASTE HEAT BOILER
20-1-E-2 SYNTHESIS GAS HEATER #2
20-1-E-3 MP STEAM GENERATOR
20-1-C-1 QUENCH GAS RECYCLE COMPRESSOR
20-1-V-4 SOOT SCRUBBER
20-1-ME-2 VENTURI SCRUBBER
20-1-E-4 COS HYDROLYSIS FEED HEATER
FROM PARALLEL 20-E-1 TRAINS
2,836,491 LBS/HR
FROM PARALLEL 20-E-2 TRAINS
FROM PARALLEL 20-E-3 TRAINS
71,400 LBS/HR.
397°F
FROM PARALLEL 20-V-4 TRAINS
440°F
352 PSIG
NET SCRUBBED GAS
393°F
20-1-E-3
20-1-E-1A1B
20-1-E-2
20-1-E-4
20-1-V-4
20-1-C-1
20-1-ME-2
20-1-P-6
TO PARALLEL 20-V-4 TRAINS
280°F
HOT PROCESS CONDENSATE
82,200 LBS/HR.
105°F
NH3 RICH CONDENSATE
TO PARALLEL 20-V-4 TRAINS.
FROM PARALLEL 20-V-4 TRAINS
52,680 LBS/HR
BLOWDOWN WATER TO TREATING UNIT
360 TON/DAY (DRY)
1600 TON/DAY (WET)
ASH TO RAILROAD LOADING
NOTES:
1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF FOUR EXCEPT AS NOTED.
2. ALL FLOW RATES ARE EXPECTED VALUES ON A TOTAL PLANT BASIS.
TO PARALLEL 20-E-1 TRAINS
TO PARALLEL 20-E-3 TRAINS
TO PARALLEL 20-E-2 TRAINS
PROCESS FLOW DIAGRAM
COAL GASIFICATION / ASH HANDLING
CASE B2-MAXIMUM METHANOL CO-PRODUCTION
EPRI
PALO ALTO, CALIFORNIA
NONE
448334-EXTC (ME-B2)-20-1
02

GAS COOLING

Process Flow Diagram EXTC(ME-B2)-21-1 shows one of the two parallel trains in the gas cooling section for Case B2. No spare train is provided.

Clean gasifier effluent from 20-1-E-4 at 440°F enters the gas cooling section and is fed to the COS hydrolysis unit 21-1-V-5A and 5B. In this unit, COS is hydrolyzed to H_2S in the presence of an activated alumina catalyst according to the selective reaction:

$$COS + H_2O \rightarrow H_2S + CO_2 \qquad (4\text{-}1)$$

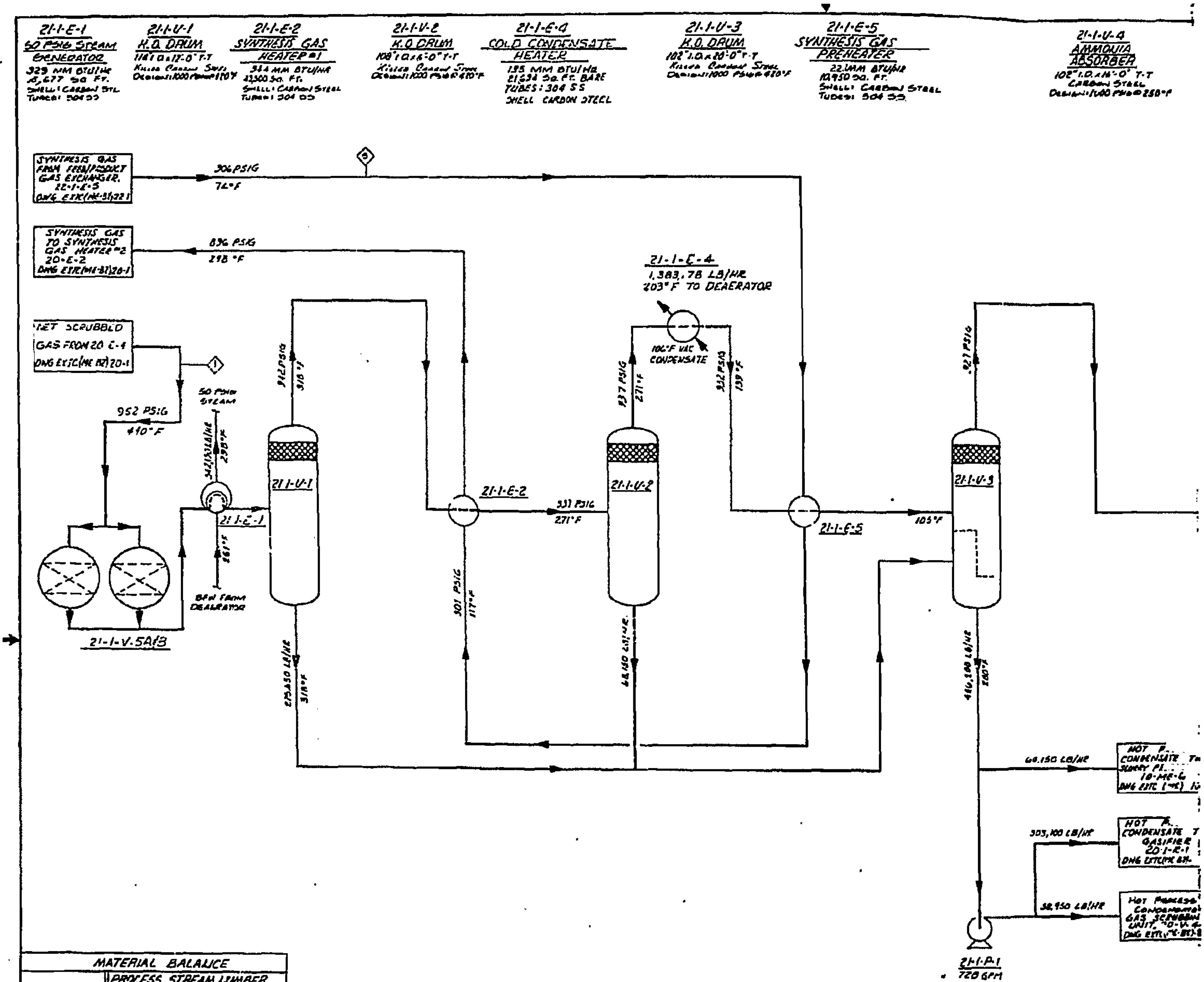
Clean effluent from the hydrolysis unit is cooled to 105°F on the tube side of a series of exchangers 21-1-E-1, 21-1-E-2, 21-1-E-4, and 21-1-E-5. Heat is recovered in exchanger 21-1-E-1 by the generation of saturated 50 psig steam. The effluent, after separation of condensate in the knockout drum 21-1-V-1, is then cooled by exchanging heat against methanol synthesis gas in 21-1-E-2. The condensate produced in cooling is separated in 21-1-V-2. Further gas cooling is obtained in exchanger 21-1-E-4 by heating vacuum condensate. The gas is then cooled by heat exchange in 21-1-E-5 against synthesis gas from the acid gas removal unit. The resultant condensate is separated in knockout drum 21-1-V-3.

Condensate from knockout drums 21-1-V-1 and 21-1-V-2 flows to 21-1-V-3. Some of the combined hot condensate from 21-1-V-3 flows to the slurry preparation unit 10-ME-6 and the remainder is pumped to the particulate scrubber 20-1-V-4 and to the gasifier 20-1-R-1 (Flow Diagram EXTC(ME-B2)-20-1).

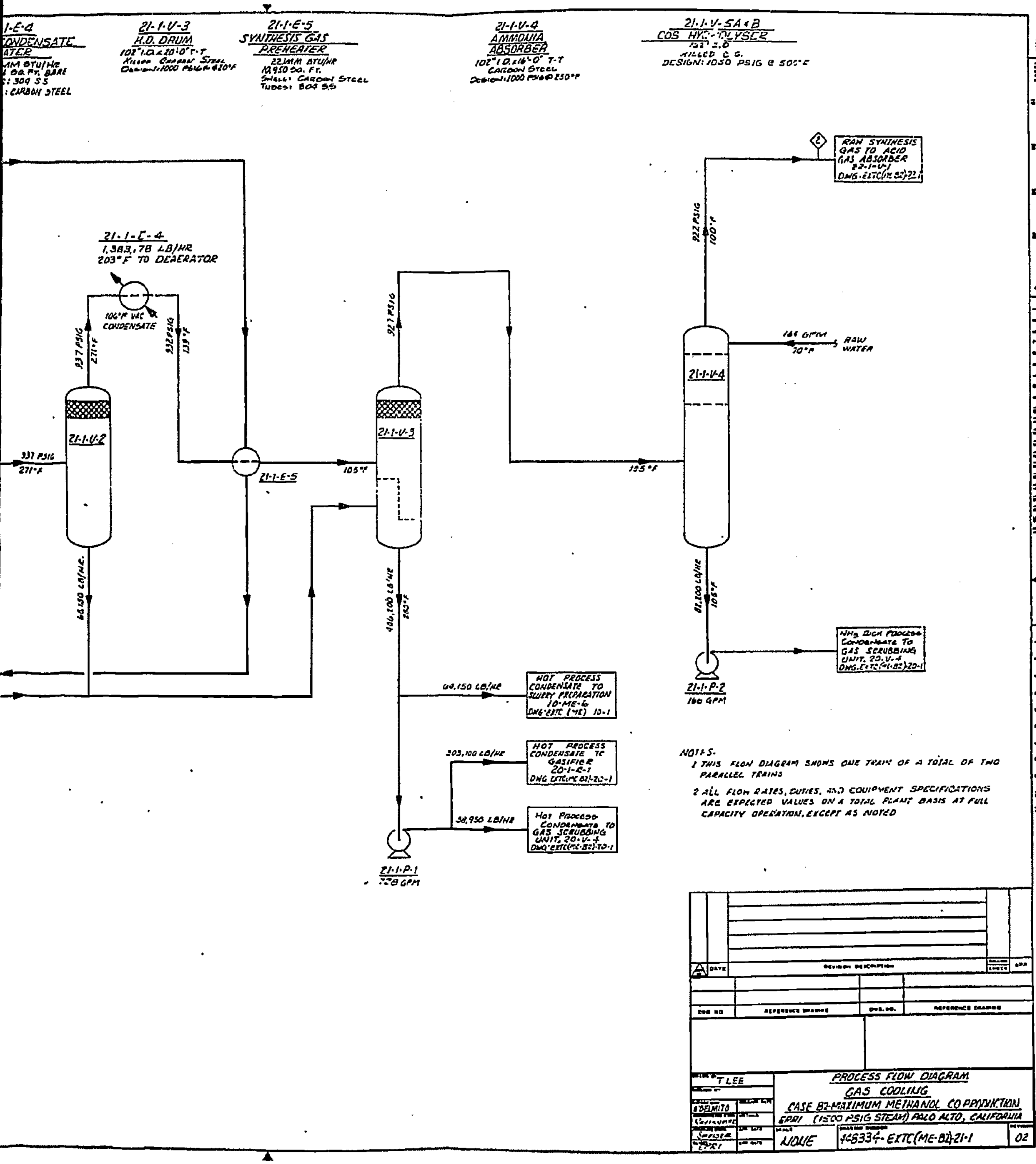
The overhead gas from knockout drum 21-1-V-3 flows to ammonia absorber 21-1-V-4, which contains six sieve-type trays. Ammonia is removed down to one ppm by contacting the gas countercurrently with raw water at 70°F. The essentially ammonia-free overhead gas at 100°F from the absorber then flows to the acid gas removal unit for removal of H_2S and COS. The ammonia-rich process condensate from the bottom of the absorber is pumped to the particulate scrubber 20-1-V-4.

Equipment Notes

All equipment is commercially available. However, the COS hydrolysis catalyst has not yet been demonstrated on a commercial scale.



MATERIAL BALANCE						
	PROCESS STREAM NUMBER					
COMPONENT	① NET SCRUBBED GAS		② AMMONIA SCRUBBER FEED		⑤ SYNTHESIS GAS	
	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	72.8	0.09	72.8	0.10	71.0	0.10
H_2	[illegible]	26.88	[illegible]	36.14	25,466.1	36.54
CO	[illegible]	39.50	38,536.4	59.12	37,175.6	53.35
CO_2	[illegible]	8.18	[illegible]	10.68	6,237.0	8.85
H_2S	96.7	0.11	94.07	1.30	0.12	--
COS	50.8	0.06	0.4	—	0.24	—
N_2	397.1	0.45	397.1	0.61	394.1	[illegible]
Ar	108.3	0.11	108.3	0.15	108.1	0.15
NH_3	118.1	0.13	10	—	—	—
H_2O VAPOR	[illegible]	[illegible]	74.8	0.10	7.5	0.01
TOTAL VAPOR MPH	[illegible]	100.00	[illegible]	100.00	[illegible]	100.00
TOTAL LB/HR	[illegible]		1,327,317		1,415,160	
MOL WT VAPOR	20.06		20.48		19.94	



21-1-V-3
H.D. DRUM
21-1-E-5
SYNTHESIS GAS PREHEATER
22MM BTU/HR
21-1-V-4
AMMONIA ABSORBER
CARBON STEEL
21-1-V-5A & B
COS HYDROLYSER
DESIGN: 1050 PSIG @ 500°F
RAW SYNTHESIS GAS TO ACID GAS ABSORBER
21-1-E-4
1,383,78 LB/HR
203°F TO DEAERATOR
106°F VAC CONDENSATE
937 PSIG
932 PSIG
927 PSIG
922 PSIG
100°F
21-1-V-2
21-1-V-3
21-1-E-5
105°F
21-1-V-4
RAW WATER
NH3 RICH PROCESS CONDENSATE TO GAS SCRUBBING UNIT
21-1-P-2
160 GPM
HOT PROCESS CONDENSATE TO SLURRY PREPARATION
HOT PROCESS CONDENSATE TO GASIFIER
HOT PROCESS CONDENSATE TO GAS SCRUBBING UNIT
21-1-P-1
NOTES.
1 THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF TWO PARALLEL TRAINS
2 ALL FLOW RATES, DUTIES, AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED
T LEE
PROCESS FLOW DIAGRAM
GAS COOLING
CASE B2-MAXIMUM METHANOL CO PRODUCTION
EPRI (1500 PSIG STEAM) PALO ALTO, CALIFORNIA
NONE
02

ACID GAS REMOVAL

Process Flow Diagram EXTC(ME-B2)-22-1 depicts one of the two parallel acid gas removal trains for Case B2. No spare train is provided.

The acid gas removal system employs Allied Chemical Corporation's Selexol process for selective removal of hydrogen sulfide (H_2S) and carbonyl sulfide (COS). The H_2S and COS in the crude gas are absorbed in Selexol solvent to the extent that sulfur in the treated gas is reduced to 5 ppmv.

The cooled, ammonia-free crude gas from the gas cooling unit is further cooled by heat exchange with the treated fuel gas in 22-1-E-5 and flows to the acid gas absorber 22-1-V-1, where it contacts chilled Selexol solvent countercurrently over a packed bed. The treated gas from the top of the absorber flows through a knock-out drum 22-1-V-3 for recovery of solvent and exchanges heat with the feed gas. Then it is routed to gas cooling Unit 21 for further heating.

The rich solvent from the bottom of the absorber is reduced in pressure through a hydraulic turbine 22-1-HT-1. Total hydraulic power from this and another turbine supplies about half of the power required by the lean solvent pump 22-1-P-1. It then flows to an intermediate pressure flash drum 22-1-V-6, where most of the dissolved hydrocarbon gases in the solvent are released. However, because of the selective absorption by the Selexol solvent, most of the dissolved H_2S and COS are retained in solution. The solvent is further let down through a second hydraulic turbine 22-1-HT-2, which supplies additional power to the lean solvent pump. It then flows to a low-pressure flash drum 22-1-V-2, where additional dissolved gases are released. These gases are routed to the acid gas knockout drum 22-1-V-5.

The rich solvent solution from the low-pressure flash drum is heated by exchange with hot regenerated lean solvent in plate exchanger 22-1-E-2 and then flows to the top of the regenerator 22-1-V-4. In the regenerator, the absorbed H_2S and CO_2 are stripped from the solution in a packed bed. Reboil heat is supplied by 115 psig steam in a vertical thermosyphon reboiler 22-1-E-3. Hot regenerated solvent is pumped back to absorber 22-1-V-1 through exchangers 22-1-E-2 and 22-1-E-1. In 22-1-E-2 heat is first exchanged with rich solution to reduce reboiler duty. Then the lean solution is chilled in exchanger 22-1-E-1 to operating temperature with refrigerant from the fluorocarbon refrigeration unit 22-1-ME-1.

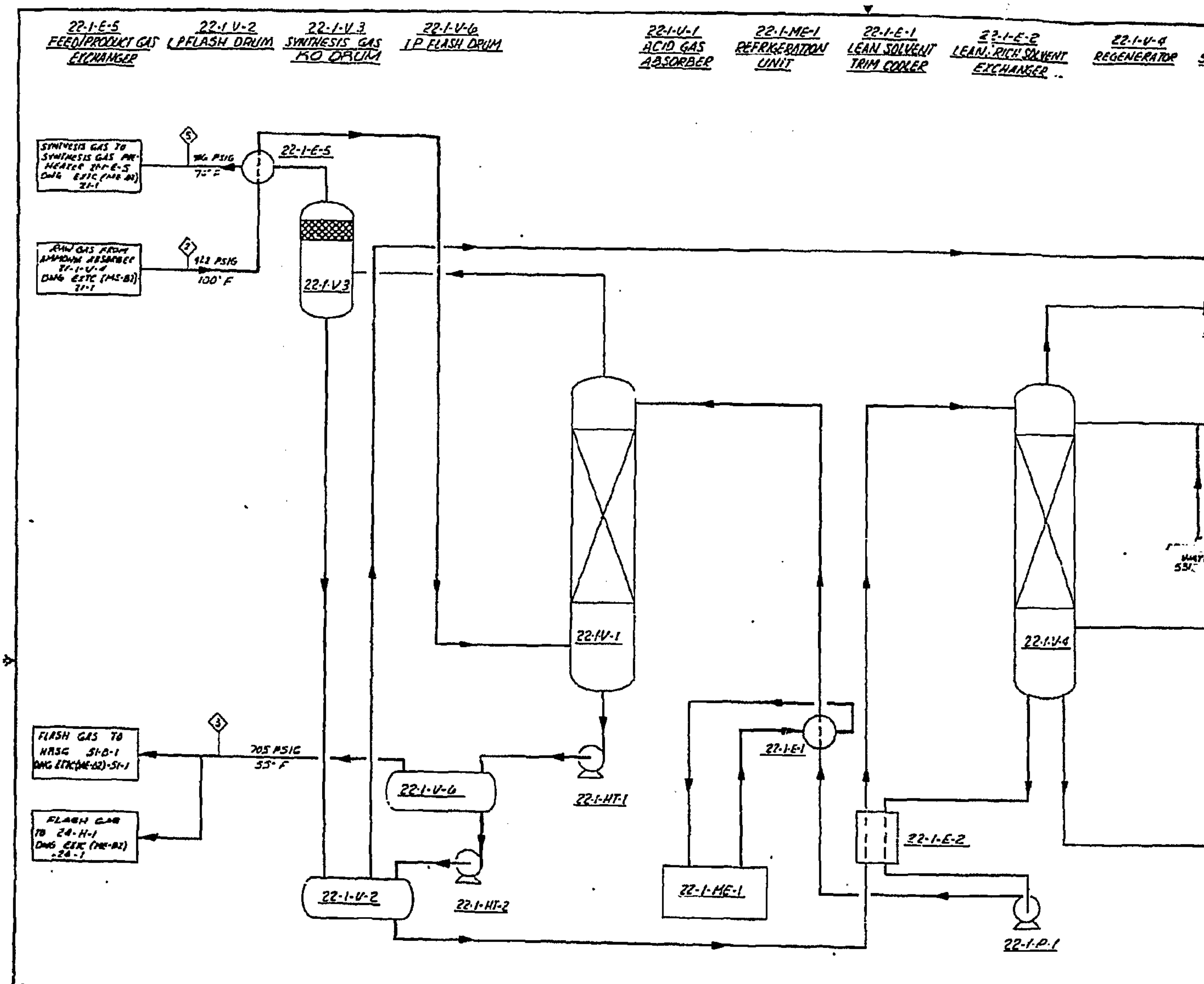
Acid gas from the regenerator overhead is cooled to 120°F in regenerator overhead condenser 22-1-E-4. Condensate resulting from this cooling step is separated in knockout drum 22-1-V-5 and then pumped back to the regenerator by 22-1-P-2. A small stream of demineralized water is added to the condensate at the discharge of 22-1-P-2, to maintain the water balance in the absorption system. The cooled acid gas from 22-1-V-5 contains about 33 percent H_2S on a volume basis and flows to the sulfur recovery unit for further processing.

Refrigeration System

The refrigeration system employed is a typical packaged fluorocarbon unit. The compressor, receiver, and condensing equipment are fabricated on skids and installed near lean solvent chiller 22-1-E-1. The capacity of the unit in each train is about 4100 tons of refrigeration duty.

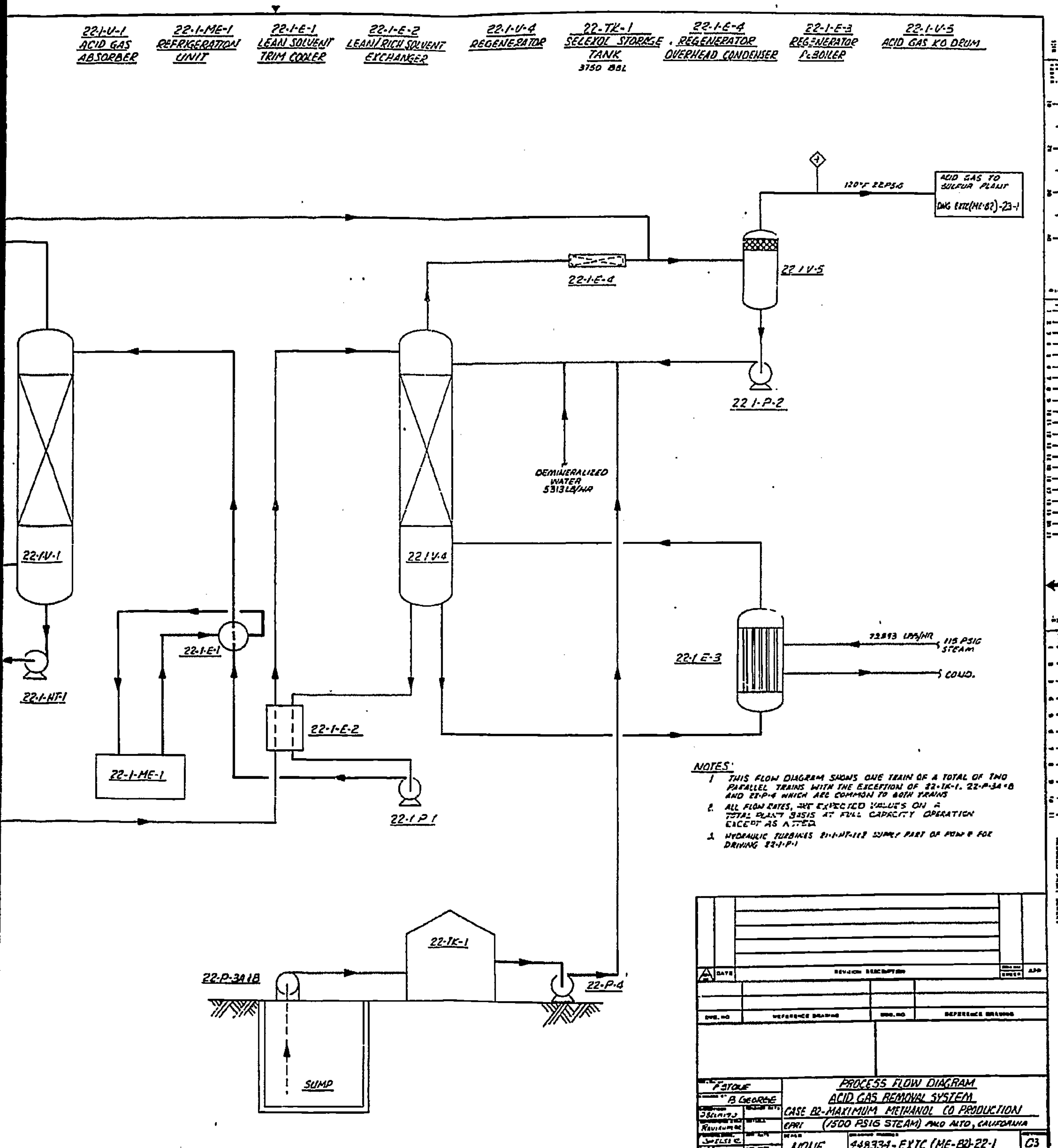
Equipment Notes

The majority of equipment in this section is carbon steel. This equipment has been used in similar service for several years. The use of plate-type exchangers for the lean/rich solvent exchanger service represents a change from previous EPRI designs. These plate-type units are less costly than conventional shell-and-tube exchangers for this service.



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER							
	2 CRUDE GAS		3 FLASH GAS		4 ACID GAS		5 SYNTHESIS GAS	
	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	778	0.10	8.5	0.15	0.9	0.03	710	0.10
H_2	253328	35.14	479	28.46	222	0.77	75,941	[illegible]
CO	[illegible]	[illegible]	493	55.32	1345	4.30	[illegible]	53.53
CO_2	[illegible]	[illegible]	33.4	11.05	15452.2	54.81	62796	8.85
H_2S	9687	1.30	2.47	1.26	9588	33.12	0.12	—
COS	04	—	—	—	.016	0.01	8.24	—
N_2	592.1	0.81	1.5	0.66	0.9	0.03	594	0.84
Ar	1083	0.15	0.1	0.05	0.1	—	108.1	0.15
NH_3	18	—	—	—	18	0.06	—	—
H_2O	748	0.10	0.1	2.05	1928	6.63	9.3	0.01
TOTAL MPH	113,561	100.00	135.87	100.00	[illegible]	100.00	[illegible]	100.00
TOTAL LB/HR	2,527,317		4787		109,665		1,415,160	
	9866							
MOL WT	20.66		24.44		37.91		19.54	



22-1-V-1
ACID GAS ABSORBER
22-1-ME-1
REFRIGERATION UNIT
22-1-E-1
LEAN SOLVENT TRIM COOLER
22-1-E-2
LEAN/RICH SOLVENT EXCHANGER
22-1-V-4
REGENERATOR
22-TK-1
SELEXOL STORAGE TANK
3750 BBL
22-1-E-4
REGENERATOR OVERHEAD CONDENSER
22-1-E-3
REGENERATOR REBOILER
22-1-V-5
ACID GAS KO DRUM
120°F 22PSIG
ACID GAS TO SULFUR PLANT
DWG EXTC(ME-82)-23-1
22-1-E-4
22 1 V-5
22 1-P-2
DEMINERALIZED WATER
5313LB/HR
22-1V-1
22 1 V-4
22-1-E-1
22-1-HT-1
73893 LBS/HR
115 PSIG STEAM
COND.
22-1 E-3
22-1-E-2
22-1-ME-1
22-1 P 1
NOTES:
1 THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF TWO PARALLEL TRAINS WITH THE EXCEPTION OF 22-TK-1, 22-P-3A+B AND 22-P-4 WHICH ARE COMMON TO BOTH TRAINS
2. ALL FLOW RATES, ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED
3. HYDRAULIC TURBINES 22-1-HT-1&2 SUPPLY PART OF POWER FOR DRIVING 22-1-P-1
22-TK-1
22-P-3A+B
22-P-4
SUMP
REVISION DESCRIPTION
DWG. NO
REFERENCE DRAWING
DWG. NO
REFERENCE DRAWING
F STOUE
B GEORGE
PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
CASE B2-MAXIMUM METHANOL CO PRODUCTION
EPRI (1500 PSIG STEAM) PALO ALTO, CALIFORNIA
NONE
448334-EXTC (ME-B2)-22-1
C3

SULFUR RECOVERY

Process Flow Diagram EXTC(ME-B2)-23-1 describes the basic sulfur plant design used. The entire sulfur plant system for Case B2 has three parallel, 50 percent capacity, sulfur recovery trains. Two operating trains and one spare train are provided for increased reliability due to the important environmental requirements this unit fulfills. Sulfur recovery is 173 short tons per day per train.

The sulfur recovery unit is a two-stage acid gas bypass type Claus unit. About one-third of the 120°F gas from the acid gas removal (Selexol) unit is burned in a sulfur furnace 23-1-H-1 to convert H_2S to SO_2. Air for combustion in the furnace is supplied by blower 23-1-BL-1. Heat from the combustion products is recovered by generating 445 psig steam in waste heat boiler 23-1-E-1. The 900°F exhaust gas from the sulfur furnace is mixed with the acid gas bypass stream and the resultant 513°F gas is fed to the sulfur converter No. 1, 23-1-R-1. The amount of acid gas bypassing the furnace is controlled to maintain a ratio of H_2S to SO_2 slightly more than the 2:1 stoichiometric ratio required for the sulfur formation reactions.

H_2S and SO_2 react in the sulfur converter to produce elemental sulfur and water according to the reaction

$$2\ H_2S + 1\ SO_2 \rightarrow 3\ S + 2\ H_2O \qquad (4\text{-}2)$$

This reaction is catalyzed by a bauxite or alumina catalyst contained within the converter. The reaction is exothermic and results in a temperature rise in the gas flowing through the converter. Since this reaction is limited by thermodynamic equilibrium, complete conversion of the H_2S and SO_2 to elemental sulfur is not achieved.

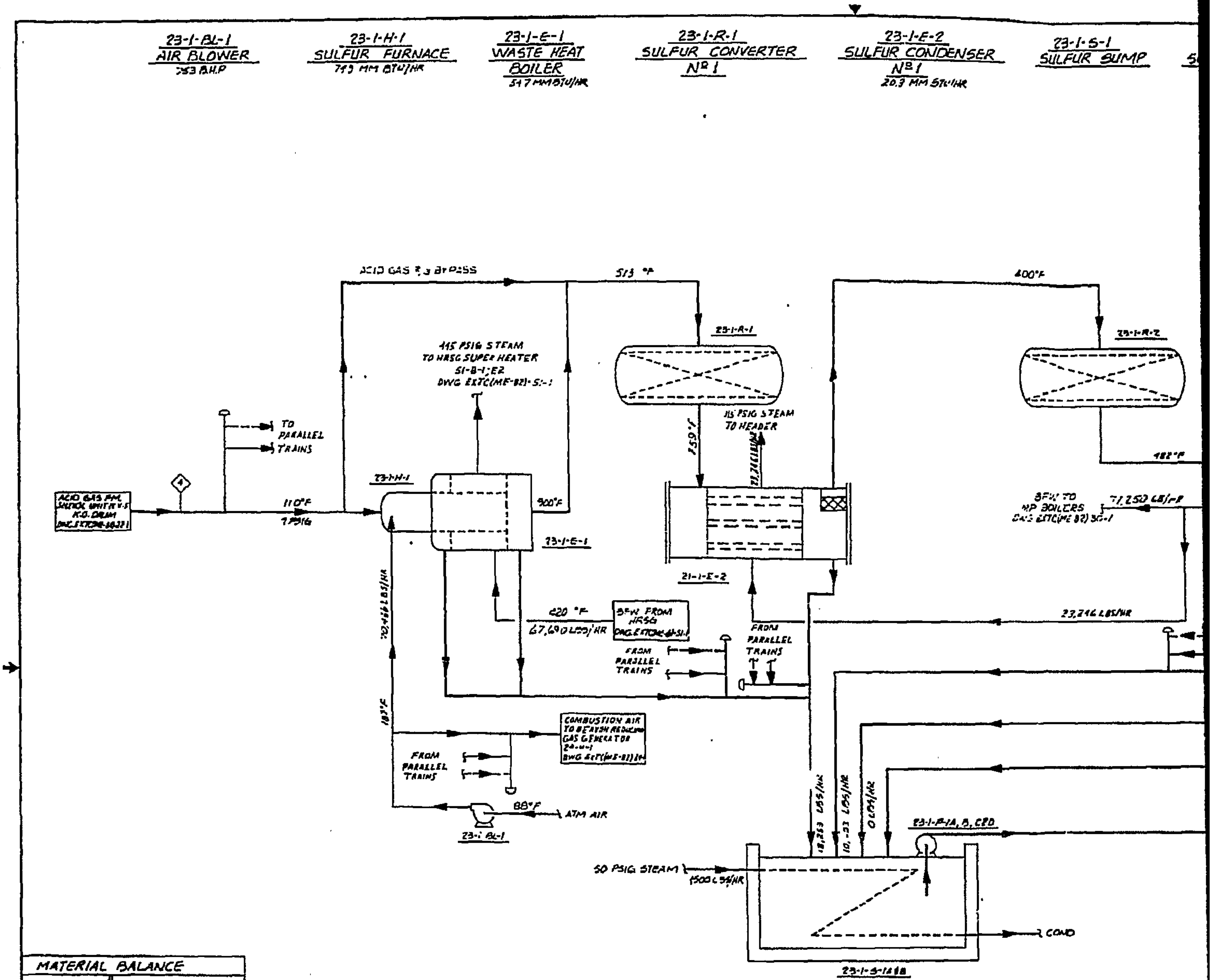
Converter effluent gas is cooled below its sulfur dew point in sulfur condenser 23-1-E-2 by generating 115 psig steam from boiler feedwater. Condensed sulfur flows by gravity to a concrete sulfur sump 23-S-1A&B. Since sulfur is solid at ambient temperature, it must be heated in the sump to take advantage of liquid phase transport to loading facilities. The sump contains low-pressure steam coils to maintain product sulfur in its molten state.

Gases from condenser 23-1-E-2 flow to sulfur converter No. 2, 23-1-R-2 where the sulfur formation reaction proceeds further. Again, the converter effluent is cooled to 285°F in 23-1-E-3 by heat transfer to medium-pressure boiler feedwater. The condensed sulfur then flows to the sulfur sumps.

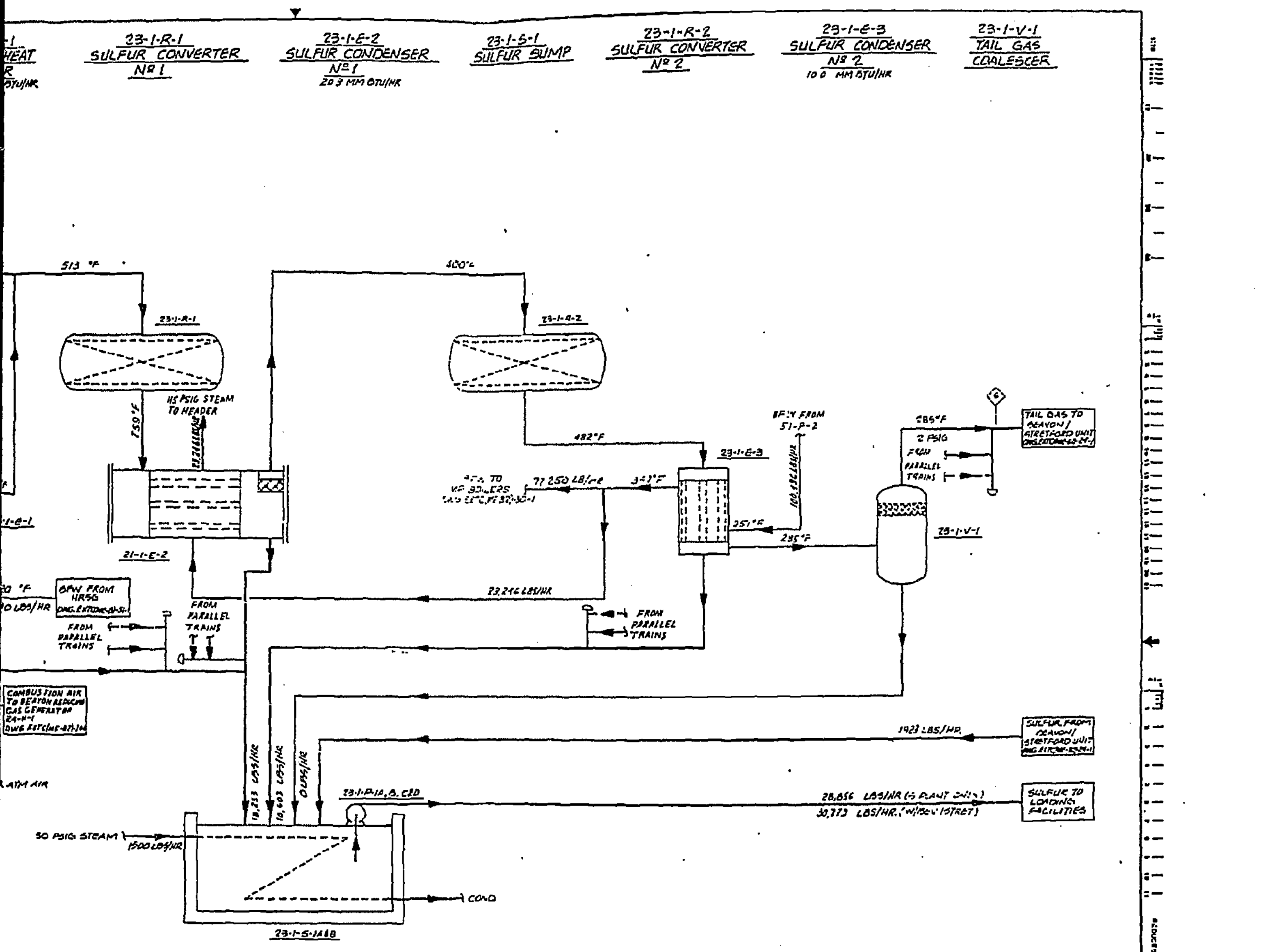
Tail gas at 285°F, still containing about 1870 lb/hr sulfur (mainly as H_2S, with smaller amounts of SO_2, COS, and elemental sulfur) flows through coalescer 23-1-V-1 and then enters Beavon/Stretford Unit 24 for final sulfur recovery to preserve air quality.

Equipment Notes

The Claus sulfur process is established commercially and, consequently, the equipment requirements are well known.



MATERIAL BALANCE				
STREAM Nº	④ ACID GAS FM SELEXOL UNIT		⑥ TAIL GAS TO TG TREATING UNIT	
COMPONENT	MPH	MOL%	MPH	MOL%
H_2	22.2	0.77	17.0	.35
CO	130.3	4.50	83.5	1.71
CO_2	1585.2	54.81	1637.5	33.51
H_2S	958.11	33.12	35.85	.74
COS	0.16	.01	.02	
N_2	0.9	0.03	1901.62	38.93
NH_3	1.8	0.06	1.2	.03
CH_4	0.9	0.03	0.60	.01
H_2O VAPOR	192.8	6.67	1183.0	24.26
SO_2			17.88	.35
S_2				
S_6			13	
S_8			58	.01
TOTAL VAPOR MPH	2892.47	100.00	4872.0	100.00
H_2O LIQ				
TOTAL LBS/HR	109,665		151,375	
MOL WT VAPOR	37.91		31.05	



1 THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE-50% PARALLEL TRAINS (TWO OPERATING AND ONE SPARE)

2 TWO SULFUR SUMPS WILL SERVICE THE THREE TRAINS

3 ALL FLOW RATES, DUTIES, AND EQUIPMENT SPECIFICATIONS ARE TYPICAL VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

4 CONFIGURATION AND SIZE OF EQUIPMENT ARE TYPICAL FOR AN EXPECTED SULFUR RECOVERY RATE OF 173.1 ST/D (93.9 % RECOVERY)

	PROCESS FLOW DIAGRAM
	SULFUR PLANT (TYPICAL)
	CASE-82 MAXIMUM METHANOL CO-PRODUCTION
EPRI	(1500 PSIG STEAM) PALO ALTO, CALIF.
NONE	45334-EXTC (ME-82)-23-1 — 01

TAIL GAS TREATING

Process Flow Diagram EXTC-(ME-B2)-24-1 describes the Beavon/Stretford system design used for the two oxygen-blown GCC plants in the study. As in the sulfur recovery unit, two 50 percent parallel operating trains and a third identical spare train are provided.

The 285°F tail gas from coalescer 23-1-V-1 in the sulfur recovery unit contains unreacted H_2S, SO_2, COS, and the elemental sulfur species S_6 and S_8. To meet strict environmental limits, the gas is processed further to remove these sulfur compounds.

The tail gas treating unit employs a proprietary process called Beavon/Stretford, which is a modification of the well-known Stretford process. The Stretford process is designed to both remove H_2S from atmospheric pressure effluent gas streams, and convert this H_2S to elemental sulfur. The Stretford process is not suitable for handling gas streams which contain substantial amounts of SO_2, COS, S_6 and S_8. The Beavon unit in this process is added to catalytically reduce (or hydrolyze, in the case of COS) these compounds to H_2S.

The reactions occurring over the cobalt molybdate catalyst in the Beavon unit are:

$$SO_2 + 3\ H_2 \rightarrow H_2S + 2\ H_2O \qquad (4\text{-}3)$$

$$COS + H_2O \rightarrow CO_2 + H_2S \qquad (4\text{-}4)$$

$$S_6 + 6\ H_2 \rightarrow 6\ H_2S \qquad (4\text{-}5)$$

$$S_8 + 8\ H_2 \rightarrow 8\ H_2S \qquad (4\text{-}6)$$

The above reactions require hydrogen. A feed gas hydrogen content 1.5 percent in excess of the stoichiometric demand is sufficient to convert essentially all sulfur compounds to H_2S, with the exception of a small residual (perhaps 50 ppmv) of COS. The tail gas stream itself does not contain enough hydrogen or enough carbon monoxide (which can be hydrolyzed to hydrogen) to react with the various sulfur compounds. Rather, a major portion of flash gas from the acid gas removal unit supplies the necessary hydrogen and carbon monoxide. The flash gas is partially combusted in reducing gas generator 24-1-H-1, and then mixed with the tail gas stream. The resulting inlet temperature to the Beavon hydrogenation reactor 24-1-V-7 is 650°F. The sulfur conversion reactions listed above, as well as the following "shift" reaction, take place in 24-1-V-7:

$$CO + H_2O \rightarrow CO_2 + H_2 \tag{4-7}$$

The effluent from 24-1-V-7 is cooled to 400°F through generation of 115 psig steam. Further cooling to 120°F takes place by direct contact with water in the bottom portion of desuperheater/absorber 24-1-T-1. Warm water from the bottom of this vessel is cooled in the fin-fan exchanger 24-1-E-3. Desuperheater/absorber 24-1-T-1 houses two internal heads, in which the water-containing desuperheating section and the Stretford packed bed absorber section are separated.

Stretford solution is pumped from filtrate tank 24-1-TK-1 to the top of the packed-bed absorber, where 99.4 percent or more of the H_2S is reacted with sodium carbonate. Oxidation of the sulfur to the elemental form is facilitated by sodium metavanadate. The absorption and oxidation reactions which occur are as follows:

$$2\ Na_2CO_3 + 2\ H_2S \rightarrow 2\ NaHCO_3 + 2\ NaHS \tag{4-8}$$

$$2\ NaHS + 2\ NaHCO_3 + 4\ NaVO_3 \rightarrow 2\ Na_2CO_3 + H_2O + S_2 + Na_2V_4O_9 + 2\ NaOH \tag{4-9}$$

The absorber provides sufficient retention time to allow the reactions to go essentially to completion. Treated gas, containing much less than 100 ppm total sulfur and traces of CH_4 and CO, is then vented to the atmosphere. The sulfur produced is of high purity, comparable to that produced in the Claus-type sulfur plant.

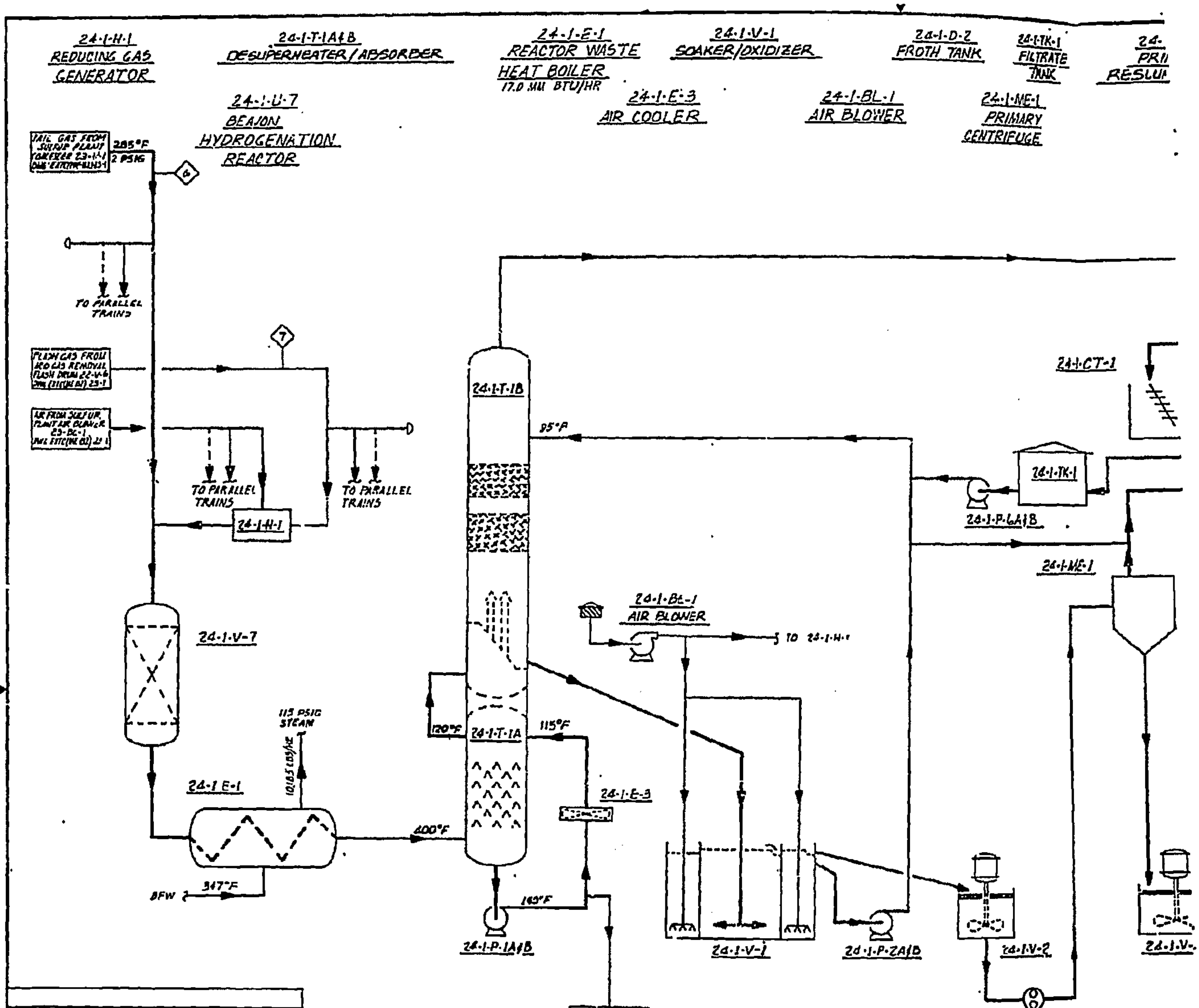
The reacted Stretford solution flows to soaker/oxidizer 24-1-V-1, where the reduced vanadate ($Na_2V_4O_9$) is oxidized to its original form by anthraquinone disulfonic acid (ADA) in the solution. The reduced ADA is subsequently regenerated by air sparged into the tank by blower 24-1-BL-1. The air also provides a medium of flotation for the sulfur which, upon reaching the top of 24-1-V-1, overflows into froth tank 24-1-V-2. The underflow from the soaker/oxidizer is pumped to filtrate tank 24-1-TK-1, via Stretford solution cooling tower 24-1-CT-1, where the heat of oxidation is rejected to the atmosphere.

Sulfur from the froth tank is pumped to the primary centrifuge 24-1-ME-1, which produces a wet sulfur cake that is reslurried in 24-1-V-3 and sent to secondary centrifuge 24-1-ME-2. The filtrate streams from the centrifuges are combined with the soaker/oxidizer underflow.

The sulfur from the secondary centrifuge is reslurried in 24-1-V-4 and pumped to the sulfur separator 24-1-EJ-1, where sulfur is melted with heat supplied by 115 psig steam in coils. Molten sulfur (1936 lb/hr) is separated from the slurry medium (primarily water) in sulfur separator 24-1-V-5. From 24-1-V-5 it flows by gravity into one of the two sumps located in Unit 23. The decanted water flows to flash drum 24-1-V-6 and then back to the secondary reslurry tank. Because certain side reactions degrade the Stretford solution, a small stream of liquid is continuously discarded from the system and pumped to effluent water treating Unit 40.

Equipment Notes

The marriage of the Beavon and Stretford processes is a fairly recent development, but it has been demonstrated commercially on a much smaller scale than is proposed here. This specific equipment has been operating successfully in many plants. Most of the plant is constructed of carbon steel. Certain sections of the Stretford unit are usually coated with coal tar epoxy to prevent corrosion by deposited sulfur, and the sulfur melter is fabricated of stainless steel.

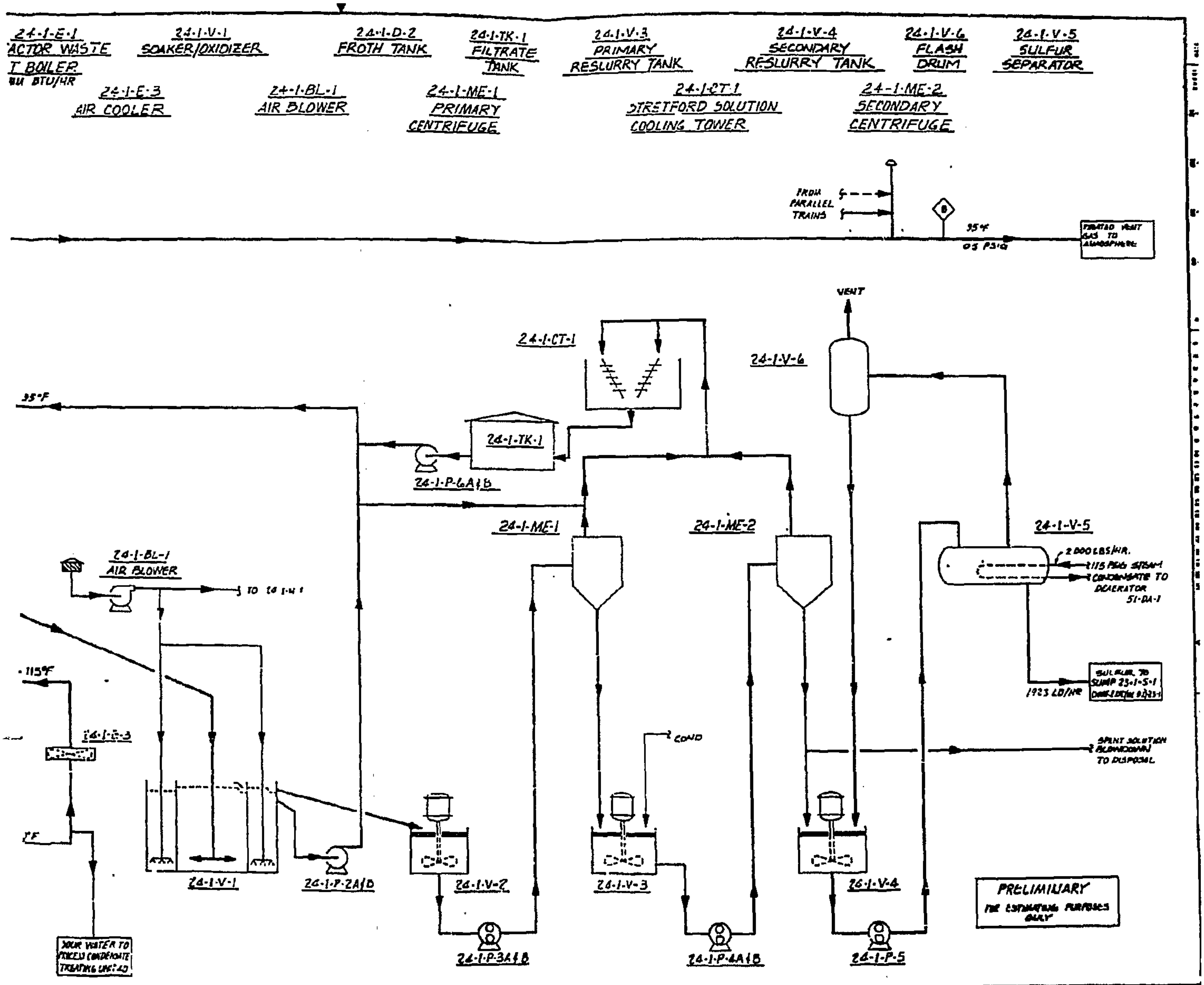


STREAM NO. / COMPONENT	⟨6⟩ TAIL GAS FM SULFUR PLANT MPH	MOL%	⟨7⟩ FLASH GAS FM SELEXOL MPH	MOL%	⟨8⟩ ■ VENT GAS MPH	MOL%
H_2	17.0	.35	36.6	24.36	13.4	.31
CO	83.5	1.83	84.26	56.08	2.59	.06
CO_2	1632.5	36.68	25.86	17.21	1823.69	42.10
H_2S	35.85	.72	2.04	1.36	.36	.01
COS	.02	—				
N_2	1901.62	38.26	1.17	.78	2144.41	47.50
NH_3	1.2	.05				
CH_4	0.6	.01	.25	.17	.47	.01
H_2O VAPOR	1182.0	23.75	.08	.05	385.96	8.75
[illegible]	17.00	.34				
[illegible]						
[illegible]	1.3	—				
[illegible]	.58	.01				
TOTAL VAPOR MPH	4877.0	100.00	151.26	100.00	4403.88	100.00
H_2O LIQ	—					
TOTAL LB/HR	151,275		3680		142,519	
MOL WT VAPOR	31.05		24.49		32.50	

■ STATED AS H_2S BUT CONSISTS OF TOTAL SULFUR.

NOTES:

1. THIS PROCESS FLOW DIAGRAM TRAIN OF A TOTAL OF THREE (TWO OPERATING AND ONE
2. ALL FLOW RATES, AND EQUIP. ARE TYPICAL VALUES ON BASIS AT FULL CAPACITY AS NOTED
3. CONFIGURATION AND SIZE OF FOR EXPECTED RATE OF 11. RECOVERY.



NOTES:

1. THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE 50% PARALLEL TRAINS (TWO OPERATING AND ONE SPARE).
2. ALL FLOW RATES, AND EQUIPMENT SPECIFICATIONS ARE TYPICAL VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED
3. CONFIGURATION AND SIZE OF EQUIPMENT ARE TYPICAL FOR AN EXPECTED RATE OF 11.5 T/D PER TRAIN SULFUR RECOVERY.

T.LEE

PROCESS FLOW DIAGRAM
BEAVON/STRETFORD UNIT (TYPICAL)
CASE B2-MAXIMUM METHANOL CO-PRODUCTION
EPRI (1500 PSIG STEAM)

NONE | 448334-EXTC(ME-B2)24-1 | 02

ZINC OXIDE TREATMENT

Process Flow Diagram EXTC(ME-B2)-25-1 shows the zinc oxide treatment unit for Case B2. There are three parallel trains, and each train has two zinc oxide vessels in series. No spare vessels are provided.

The heated gas from 20-1-E-2 flows through zinc oxide beds 25-1-V-1A&B which remove nearly all of the remaining sulfur compounds according to the following reactions:

$$H_2S + ZnO \rightarrow ZnS + H_2O \qquad (4\text{-}10)$$

$$COS + ZnO \rightarrow ZnS + CO_2 \qquad (4\text{-}11)$$

Essentially all of the H_2S is removed, and approximately 90 percent of the entering COS is removed. The effluent gas contains less than 0.5 ppmv sulfur as required by the methanol plant catalyst.

Although the zinc oxide treatment unit is shown with the methanol plant, its cost is stated separately in the economic tables of the report.

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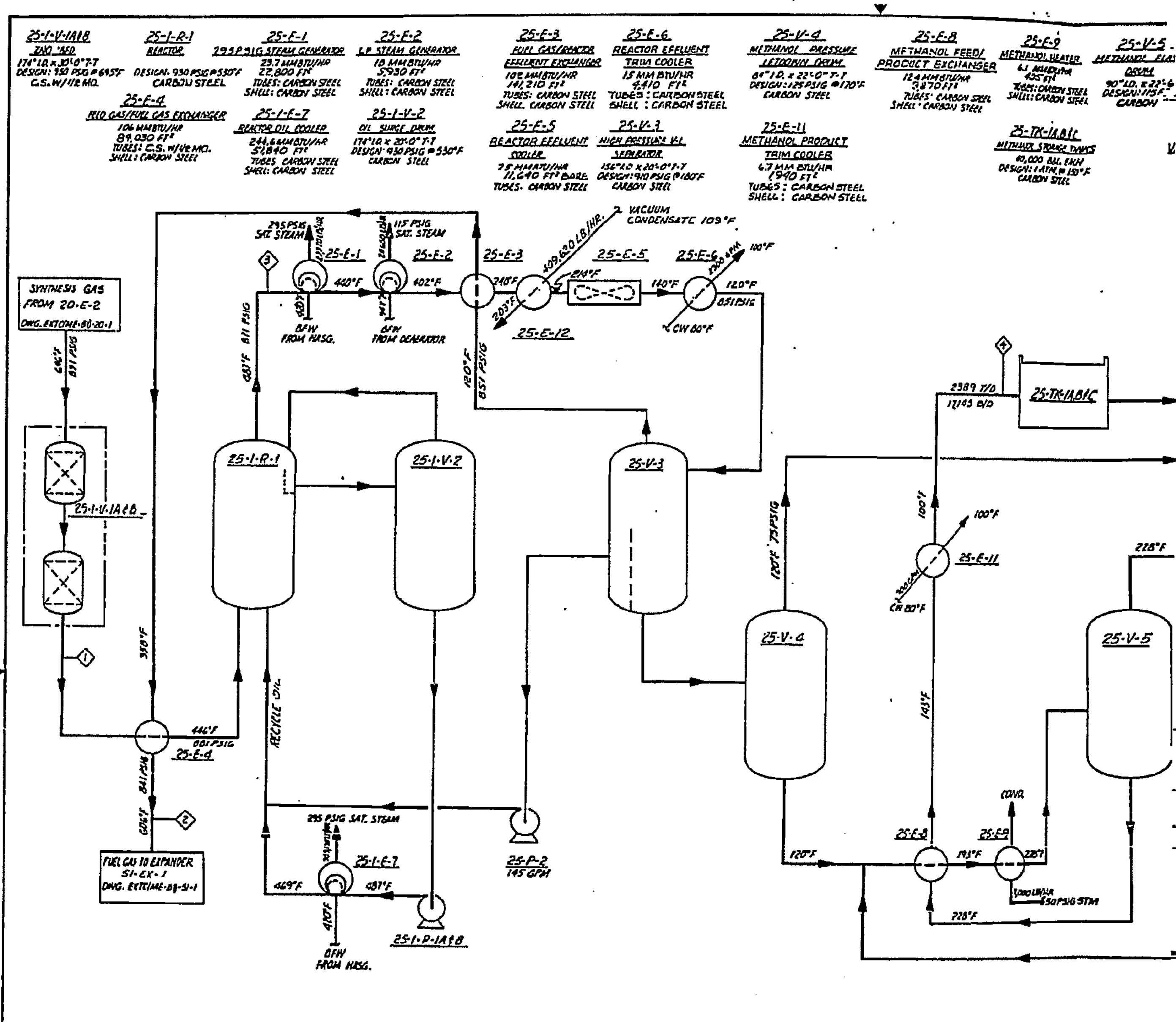
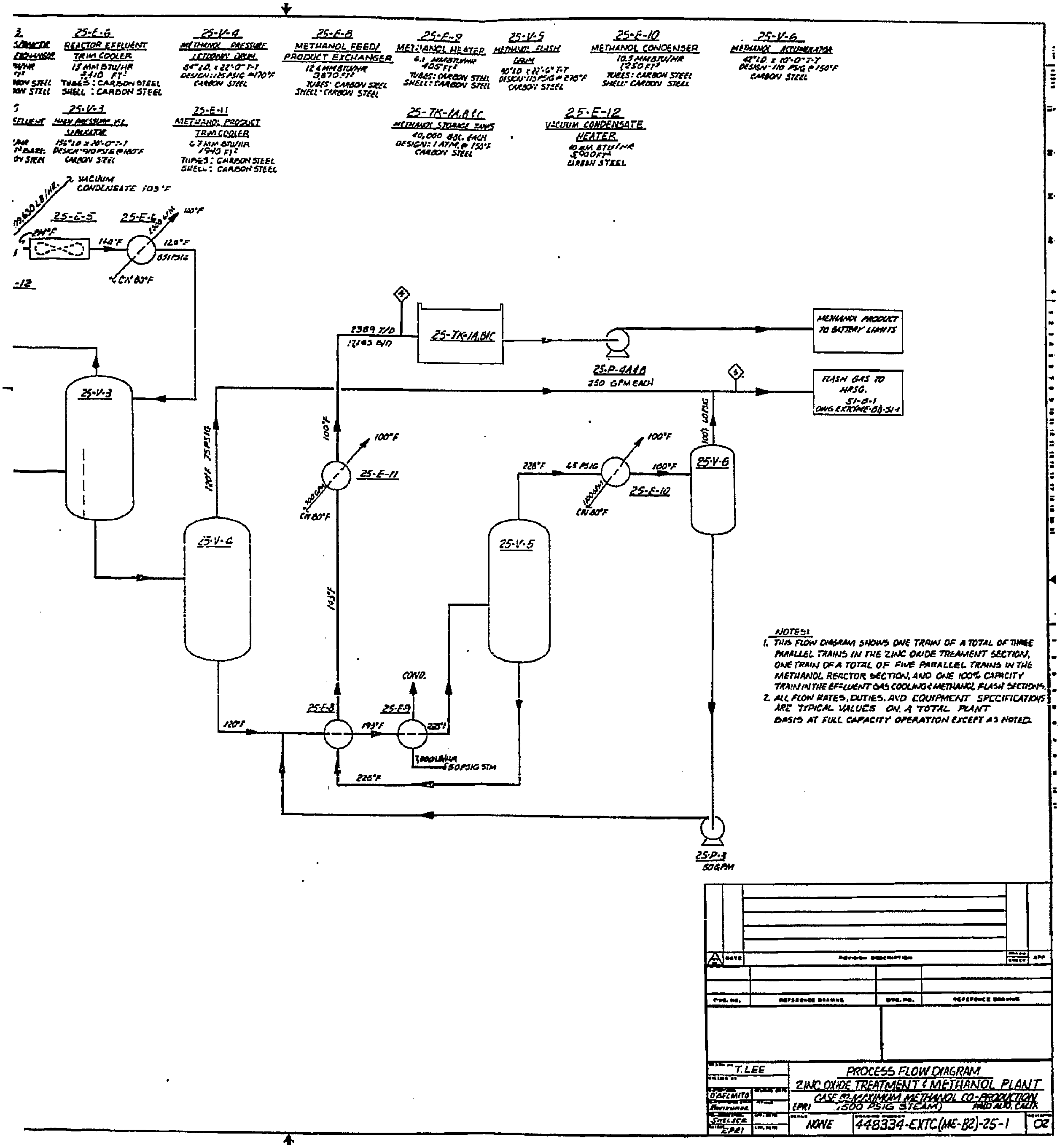


COMPONENT	① SYNTHESIS GAS MPH	MOL%	② FUEL GAS MPH	MOL%	③ REACTOR EFFLUENT MPH	MOL%	④ METHANOL PRODUCT MPH	WT%	⑤ FLASH GAS MPH	MOL%
CO	37,995.8	[illegible]	3,652.0	[illegible]	[illegible]	[illegible]	0.2	—	78.9	[illegible]
H_2	25,341	35.9	[illegible]	[illegible]	[illegible]	22.7	—	—	15.8	8.95
CO_2	6,279.6	8.85	6,245.0	[illegible]	6,279.6	[illegible]	5.6	0.10	69.0	39.09
CH_4	71.0	0.10	74.6	[illegible]	71.0	0.12	—	—	0.6	[illegible]
N_2	594.3	0.84	593.7	1.19	594.9	1.02	—	—	1.2	0.68
Ar	108.	0.15	108.1	0.21	108.1	0.19	—	—	—	—
CH_3OH	—	—	477.8	0.91	6,257.6	10.75	5,773.2	97.83	11.6	6.57
H_2O	9.3	0.01	98.5	0.19	133.3	0.23	35.3	0.60	.01	0.06
HIGHER ALCOHOLS	—	—	—	—	83.1	0.15	83.1	1.41	—	—
PROCESS OIL	—	—	—	—	168.3	0.29	—	—	—	—
TOTAL MOLS/HR	70,562.8	100.00	9,258.6	100.00	[illegible]	100.00	[illegible]	100.00	176.5	100.00
LB/HR	1,405,143		1,219,189		1,459,065		190,284		5675	
MW	19.34		23.96		25.07		32.27		32.15	


25-E-6
REACTOR EFFLUENT TRIM COOLER
15 MM BTU/HR
TUBES : CARBON STEEL
SHELL : CARBON STEEL
25-V-4
METHANOL PRESSURE LETDOWN DRUM
CARBON STEEL
25-E-8
METHANOL FEED/ PRODUCT EXCHANGER
12.6 MM BTU/HR
3870 FT²
TUBES: CARBON STEEL
SHELL: CARBON STEEL
25-E-9
METHANOL HEATER
405 FT²
TUBES: CARBON STEEL
SHELL: CARBON STEEL
25-V-5
METHANOL FLASH DRUM
CARBON STEEL
25-E-10
METHANOL CONDENSER
10.5 MM BTU/HR
(250 FT²)
TUBES: CARBON STEEL
SHELL: CARBON STEEL
25-V-6
METHANOL ACCUMULATOR
CARBON STEEL
25-V-3
HIGH PRESSURE MEL SEPARATOR
CARBON STEEL
25-E-11
METHANOL PRODUCT TRIM COOLER
6.7 MM BTU/HR
1940 FT²
TUBES: CARBON STEEL
SHELL: CARBON STEEL
25-TK-1A,B,&C
METHANOL STORAGE TANKS
40,000 BBL EACH
DESIGN: 1 ATM @ 150°F
CARBON STEEL
25-E-12
VACUUM CONDENSATE HEATER
40 MM BTU/HR
5900 FT²
CARBON STEEL
VACUUM CONDENSATE 105°F
25-E-5
25-E-6
140°F
120°F
CW 80°F
2389 T/D
17,165 B/D
25-TK-1A,B,C
METHANOL PRODUCT TO BATTERY LIMITS
25-P-4A&B
250 GPM EACH
FLASH GAS TO HRSG.
25-V-3
25-V-4
25-V-5
25-V-6
100°F
25-E-11
CW 80°F
228°F
65 PSIG
100°F
25-E-10
CW 80°F
193°F
COND.
25-E-8
25-E-9
120°F
193°F
225°F
220°F
25-P-3
50 GPM
NOTES:
1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE PARALLEL TRAINS IN THE ZINC OXIDE TREAMENT SECTION, ONE TRAIN OF A TOTAL OF FIVE PARALLEL TRAINS IN THE METHANOL REACTOR SECTION, AND ONE 100% CAPACITY TRAIN IN THE EFFLUENT GAS COOLING & METHANOL FLASH SECTIONS.
2. ALL FLOW RATES, DUTIES, AND EQUIPMENT SPECIFICATIONS ARE TYPICAL VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
T. LEE
PROCESS FLOW DIAGRAM
ZINC OXIDE TREATMENT & METHANOL PLANT
CASE B2-MAXIMUM METHANOL CO-PRODUCTION (1500 PSIG STEAM)
EPRI
PALO ALTO, CALIF.
NONE
448334-EXTC(ME-B2)-25-1
02

METHANOL PLANT

Process Flow Diagram EXTC(ME-B2)-25-1 also shows the methanol plant for Case B2. There are five parallel reactors and circulating oil loops, a single train for reactor effluent gas cooling and product methanol flashing. No spare trains are provided.

The methanol plant employs the Chem Systems process for producing methanol in a three-phase fluidized bed reactor.

The synthesis gas from the zinc oxide treatment unit is cooled to methanol reaction temperature in 25-E-4 by heating unconverted reactor product gas, before flowing to the methanol synthesis reactor 25-1-R-1.

The gas flows upward in the reactor concurrent with an inert hydrocarbon liquid (oil) containing fluidized catalyst particles. The liquid limits the temperature rise as it absorbs the heat liberated by the reaction. Phase separation between solid, liquid and vapor occurs at the top of the reactor. The catalyst remains in the reactor. The hydrocarbon liquid, separated from both catalyst and vapor, is recirculated by pumps 25-1-P-1A&B through exchanger 25-1-E-7 to the bottom of the reactor. Cooling of the oil occurs in 25-1-E-7 by generation of 295 psig steam.

The reactor effluent gas is cooled in a series of exchangers to condense the crude methanol and any hydrocarbon liquid that has vaporized. The hot gas generates 295 psig steam in 25-E-1 and 115 psig steam in 25-E-2. Further cooling is obtained by heat exchange with product fuel gas in 25-E-3, by exchange with vacuum condensate in 25-E-12, and then by air fan 25-E-5. Final cooling is done with cooling water in 25-E-6.

The cooled reactor effluent flows to separator 25-V-3 where the unconverted product gas, crude methanol, and hydrocarbon liquid are separated. The product fuel gas is heated in 25-E-3 and 25-E-4 and then flows to Unit 21 for further heating. The hydrocarbon liquid is recycled back to the reactor by 25-P-2. The crude methanol is routed to the product flash section to reduce the vapor pressure for storage by releasing dissolved gases.

In the product flash section (not a Chem Systems design), the high-pressure methanol is flashed at low-pressure in flash drum 25-V-4. The flash gas is routed to the heat recovery steam generator (HRSG) in Unit 51 where it is burned. The low-pressure crude methanol is heated by exchange with product methanol in 25-E-8 and is further heated with 50 psig steam in 25-E-9.

The hot crude methanol is then flashed in flash drum 25-V-5. The flash gas is cooled with cooling water in 25-E-10 and then flows to knockout drum 25-V-6. The overhead gas from 25-V-6 is routed to the HRSG where it is burned. Methanol recovered in 25-V-6 is rich in dissolved gases and is recycled, joining the low-pressure crude methanol before 25-E-8.

The hot product methanol from flash drum 25-V-5 is cooled in 25-E-8 and further cooled by cooling water in 25-E-11. The cooled product methanol is stored in three 40,000 bbl floating roof tanks 25-TK-1A,B&C. Transfer pumps 25-P-4A&B are provided to transfer the methanol product to battery limits.

Equipment Notes

The Chem Systems liquid phase methanol process is in the early stages of development. Early work with a bench scale unit has demonstrated the feasibility of the process using a commercial catalyst. A process development unit has been constructed and operated for short periods of time. Chem Systems is presently working on solving catalyst deactivation and attrition problems. Use of catalyst powder, instead of tablets directly in a catalyst liquid system, is being pursued.

The majority of the equipment in this section is carbon steel and, with the exception of the reactor and its phase separation system, has been used in similar service for many years.

STEAM, BOILER FEEDWATER, AND CONDENSATE

Process Flow Diagram EXTC(ME-B2)-30-1 schematically represents the steam, boiler feedwater (BFW), and condensate systems for Case B2.

The process plant steam generation is integrated with the combined-cycle system. The steam system operates at six levels:

- High-Pressure (HP) — 1450 psig, 900°F at the 51-T-1A turbine inlet
- Intermediate Pressure (IP) — 445 psig, 900°F at the 51-T-1B turbine inlet
- Medium-Pressure (MP) — 295 psig at gas turbine combustor inlet
- Medium-Pressure (MP) — 115 psig
- Low-Pressure (LP) — 50 psig
- Very Low-Pressure (VLP) — 15 psig for consumption in deaerator

High-pressure (HP) steam generation is carried out in the gas cooling unit 20-1-E-1 with additional generation and superheating in the heat recovery steam generator (HRSG) 51-1-B-1 of gas turbine 50-1-GT-1. There are five gas turbines and each has two attendant HRSGs. The saturated HP steam from 20-1-E-1 combines with saturated HP steam from the HRSG HP evaporator 51-1-B-1:E-3. The combined stream is superheated to 900°F in 51-1-B-1:E-1 and used to drive the single back-pressure-type turbine 51-T-1A. The HP end of turbine 51-T-1A, a machine of 82.1 percent isentropic efficiency, takes steam at 1450 psig, 900°F and exhausts at 445 psig.

Saturated intermediate pressure (IP) steam at 445 psig is obtained from the IP steam generators located in the sulfur plant, and from the gas turbine air cooler 50-1-E-1. The saturated IP steam, together with the exhaust steam from 51-T-1A is superheated to 900°F in the HRSG reheater 51-1-B-1:E-2. The superheated IP steam at 385 psig, 950°F is then used in the IP end of 51-T-1B, a machine of 85.7 percent isentropic efficiency. The low-pressure end of 51-T-1B exhausts steam at 85.9 psig.

Medium-pressure saturated steam at 295 psig is generated in the methanol plant and is combined with the fuel gas prior to entering the combustor of gas turbine 50-1-GT-1.

Steam for the 115 psig header is obtained from steam generators in gas cooling unit 20-1-E-3, from the sulfur plant, and from the tail gas treating unit. A portion of the 115 psig steam is supplied to the sulfur heater and the acid gas removal unit reboiler. The remainder is combined with 51-T-1B exhaust at 93.8 psig for consumption in MP turbine 51-T-2 and in BFW pump turbine 51-T-4. The MP turbine and the BFW pump driver are condensing turbines exhausting at 2-1/2 inches Hg absolute. 51-T-2 has an isentropic efficiency of 87.4 percent.

The 50 psig steam header is supplied by steam generated in gas cooling unit 21-1-E-1. The 50 psig steam is primarily used in condensing turbine-generator 51-T-3 for making additional electric power while small amounts are used for steam tracing, process water treating, methanol flashing, the sulfur pit, and deaerator.

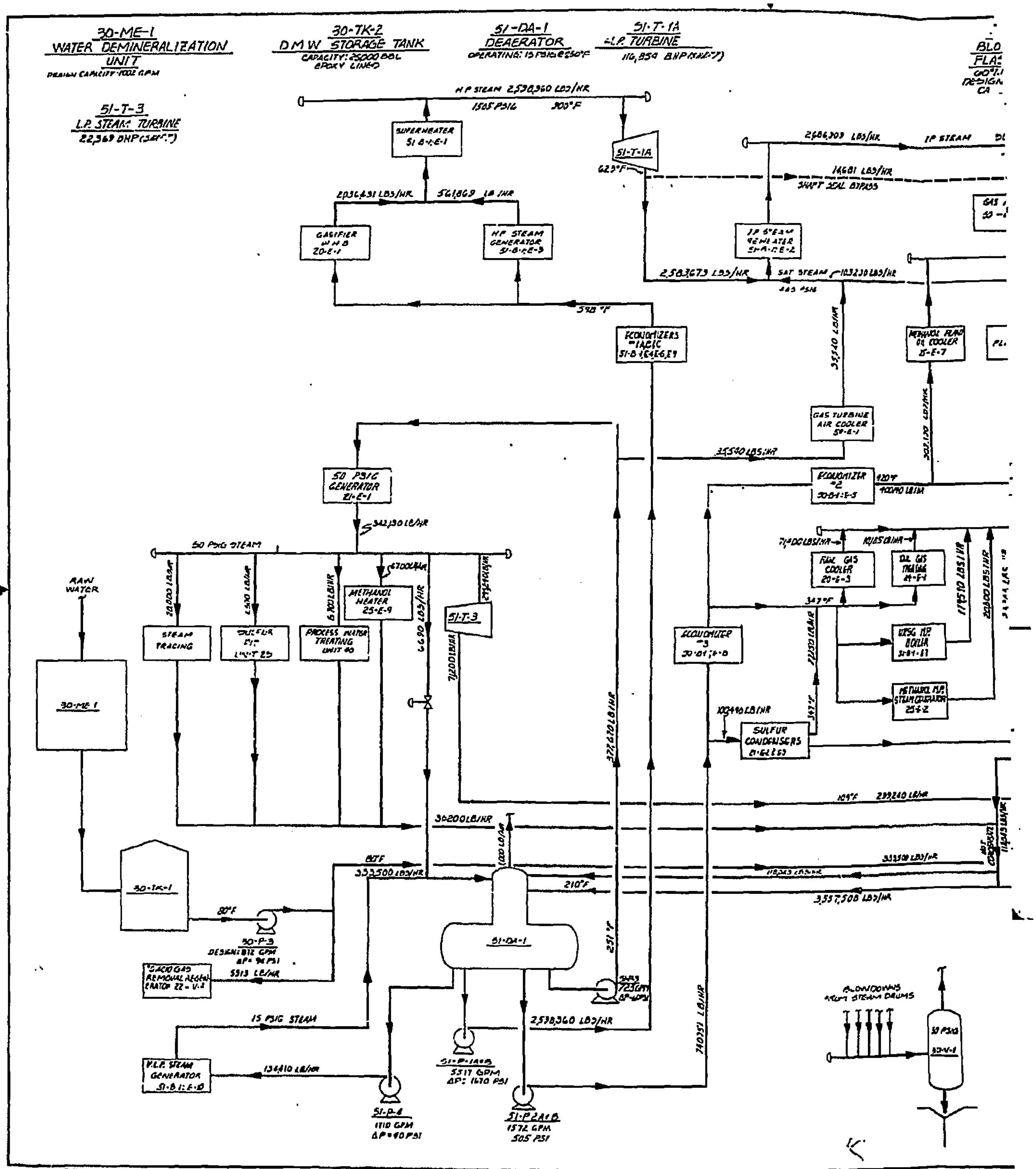
15 psig steam is supplied by steam generation in HRSG coil 51-1-B-1:E-10. This very low-pressure steam is used entirely in deaerator 51-DA-1.

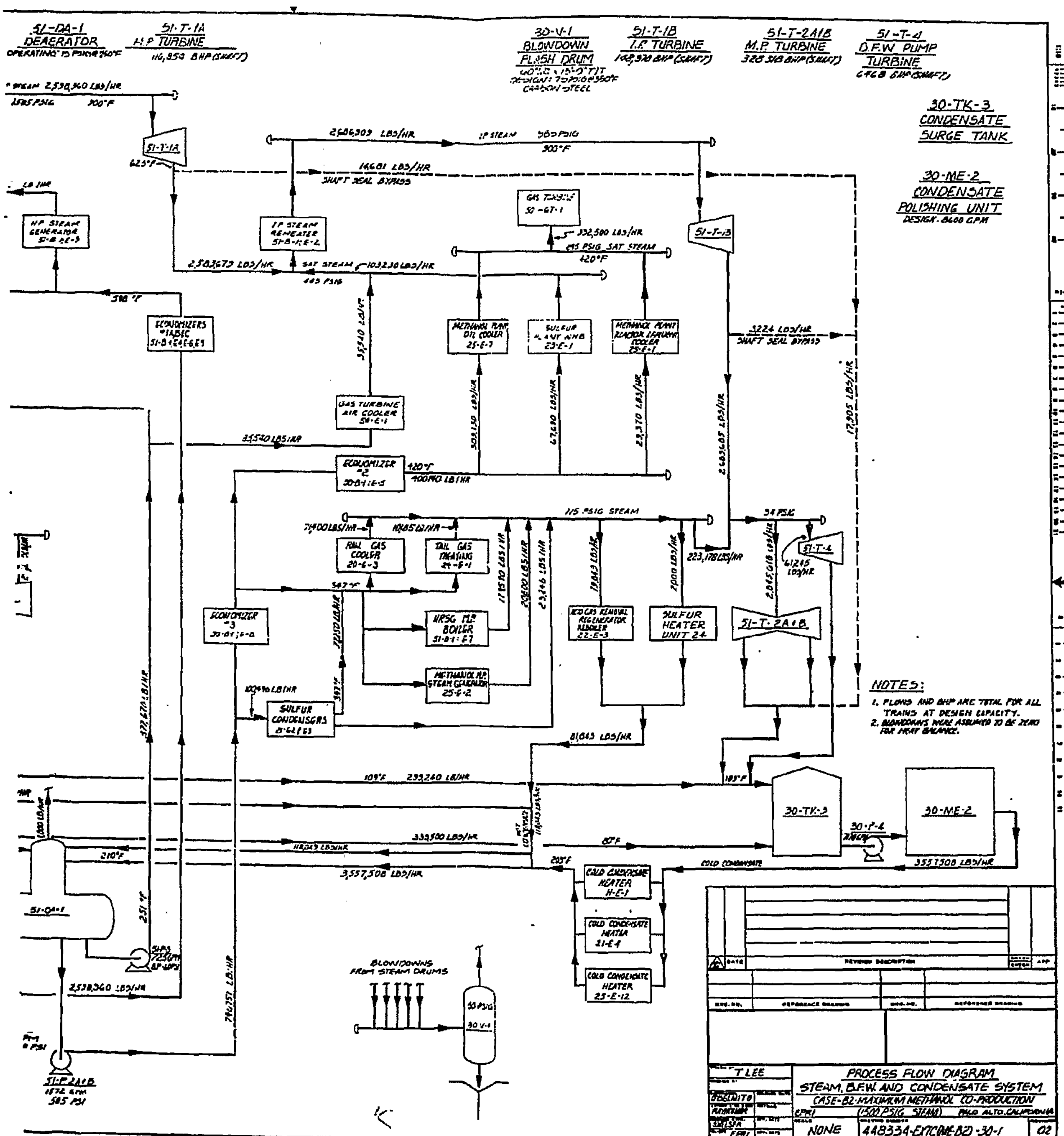
Raw water is treated in an automatic ion exchange demineralizer 30-ME-1, consisting of three strong-acid cation columns, one degasifier (with 10-minute holdup vessel) and three strong-base anion columns. Two of the three cation and anion columns can handle the design flow of raw water either for the two-hour period required for resin regeneration or for the longer time period required for resin changeout. Treated water, suitable for generation of 1500 psig steam is stored in a tank 30-TK-2, which has a 24-hour capacity. Demineralized water is pumped to condensate surge tank 30-TK-3 (30-minute holdup), where it combines with the vacuum condensate from condensers 51-E-11, 51-E-12, and 51-E-14.

The turbine surface condensers 51-E-11, 51-E-12, and 51-E-14 are single-shell single-pass units with divided water boxes. The tubes are 90/10 copper/nickel, 7/8 inch OD, 22 BWG wall thickness. The noncondensable gas removal and priming equipment includes positive displacement rotary vacuum pumps and a recirculating ball-type condenser tube cleaning system. Motor-driven condensate pumps transport the condensate to condensate storage tank 30-TK-3, which is sized for 30-minute capacity at design flow rate.

Condensate polishing unit 30-ME-2 affords further protection to the steam generation units by treating the combined stream of demineralized water and condensate with strong acid and base in four vessels.

The vacuum condensate from polishing unit 30-ME-2 flows to the deaerator after heat recovery from the gasifier effluent in 21-1-E-4, from air compressor intercoolers 11-1-E-1 and from 25-E-12. The hot condensate from the 115 psig and 50 psig steam users also flows to the deaerator. The deaerator providing 10 minute storage is a horizontal tray-type unit operating at 15 psig.


30-ME-1
WATER DEMINERALIZATION UNIT
30-TK-2
DMW STORAGE TANK
51-DA-1
DEAERATOR
51-T-1A
L.P. TURBINE
51-T-3
L.P. STEAM TURBINE
HP STEAM
SUPERHEATER
GASIFIER
HP STEAM GENERATOR
IP STEAM
ECONOMIZERS
50 PSIG GENERATOR
50 PSIG STEAM
STEAM TRACING
PROCESS WATER TREATING UNIT 40
METHANOL HEATER 25-E-9
RAW WATER
30-ME-1
30-TK-1
30-P-3
51-DA-1
GAS TURBINE AIR COOLER
ECONOMIZER #2
ECONOMIZER #3
SULFUR CONDENSERS
15 PSIG STEAM
V.L.P. STEAM GENERATOR
51-P-4
51-P2A&B
BLOWDOWNS FROM STEAM DRUMS


51-DA-1
DEAERATOR
51-T-1A
H.P. TURBINE
30-V-1
BLOWDOWN FLASH DRUM
51-T-1B
I.P. TURBINE
51-T-2A/B
M.P. TURBINE
51-T-4
B.F.W. PUMP TURBINE
30-TK-3
CONDENSATE SURGE TANK
30-ME-2
CONDENSATE POLISHING UNIT
DESIGN - 8600 GPM
IP STEAM
SHAFT SEAL BYPASS
HP STEAM GENERATOR
IP STEAM REHEATER
GAS TURBINE
SAT STEAM
ECONOMIZERS
METHANOL PLANT OIL COOLER 25-E-7
SULFUR PLANT WHB 23-E-1
METHANOL PLANT REACTOR EFFLUENT COOLER 25-E-1
GAS TURBINE AIR COOLER
ECONOMIZER #2
115 PSIG STEAM
RAIL GAS COOLER 20-E-3
TAIL GAS TREATING 24-E-1
ECONOMIZER #3
HRSG M.P. BOILER
METHANOL M.P. STEAM GENERATOR 25-E-2
SULFUR CONDENSERS
ACID GAS REMOVAL REGENERATOR REBOILER 22-E-3
SULFUR HEATER UNIT 24
51-T-2A+B
NOTES:
1. FLOWS AND BHP ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. BLOWDOWNS WERE ASSUMED TO BE ZERO FOR HEAT BALANCE.
30-TK-3
30-ME-2
30-P-4
COLD CONDENSATE
COLD CONDENSATE HEATER H-E-1
COLD CONDENSATE HEATER 21-E-4
COLD CONDENSATE HEATER 25-E-12
BLOWDOWNS FROM STEAM DRUMS
50 PSIG
30 V-1
51-DA-1
51-P-2A+B
585 PSI
T LEE
PROCESS FLOW DIAGRAM
STEAM, B.F.W. AND CONDENSATE SYSTEM
CASE-B2-MAXIMUM METHANOL CO-PRODUCTION
(1500 PSIG STEAM)
EPRI
PALO ALTO, CALIFORNIA
NONE
448334-EI7C(ME-B2)-30-1
02

COMBUSTION GAS TURBINE

Process Flow Diagram EXTC(NE-B2)-50-1 shows one of the five parallel combustion gas turbines for Case B2. No spare turbine is provided.

295 psig saturated steam generated in the process plant is added to the unconverted residual gas from the methanol plant after expansion in fuel gas expander 51-1-EX-1. The combined stream flows to the gas turbine combustor at 245 psig where it is burned with excess air supplied by air compressor 50-1-C-1. Effluent gases exit the combustor at 2000°F and flow to the combustion gas turbine 50-1-GT-1. A small fraction of compressed air is cooled by IP steam generation in 50-1-E-1 before being injected into the turbine to cool the rotors.

The combustion gases are expanded in the combustion gas turbine, producing 519.2 MW net power in generator 50-1-G-1. The effluent gases at 982°F flow to the heat recovery steam generator (HRSG) in Unit 51. The turbine drives the air compressor and electric generator 50-1-G-1. Detailed performance information on the combustion gas turbine is presented in Appendix A.

Equipment Notes

The combustion gas turbine with a combustor outlet temperature of 2000°F is commercially available at the present time. The hot parts of the machine will be fitted with thermal barrier coatings.

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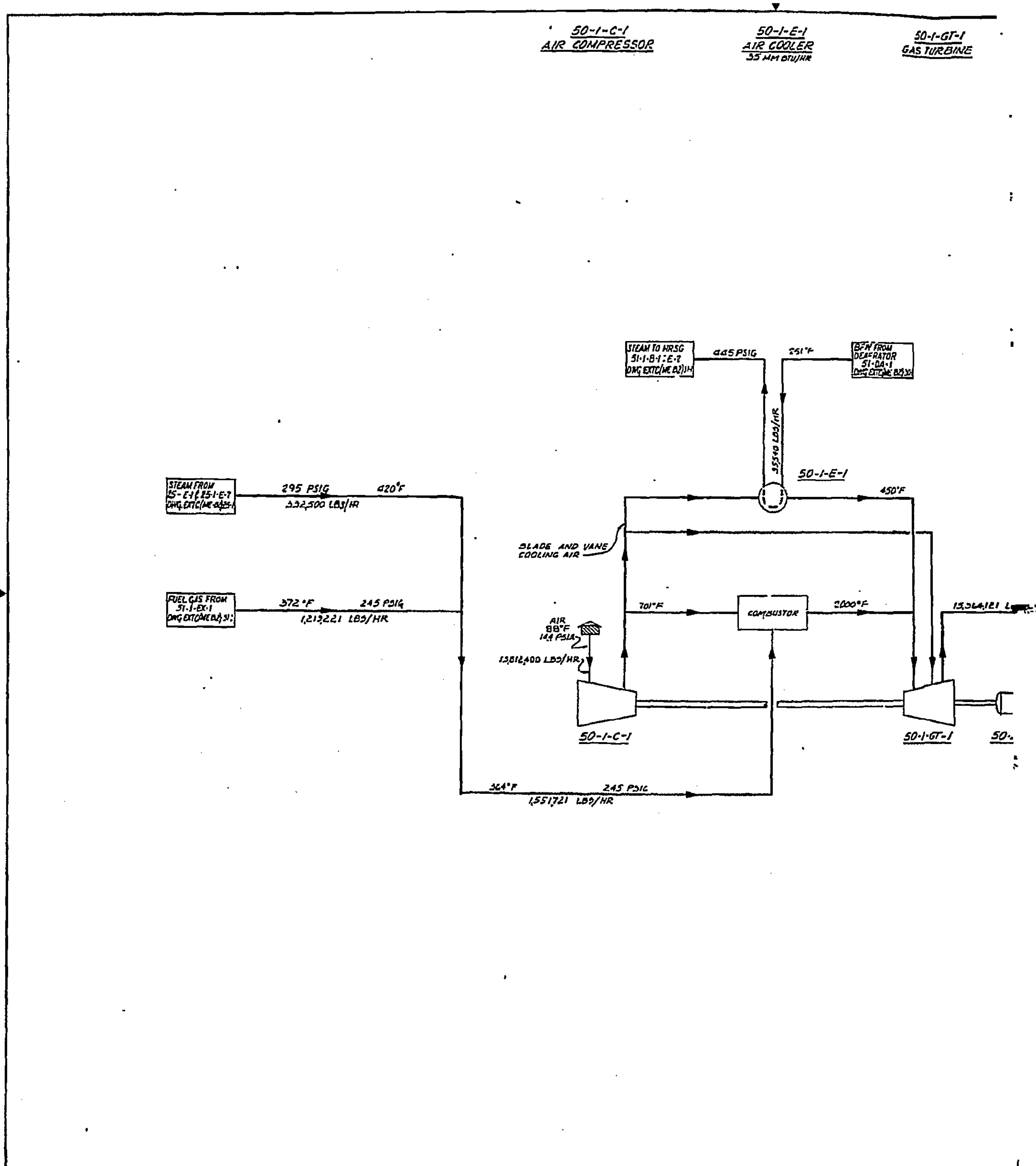


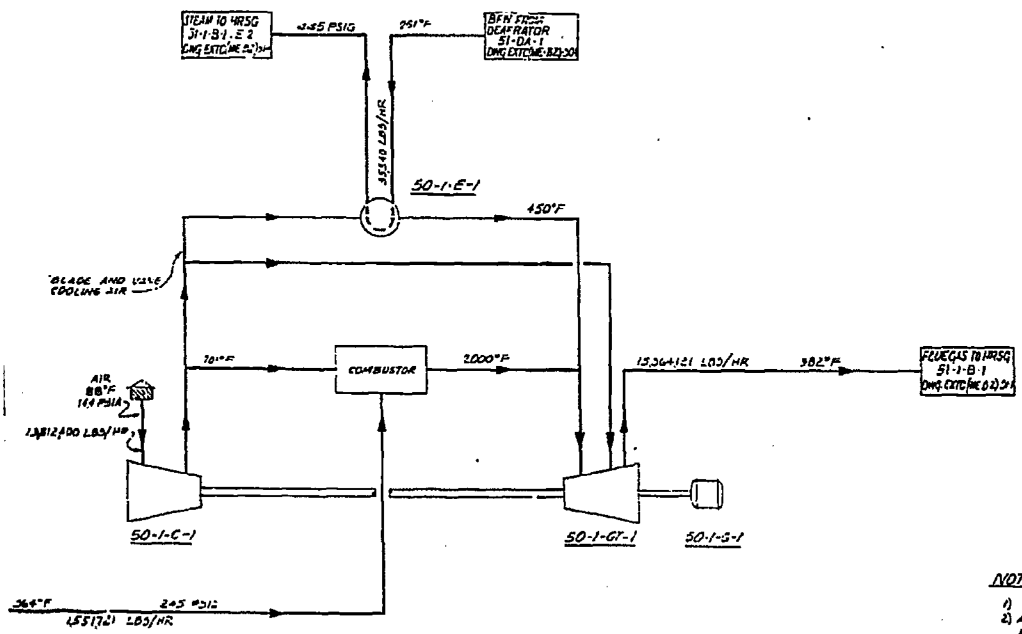
50-1-C-1
AIR COMPRESSOR
50-1-E-1
AIR COOLER
35 MM BTU/HR
50-1-GT-1
GAS TURBINE
STEAM TO HRSG
51-1-B-1 : E-2
445 PSIG
251°F
BFW FROM
DEAERATOR
51-DA-1
35540 LBS/HR
50-1-E-1
STEAM FROM
295 PSIG
420°F
332,500 LBS/HR
450°F
BLADE AND VANE
COOLING AIR
FUEL GAS FROM
51-1-EX-1
372°F
245 PSIG
1,213,221 LBS/HR
701°F
COMBUSTOR
2000°F
AIR
88°F
14.4 PSIA
13,812,400 LBS/HR
50-1-C-1
50-1-GT-1
364°F
245 PSIG
1,551,721 LBS/HR

50-1-C-1
AIR COMPRESSOR

50-1-E-1
AIR COOLER
35 MM BTU/HR

50-1-GT-1
GAS TURBINE

50-1-G-1
GENERATOR
513,130 KW(NET)



NOTE:

1) THIS FLOW DIAGRAM SHOWS ONE TRAIN OF FIVE
2) ALL FLOW RATES, DUTIES AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

DATE	REVISION DESCRIPTION		

DWG. NO.	REFERENCE DRAWING	DWG. NO.	REFERENCE DRAWING

T.LEE
D'DELMITO
RAVIKUMAR
SHULZER
EPRI

PROCESS FLOW DIAGRAM
COMBUSTION GAS TURBINE
CASE B2-MAXIMUM METHANOL CO-PRODUCTION
(1500 PSIG STEAM)

EPRI PALO ALTO, CALIFORNIA

NONE

448334-EXTC(ME-B2)-50-1

02

HEAT RECOVERY STEAM GENERATOR AND STEAM TURBINES

Process Flow Diagram EXTC-(ME-B2)-51-1 shows the heat recovery steam generators (HRSG) 51-1-B-1 and the steam turbines for Case B2. There are ten operating HRSG units, one primary steam turbine 51-T-1A, 1B and 2, and one secondary steam turbine 51-T-3. The primary and secondary steam turbines drive generators for production of electric power. Steam turbine 51-T-4 is used to drive the high-pressure boiler feedwater pump. Additional electric power is generated by the expansion of high-pressure fuel gas in five fuel gas expanders. No spare turbines or HRSGs are provided.

Two HRSGs are coupled with each gas turbine to recover heat from the turbine exhaust gas, which leaves the turbines at 982°F. Flash gas from the methanol unit and a part of Selexol flash gas are burned in the HRSG duct burner. The combustion products are combined with turbine exhaust gas to give a flue gas at 986°F. Radiation heat losses occur throughout the HRSG and are to be realized immediately following the duct burner, whereby the HRSG flue gas inlet temperature is 979°F. The HRSG performs superheating, high-pressure (HP), medium-pressure (MP), and very low-pressure (VLP) steam generation, and boiler feedwater heating. The arrangement of the heat recovery sections of the HRSG in the direction of flue gas flow is as follows:

Superheater	51-1-B-1:E-1	and Reheater 51-1-B-1:E-2
HP Evaporator	51-1-B-1:E-3	
Economizer 1A	51-1-B-1:E-4	
Economizer 2	51-1-B-1:E-5	
Economizer 1B	51-1-B-1:E-6	
MP Evaporator	51-1-B-1:E-7	
Economizer 3	51-1-B-1:E-8	
Economizer 4	51-1-B-1:E-9	
VLP Evaporator	51-1-B-1:E-10	

Saturated HP steam from 20-1-E-1 and saturated HP steam from the HP evaporator is superheated to 900°F in the HRSG superheater 51-1-B-1:E-1. The HRSG superheater outlet supplies the HP feed of back-pressure steam turbine 51-T-1A. Expanded steam from 51-T-1A combines with process generated saturated IP steam and is reheated to 900°F in 51-1-B-1:E-2. This steam supplies the feed to IP back-pressure turbine 51-T-1B. Saturated MP steam generated in process areas combines with the IP turbine exhaust to drive both the MP power turbine 51-T-2 and the

HP BFW pump turbine 51-T-4. These are condensing turbines exhausting at 2-1/2 inches Hg absolute.

HP BFW from the deaerator 51-DA-1 is pumped through high-pressure boiler feedwater pump 51-P-1; preheated to 347°F in economizer 4; heated to 410°F in economizer 1B; and further heated to saturation temperature 598°F in economizer 1A. Both HP steam generator 51-1-B-1:E-3 and the gasifier HP steam generator are supplied by this 598°F boiler feedwater. The operating HP BFW pump is driven by steam turbine 51-T-4 and the spare is motor driven.

Both IP BFW and MP BFW needs are met by boiler feedwater pump 51-P-2. A portion of the pump discharge stream supplies IP steam generators in the process areas. The balance is "let down" to supply MP process area steam generators MP BFW heating to 347°F in economizer 3, while heating to 420°F is done in economizer 2. The HP methanol plant residual gas heated to 606°F in 25-E-4, is expanded to 245 psig in fuel gas expander 51-1-EX-1 generating electric power. The expanded gas at 372°F is routed to gas turbine 51-1-GT-1.

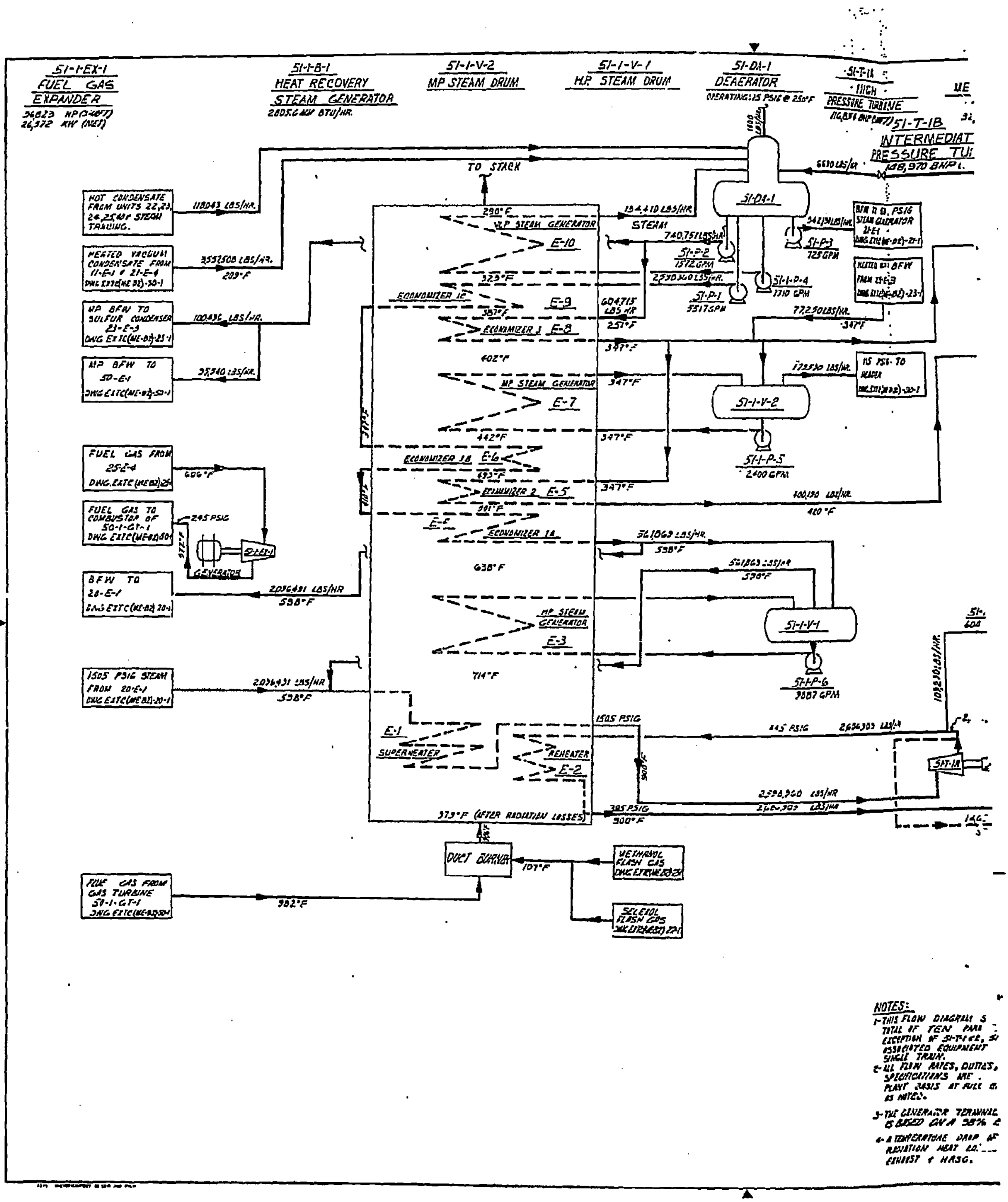
LP BFW is supplied to process area LP steam generators by 51-P-3. LP steam supplies process heating, deaerator heating and LP steam turbine 51-T-3.

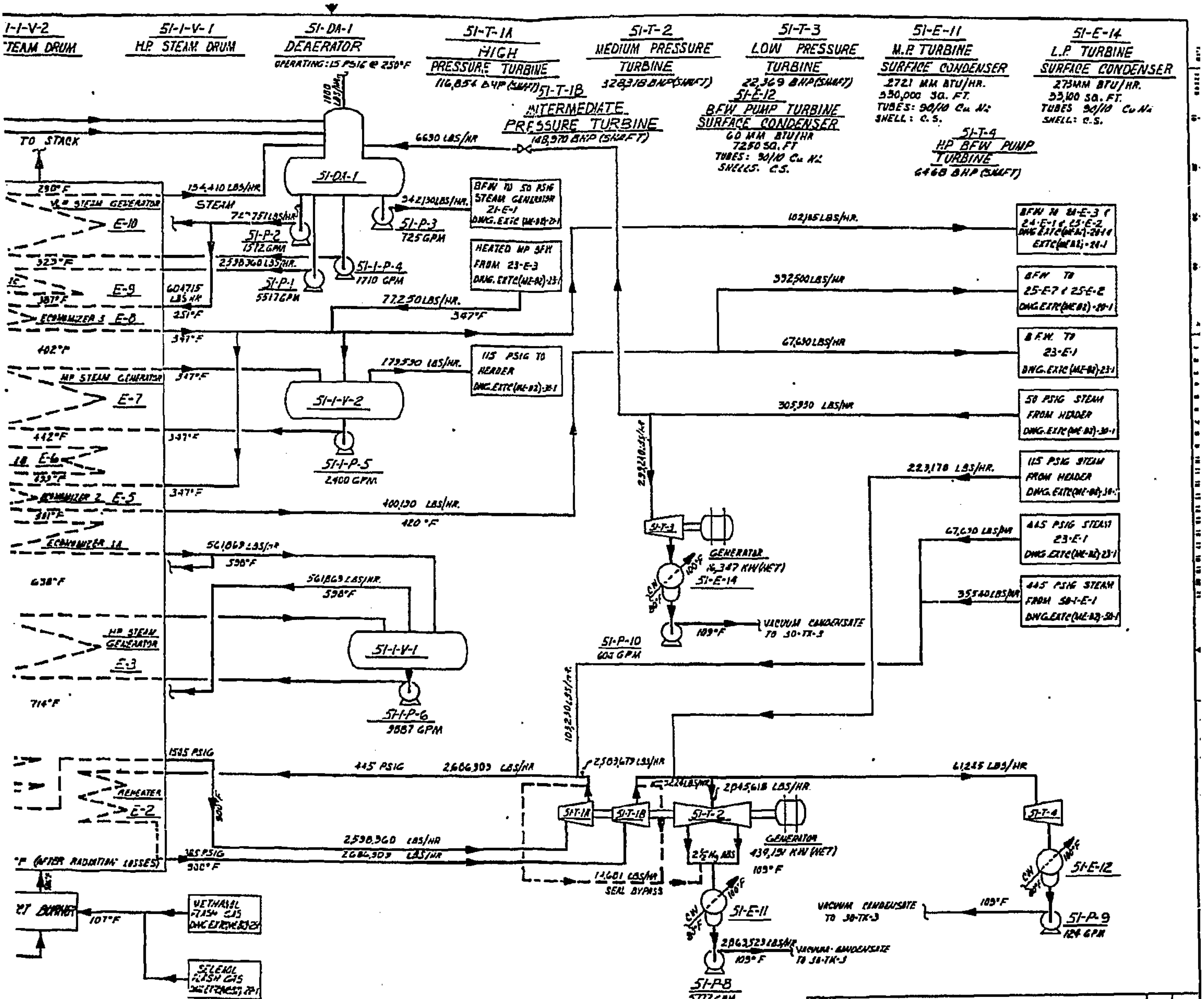
The secondary steam turbine 51-T-3 uses excess saturated LP steam from the process plant to generate a small quantity of additional electric power. This turbine is a condensing type with exhaust conditions of 2-1/2 inches Hg absolute.

Additional deaerating steam is supplied to 51-DA-1 by VLP evaporator 51-1-B-1:E-10, which is fed by VLP BFW circulation pump 51-1-P-4.

The HP and MP evaporator are supplied with steam drums 51-1-V-1 and 51-1-V-2, respectively, and BFW circulation pumps 51-1-P-6 and 51-1-P-5. BFW is pumped through the evaporator at feed to steam mass ratio of 6:1.

The HRSG "pinch-point" temperature used in designing the evaporation and economizing coils has been set at 40°F, in an effort to optimize the trade-off between initial cost and plant efficiency. The stack gas outlet temperature is 290°F, allowing the gas side surface of VLP evaporator 51-1-B-1:E-10 to operate a safe margin above the dew point of the SO_2-bearing stack gas.


51-1-EX-1
FUEL GAS
EXPANDER
51-1-B-1
HEAT RECOVERY
STEAM GENERATOR
51-1-V-2
MP STEAM DRUM
51-1-V-1
HP STEAM DRUM
51-DA-1
DEAERATOR
51-T-1B
INTERMEDIATE
PRESSURE TU
TO STACK
HOT CONDENSATE FROM UNITS 22,23, 24,25 & STEAM TRACING.
HEATED VACUUM CONDENSATE FROM 11-E-1 & 21-E-4
HP BFW TO SULFUR CONDENSER 23-E-3
MP BFW TO 50-E-1
FUEL GAS FROM 25-E-4
FUEL GAS TO COMBUSTOR OF 50-1-GT-1
BFW TO 28-E-1
1505 PSIG STEAM FROM 20-E-1
FLUE GAS FROM GAS TURBINE 51-1-GT-1
DUCT BURNER
METHANOL FLASH GAS
SELEXOL FLASH GAS
GENERATOR
LP STEAM GENERATOR
E-10
ECONOMIZER 1C
E-9
ECONOMIZER 3
E-8
MP STEAM GENERATOR
E-7
ECONOMIZER 1B
E-6
ECONOMIZER 2
E-5
E-4
ECONOMIZER 1A
HP STEAM GENERATOR
E-3
E-1
SUPERHEATER
REHEATER
E-2
51-DA-1
51-P-2
51-P-3
51-1-P-4
51-P-1
51-1-V-2
51-1-P-5
51-1-V-1
51-1-P-6
51-T-1A
NOTES:
1-THIS FLOW DIAGRAM
2-ALL FLOW RATES, DUTIES,
3-THE GENERATOR TERMINAL
4-A TEMPERATURE DROP OF



NOTES:

1- THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF TEN PARALLEL TRAINS WITH THE EXCEPTION OF 51-T-1 & 2, 51-1-T-3 AND 51-DA-1 AND ASSOCIATED EQUIPMENT WHICH CONSIST OF A SINGLE TRAIN.

2- ALL FLOW RATES, DUTIES, AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

3- THE GENERATOR TERMINAL NET POWER GENERATION IS BASED ON A 98% EFFICIENCY.

4- A TEMPERATURE DROP OF ABOUT 7°F ALLOWS FOR RADIATION HEAT LOSSES IN THE GAS TURBINE EXHAUST & HRSG.

DATE	REVISION DESCRIPTION	CHECK	APP

DWG NO.	REFERENCE DRAWING	DWG NO.	REFERENCE DRAWING

V. FRIAS
G. BELMITO
RAMAMMER
SARLSAR
EPRI

PROCESS FLOW DIAGRAM
HEAT RECOVERY & STEAM GENERATION
CASE B2-MAXIMUM METHANOL CO-PRODUCTION (1500 PSIG STM)
EPRI PALO ALTO CALIF.

SCALE: NONE
448334 - EXT.C. (ME-B2) -51-1
03

Section 5

PROCESS DISCUSSION

Two grass roots plants (one power generation base case and one methanol/electric power coproduction case) based on oxygen-blown Texaco gasifiers integrated with current state-of-the-art combined-cycle generating equipment are shown schematically on Block Flow Diagrams EXTC(ME-A2)-1-1 and EXTC(ME-B2)-1-1. These plants consume 10,000 short tons per day of Illinois No. 6 coal, fed to the gasifiers in a water slurry containing 66.5 weight percent solids. Case A2 generates electric power only and provides the base cost for producing electric power, which is used as a credit in methanol/electric power coproduction Case B2.

Since each case uses the same coal and oxidant feed rate, the coal receiving and conveying, coal grinding and slurrying, oxidant feed, and coal gasification units are the same for both cases. The major differences in the processing schemes occur in the acid gas removal units, the heat recovery steam generators (HRSGs), and in the methanol synthesis.

PERFORMANCE COMPARISONS

Table 5-1 provides overall system performance comparisons for the conventional GCC design (Case A2) and the plant coproducing methanol and electric power (Case B2). As each of these plants was designed to process 10,000 tons/day of coal, comparisons are difficult to make as the two plants have different electric power production capacities. Therefore, in order to understand system differences, the Case B2 design (methanol coproduction case) was scaled linearly to produce the same net power as the Case A2 design. The scaled results for Case B2 are shown in the third column of Table 5-1.

The Case A2 results indicate that 1,106.52 MW would require a coal feed rate of 798,333 lbs/hr (MF) of coal. The scaled Case B2 results demonstrate that an additional coal feed rate of 291,791 lbs/hr (MF) would result in the production of 3,118 tons/day of methanol (or 10,520 FOE barrels/day of methanol) if the "once-through" concept used in this study is employed.

Comparing the power generations and consumptions for the scaled Case B2 (methanol coproduction) with the Case A2 (conventional GCC) system surfaces some interesting characteristics to be associated with the inclusion of a "once through" methanol plant in a GCC system. First, it can be seen that the gas turbines produce 16.7 MW more power in the coproduction design, even though the fuel gas for this case has a lower heating value than that in the GCC case. The two fundamental reasons for this slightly increased gas turbine power output can be explained with reference to the following table:

	Conventional Texaco Based GCC Power Plant with No Methanol Production	Methanol Coproduction Case (Same Design as Case B2), Scaled to Produce Same Quantity of Electricity as Case A2
Case Designation	A2	B2 Scaled Up
HHV of fuel to turbine, BTU/SCF	289.8	284.0
LHV of fuel to turbine, BTU/SCF	271.6	270.5
Fuel chemical heat to turbine (HHV basis), 10^6 BTU/hr	7,864.3	7,636.6
Fuel chemical heat to turbine (LHV basis), 10^6 BTU/hr	7,368.9	7,273.8
Fuel $CO:H_2$ ratio, moles/mole	1.468	2.427
Steam injection rate, lbs/hr	0	454,029
Percentage of turbine gross work required to power the air compressors	56.6	54.0

This table shows that although the fuel chemical heat flowing to the gas turbine on an HHV basis has been depleted by 3 percent due to the methanol synthesis, the chemical heats on a LHV basis are almost identical. This is due to the fact that more of the hydrogen in the raw fuel gas than carbon monoxide has been removed during the methanol synthesis step. Therefore, the resulting fuel gas after methanol synthesis is richer in carbon monoxide and is therefore a "better" fuel for power generation. Also, the coproduction Case (B2) employs steam injection whereas A2 does not. This steam replaces a small part of the turbine's cooling air requirement, thereby reducing the power required by the air compressors from almost 57 percent of the gross turbine output to 54 percent of gross turbine work. This reduction in air compressor power requirement in the coproduction case increases the electricity available at the generator terminals.

Second, Table 5-1 indicates that the steam turbine power generated in the scaled-up coproduction Case (B2) is 53.78 MW greater than the equivalent steam turbine power in the GCC Case (A2). This increase is primarily due to the additional steam raised by cooling the extra high temperature raw gas produced from the incremental 291,791 lbs/hr (MF) of coal being gasified. The incremental high pressure steam generation in the coproduction case amounts to 639,654 lbs/hr.

Finally, Table 5-1 shows that the scaled coproduction case consumes an extra 70.48 MW to satisfy internal power requirements over the power required by the GCC plant. The 70.48 MW is dominated by the additional power requirement of the oxidant production and feed system (52.88 MW). It is of interest to note that a requirement of 70.48 MW to produce 3,118 tons/day of methanol is equivalent to the consumption of 542.5 kWh/ton. A dedicated coal-to-methanol plant requires between 700 kWh/ton and 900 kWh/ton to satisfy internal power requirements.

GASIFIER MATERIAL BALANCE

A gasifier material balance for full capacity operation is given in Table 5-2.

The coal quantity and composition, and oxidant composition were selected by EPRI. The yields are typical of a Texaco single-stage entrained oxygen-blown gasifier operating at 600 psig. (1) As noted in another EPRI report (2), pressure has little effect on gasifier yields. Therefore, the yields at 600 psig were deemed to be acceptable for use with the 1,000 psig gasifier used in this study.

The coal feed is 10,000 tons per day of Illinois No. 6, fed in a 66.5 weight percent slurry. For this particular coal, slurry concentrations in excess of 60 percent solids have been achieved. The first plants built will probably employ lower slurry concentrations. It is important to bear in mind, however, 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.

Table 5-1

SYSTEM PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

CASE DESCRIPTION	Conventional Texaco Based GCC Power Plant With No Methanol Production	Texaco GCC Plant With Once-Through Methanol Coproduction (As Designed)	Methanol Coproduction Case (Same Design As Case B2), Scaled To Produce Same Quantity Of Electricity As Case A2
CASE DESIGNATION	A2	B2	B2-Scaled-Up
GASIFICATION SYSTEM			
Coal Feed Rate, lbs/hr m.f.	798,333	798,333	1,090,124
Oxygen/Coal Feed*, lb/lb m.f.	0.858	0.858	0.858
Oxidant Temperature, °F	287	287	287
Slurry Water/Coal Ratio, lb/lb m.f.	0.503	0.503	0.503
Gasification Section Average Pressure, psig	1,000	1,000	1,000
Crude Gas Temperature, °F	2,300-2,600	2,300-2,600	2,-300-2,600
Crude Gas HHV (Dry Basis), BTU/SCF**	281.1	281.1	281.1
POWER SYSTEM			
Temperature of Fuel Gas to Gas Turbine	339	372	372
Gas Turbine Inlet Temperature, °F	2,000	2,000	2,000
Steam Conditions, psig/°F/°F	1450/900/900	1450/900/900	1450/900/900
Surface Condenser Pressure, inches Hg abs.	2.5	2.5	2.5
Stack Gas Exit Temperature, °F	290	290	290
Gas Turbine Power#, MW	692.25	519.19	708.95
Steam Turbine Power#, MW	561.43	450.54	615.21
Fuel Gas Expander Power#, MW	36.64	26.37	36.01
Oxygen Plant Power##, MW	1.72	1.72	2.35
Power Consumed, MW	185.52	187.48	256.00
Net System Power, MW	1,106.52	810.34	1,106.52
METHANOL			
Methanol Produced, 10^3 gal/day	--	688.0	939.5
tons/day	--	2,283.4	3,118.0
FOEBδ/day	--	7,705	10,520
10^6 BTU/day	--	45,072	61,546
OVERALL SYSTEM			
Process and Dearator Makeup Water, gpm	602	1,245††	1,700††
gpm/1000 MW	544	1,536	2,097
Cooling Tower Makeup Water†, gpm	8,299	6,875	9,388
gpm/1000 MW	7,495	8,478	11,577
Cooling Tower Heat Rejection, % of coal HHV	42.9	35.5	35.5
Air Cooler Heat Rejection, % of coal HHV	0.30	1.28	1.28
Net System Heat Rate, BTU/kWh	9,214	--	--
Overall System Efficiency (Coal → Power and Methanol), % of coal HHV	37.0	45.5	45.5
Efficiency of Methanol Production, % of coal HHV	--	68.8%	68.8%

* Dry basis, 100 percent oxygen
\# At generator terminals
† Includes condensate from oxidant feed in Unit 11
δ FOEB = Barrels of distillate fuel oil (5.85×10^6 BTU/BBL) with high heating value equivalent to methanol produced
** Excluding HHV of H_2S, COS, NH_3
\#\# From power recovery expander in 11-ME-1
†† Includes water for steam injection into gas turbine

TABLE 5-2

GASIFIER MATERIAL BALANCE - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Feeds

	T (°F)	lb/hr	lb mol/hr
Coal	140		
Moisture		35,000	1,942.8
Ash		80,000	-
MAF Coal			
Carbon as C		554,985	46,205.9
Hydrogen as H_2		42,525	21,094.6
Oxygen as O_2		80,022	2,500.8
Nitrogen as N_2		9,985	356.4
Sulfur as S		30,816	961.1
TOTAL COAL		833,333	
Oxidant	290		
Oxygen as O_2		684,687	21,397.3
Argon as Ar		4,326	108.3
Nitrogen as N_2		9,241	329.9
TOTAL OXIDANT		698,254	21,835.5
Slurry Water	140	366,553	20,364.1
TOTAL FEEDS		1,898,140	

Effluents

	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	2300-2600			
CH_4		1,158	72.2	0.08
H_2		52,364	25,974.2	28.84
CO		1,071,001	38,236.4	42.45
CO_2		345,232	7,844.4	8.71
H_2S		30,907	906.9	1.01
COS		3,256	54.2	0.06
N_2		16,725	597.1	0.66
Ar		4,326	108.3	0.12
H_2O		290,137	16,106.4	17.88
NH_3		3,034	178.1	0.19
TOTAL GASIFIER EFFLUENT		1,818,140	90,078.2	100.00
Ash				
Carbon		Nil		
Ash		80,000		
TOTAL ASH		80,000		
TOTAL EFFLUENTS		1,898,140		

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 coal 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 via the coal slurry. At gasification temperature, the gasifiers are assumed capable of destroying recycle ammonia. The presence of ammonia in the process condensate has the potentially beneficial effect of neutralizing dissolved carbon dioxide.

PROCESS ENERGY BALANCES

Tables 5-3 and 5-4 contain the overall process energy balances at 100 percent capacity operation for the two cases. The boundary for each balance encompasses the entire plant, exclusive of the cooling tower but including the power demand of the pumps and fans. Energy contents of streams crossing the boundary are expressed as the sum of the streams'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. All of the energy balances close within one-quarter of one percent. Discrepancies result from approximations applied in calculating enthalpies for some process streams.

Energy balance comparisons are presented in Table 5-5 and these are derived from the two previous tables. Coal charged at 10,000 ton/day is equivalent to 10,201 x 10^6 Btu/hr HHV. The significant difference between the two cases lie in the power production, methanol product, and steam turbine condenser categories.

POWER CONSUMPTION SUMMARY

Table 5-6 presents power consumption for the major plant sections under conditions of normal operation at 100 percent capacity factor.

In Unit 10, the coal pulverizing equipment (10-ME-4) requires most of the power. Since the power for pulverization is very sensitive to both the equipment used and the coal properties, the power for coal handling is preliminary pending grinding tests.

Unit 11 power requirements are due to the air and oxygen compressor drivers (11-C-1-M and 11-C-2-M) which are synchronous, and Type II induction motors, respectively. All intercoolers are water-cooled and the power of the cooling tower pumps is included elsewhere.

Table 5-3

ENERGY BALANCE - CASE A2
NO METHANOL, 1500 PSIG SATURATED SYSTEM

Basis: 60°F, water as liquid, 3413 Btu/kWh

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
HEAT IN						
Coal	10,196	5				10,201
Air to Oxidant Feed System		22	52			74
Air to Sulfur Recovery			1			1
Air to Combustion Gas Turbines		140	335			475
Demineralized and Raw Water		1				1
Auxiliary Power Inputs					627	627
TOTAL	10,196	168	388	0	627	11,379
HEAT OUT						
Oxidant Compressors Inter/Aftercooling		267				267
Condensate From Air Separation Plant		1				1
Air Separation Plant Vent Gas		18	22			40
Quench Process Water Cooling, 20-E-5		82				82
Gasifier Heat Losses				82		82
Ash Cake		6				6
Process Wastewater		3				3
Selexol Refrigeration Cooler		58				58
Selexol Overhead Condenser		8				8
Tail Gas Treating Unit Cooling		23				23
Sulfur By-Product (27,770 lb/hr)	111	2				113
Tail Gas Unit Vent Gas	7	1	6			14
Steam Heat Losses*			30			30
Gas Turbine Power					2,363	2,363
Expander					125	125
Steam Turbine Power					1,916	1,916
HRSG Flue Gas		1,248	836			2,084
Steam Turbine, Gas Turbine and HRSG Loss**				221		221
Steam Turbine Condensers		3,828				3,828
Motor and Mechanical Losses					99	99
TOTAL	118	5,545	894	303	4,503	11,363

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.14\%$

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

** Includes mechanical losses from steam turbine, steam turbogenerator, and gas turbine generator; radiation losses from gas turbine and HRSG

TABLE 5-4

ENERGY BALANCE - CASE B2
METHANOL COPRODUCTION, 1500 PSIG SATURATED STEAM, LOW-CONVERSION METHANOL PLANT

Basis: 60°F, water as liquid, 3413 Btu/kWh

	10^6 Btu/hr					
	HHV	Sensible	Latent	Radiation	Power	Total
HEAT IN						
Coal	10,196	5				10,201
Air to Oxidant Feed System		22	52			74
Air to Sulfur Recovery			1			1
Air to Combustion Gas Turbines		94	225			319
Demineralized and Raw Water		6				6
Auxiliary Power Inputs					634	634
TOTAL	10,196	127	278	0	634	11,235
HEAT OUT						
Oxidant Compressors Inter/Aftercooling		267				267
Condensate From Air Separation Plant		1				1
Air Separation Plant Vent Gas		18	22			40
Quench Process Water Cooling, 20-E-5		82				82
Gasifier Heat Losses				82		82
Ash Cake		6				6
Process Wastewater		3				3
Methanol Product	1,878	5				1,883
Methanol Plant		107				107
Selexol Refrigeration Cooler		81				81
Selexol Overhead Condenser		28				28
Tail Gas Treating Unit Cooling		28				28
Sulfur By-Product (30,799 lb/hr)	123	2				125
Tail Gas Unit Vent Gas	6	1	8			15
Steam Heat Losses*			30			30
Gas Turbine Power					1,772	1,772
Steam Turbine Power					1,538	1,538
Expander, Fuel Gas					90	90
HRSG Flue Gas		883	850			1,733
Steam Turbine, Gas Turbine and HRSG Losses**				153		153
Steam Turbine Condensers		3,054				3,054
Motor and Mechanical Losses					96	96
TOTAL	2,007	4,566	910	235	3,496	11,214

$\frac{\text{In-Out}}{\text{In}} \times 100\% = 0.19\%$

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

** Includes mechanical losses from steam turbine, steam turbogenerator, and gas turbine generator; radiation losses from gas turbine and HRSG

TABLE 5-5

ENERGY BALANCE COMPARISON - OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes	Yes Case B2 Scaled To Produce Same Quantity of Electricity as Case A2
		10^6 Btu/hr	
	A2	B2	B2 - Scaled-Up
Case Designation			
HEAT IN			
Coal	10,201	10,261	13,929
Air to Sulfur Recovery	74	74	101
Air to Sulfur Recovery	1	1	1
Air to Combustion Gas Turbines	475	319	436
Demineralized and Raw Water	1	6	8
Auxiliary Power Inputs	627	634	866
TOTAL	11,379	11,235	15,341
HEAT OUT			
Oxidant Compressors Inter/Aftercooling	267	267	365
Condensate From Air Separation Plant	1	1	1
Air Separation Plant Vent Gas	40	40	55
Process Water Cooling	82	82	112
Gasifier Heat Losses	82	82	112
Ash Cake	6	6	8
Process Wastewater	3	3	4
Selexol Refrigeration Cooler	58	81	111
Selexol Overhead Condenser	8	26	38
Tail Gas Treating Unit Cooling	23	28	38
Sulfur By-Product	113	125	171
Tail Gas Unit Vent Gas	14	15	21
Steam Heat Losses*	30	30	41
Gas Turbine Power	2,363	1,772	2,420
Steam Turbine Power	1,916	1,538	2,100
Fuel Gas Expander	125	90	123
HRSG Flue Gas	2,084	1,733	2,366
Steam Turbine, Gas Turbine and HRSG Losses**	221	153	209
Steam Turbine Condensers	3,828	3,054	4,170
Motor and Mechanical Losses	99	96	131
Methanol Plant	0	107	146
Methanol Product	0	1,883	2,571
TOTAL	11,363	11,214	15,313
Heat In - Heat Out	16	21	28
$\frac{\text{In-Out}}{\text{In}} \times 100\%$	0.14%	0.19%	0.19%

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

** Includes mechanical losses from steam turbine, steam turbogenerator, and gas turbine generator; radiation losses from gas turbine and HRSG

TABLE 5-6

POWER CONSUMPTION SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS
(Expressed as Kilowatts at 100 percent capacity Factor)

Methanol Coproduction		None	Yes	Yes Case B2 Scaled To Produce Same Quantity of Electricity as Case A2
Case Designation		A2	B2	B2-Scaled-Up
Unit No.	Description			
10	Coal Handling*	8,840	8,840	12,071
11	Oxidant Feed	144,687	144,687	197,570
20	Gasification	1,928	1,928	2,633
21	Gas Cooling	55	55	75
22	Acid Gas Removal	4,185	5,769	7,878
23	Sulfur Recovery	549	638	871
24	Tail Gas Treating	3,170	5,054	6,901
25	Methanol Unit	-	1,697	2,317
30	Raw Water Treating	752	852	1,163
32-45	General Facilities**	19,680	16,700	22,804
50	Gas Turbine	-	-	-
51	Steam System	1,632	1,220	1,666
51	Surface Condenser#	36	36	49
TOTAL PLANT POWER		185,514	187,476	255,998

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

** About 90 percent of this power demand is attributable to the cooling water system pumps and fans

For mechanical vacuum pump

The recycle gas compressors (20-C-1; one per train) requires about 80 percent of the power used in Unit 20. One recycle gas compressor is employed for each train in this report. Miscellaneous pumps use most of the remaining 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 the most power. However, 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.

In Unit 25, the recycle oil pumps consume the most power.

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

In the General Facilities section, the cooling water circulation pumps and the cooling tower fans required the most power. The entire cooling water system is responsible for approximately 90 percent of the total power demand of the general facilities (Units 32 through 45).

Pumps consume the bulk of the power supplied to Units 50/51 with the HP BFW circulation pumps (51-P-1) being the largest consumers.

REFERENCES

1. "Economic Studies of Coal Gasification Combined Cycle Systems for Electric Power Generation," EPRI AF-642, January 1978.

2. "Effects of Sulfur Emission Controls on the Cost of Gasification Combined Cycle Power Systems," EPRI AF-916, October 1978.

Section 6

CAPITAL AND OPERATING COST ESTIMATES

PLANT FACILITIES INVESTMENT

Plant investment (in $1000; mid-1978) for all sections of the plant for both cases, plus a reference case from an earlier EPRI report, are presented in Table 6-1. Tables 6-2 and 6-3 give the breakdown of plant facilities investment for the major units of each case. The estimates in Table 6-1 through 6-3 contain allowances for both project and process contingencies. The project contingency allowance is intended to cover additional 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 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 process contingency allowances used for each major subsection of both plant designs are shown in Table 6-4. These allowances were supplied by EPRI.

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

COMPARISON OF CASES A2 AND EXTC-79 (TEXACO GCC PLANT DESIGNS)

The Case A2 design is comparable to the design of the reference case, EXTC-79[1], with the exception of the gasifier pressure and the gas cooler configuration. Operation of the gasifier in Case A2 at 1000 psig requires higher discharge pressure for both coal slurry pumps and the oxidant feed compressor relative to the reference case with a gasifier operating pressure of 600 psig.

[1] Published in EPRI Report No. AP-1624, November, 1980.

Table 6-1

SUMMARY OF PLANT FACILITIES INVESTMENT ESTIMATES IN $1000*
METHANOL COPRODUCTION OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes	None
Raw Gas Cooling	Saturated Steam	Saturated Steam	Saturated Steam
Case Designation	A2	B2	EXTC-79**
PLANT FACILITIES INVESTMENT			
Coal Handling	37,556	37,556	33,213
Oxidant Feed	176,354	176,354	176,404
Gasification and Ash Handling	32,561	32,561	18,420
Gas Cooling	91,292	82,157	104,198
Particulate Removal	9,454	9,454	
COS Hydrolysis	-	3,812	-
Acid Gas Removal	15,262	18,052	13,721
Sulfur Recovery	6,231	7,876	5,741
Tail Gas Treating	9,635	12,169	10,369
ZnO Treating	-	9,358	-
Methanol Synthesis	-	37,364	-
Fuel Storage	5,217	-	-
Steam, Condensate, BFW	4,644	6,502	4,947
Fuel Gas Expansion	19,946	14,263	-
Combined Cycle	377,872	285,022	353,730
General Facilities	69,132	65,810	71,708
Initial Catalyst and Chemicals	3,854	8,343	4,042
Subtotal	859,010	806,653	796,493

*Including process and project contingencies plus Illinois sales tax (Mid-1978 dollars)

**EPRI Report AP-1624 for a Texaco gasification combined-cycle plant with the following features:

- 2000°F gas turbine
- Saturated HP steam generation in the gas cooling section
- 1450 psig/900°F/900°F steam cycle in these turbine casings
- Gasifier operating at 600 psig

Table 6-2

PLANT FACILITIES INVESTMENT - CASE A2
NO METHANOL, 1500 PSIG SATURATED STEAM

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Total Cost $/kW**	Percent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment: $1000*	$/kW**
Coal Handling, Preparation and Feeding	15,029	6,194	10,677	757	32,657	29.51	4.5	0	4,899	37,556	33.94
Oxidant Feed	83,577	30,015	37,926	1,833	153,351	138.58	20.9	0	23,003	176,354	159.37
Gasification and Ash Handling	14,015	3,511	7,203	707	25,436	22.99	3.5	3,310	3,815	32,561	29.43
Gas Cooling	28,802	15,832	25,661	1,502	71,797	64.88	9.8	8,725	10,770	91,292	82.50
Particulate Removal	3,298	1,813	2,938	172	8,221	7.43	1.1	0	1,233	9,454	8.54
Acid Gas Removal	6,420	2,322	4,201	328	13,271	11.99	1.8	0	1,991	15,262	13.79
Sulfur Recovery	2,341	1,102	1,854	121	5,418	4.90	.7	0	813	6,231	5.63
Tail Gas Treating	3,746	1,713	2,747	172	8,378	7.57	1.1	0	1,257	9,635	8.71
Methanol Storage	2,117	613	708	40	3,478	3.14	.5	1,218	521	5,217	4.72
Steam, Condensate, BFW	1,879	765	1,302	92	4,038	3.65	.6	0	606	4,644	4.20
Fuel Gas Expansion	9,050	2,648	5,182	466	17,346	15.68	2.4	0	2,600	19,946	18.03
Combined Cycle	171,442	50,160	98,159	8,823	328,584	296.95	44.9	0	49,288	377,872	341.49
General Facilities	30,139	12,320	16,764	892	60,115	54.34	8.2	0	9,017	69,132	62.48
Initial Catalyst and Chemicals	-	-	-	-	3,854	3.48	0.5	-	-	3,854	3.48
Subtotal	371,855	129,063	215,322	15,905	735,944	665.09	100.0	13,253	109,813	859,010	776.31

PLANT FACILITIES INVESTMENT SUMMARY

	$1000*	$/kW**
Process Plant Investment and General Facilities	735,944	665.09
Process Contingency	13,253	11.98
Project Contingency	109,813	99.24
Total Plant Investment	859,010	776.31

*Mid-1978 dollars

**Based on 100 percent plant design power output of 1,106,522 kW net

#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 6-3

PLANT FACILITIES INVESTMENT - CASE B2
METHANOL COPRODUCTION**, 1500 PSIG SATURATED STEAM

Plant Section	Cost Breakdown Without Contingencies: Direct Field Mat'l#	Direct Field Labor##	Eng. & Support Costs§	Sales Tax	Total Cost $1000*	Per-cent	Contingencies: Process $1000*	Project $1000*	Total Plant Investment $1000*
Coal Handling, Preparation and Feeding	15,029	6,194	10,677	757	32,657	4.8	0	4,899	37,556
Oxidant Feed	83,577	30,015	37,926	1,833	153,351	22.4	0	23,003	176,354
Gasification and Ash Handling	14,015	3,511	7,203	707	25,436	3.7	3,310	3,815	32,561
Gas Cooling	26,331	14,475	23,461	1,373	65,640	9.6	6,671	9,846	82,157
Particulate Removal	3,298	1,813	2,938	172	8,221	1.2	0	1,233	9,454
COS Hydrolysis	1,330	731	1,185	69	3,315	.5	0	497	3,812
Acid Gas Removal	7,601	2,744	4,964	388	15,697	2.3	0	2,355	18,052
Sulfur Recovery	2,961	1,392	2,343	153	6,849	1.0	0	1,027	7,876
Tail Gas Treating	4,733	2,164	3,468	217	10,582	1.5	0	1,587	12,169
ZnO Treating	4,908	884	2,098	247	8,137	1.2	0	1,221	9,358
Methanol Synthesis	15,039	2,888	6,276	706	24,909	3.6	8,719	3,736	37,364
Steam, Condensate, BFW	2,643	1,072	1,812	127	5,654	.8	0	848	6,502
Fuel Gas Expansion	6,471	1,894	3,705	333	12,403	1.8	0	1,860	14,263
Combined Cycle	129,313	37,836	74,041	6,655	247,845	36.1	0	37,177	285,022
General Facilities	28,779	11,642	15,942	863	57,226	8.3	0	8,584	65,810
Initial Catalyst and Chemicals	-	-	-	-	8,343	1.2	-	-	8,343
Subtotal	346,028	119,255	198,039	14,600	686,265	100.0	18,700	101,688	806,653

PLANT FACILITIES INVESTMENT SUMMARY

	$1000*
Process Plant Investment and General Facilities	686,265
Process Contingency	18,700
Project Contingency	101,688
Total Plant Investment	806,653

*Mid-1978 dollars

**100 percent plant design power output is 810,344 kW; methanol production rate is 2,283.4 tons/day or 7,705 FOE barrels/day

#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 6-4

PROCESS CONTINGENCIES

Unit	Process Contingency Percent (Both Cases)
Coal Handling, Preparation and Feeding	0
Oxidant Feed	0
Gasification	15
Ash Handling	5
Gas Cooling and Particulate Removal	0-20*
COS Hydrolysis	0
Acid Gas Removal	0
Sulfur Recovery (Claus)	0
Tail Gas Treating	0
ZnO Treating	0
Methanol Synthesis	35
Steam, Condensate, BFW	0
Fuel Gas Expansion	0
Support Facilities	0
Combined Cycle	0

* 20 percent was applied to the high-temperature heat recovery equipment generating saturated steam in the raw gas-cooling section in Cases A2 and B2; 0 percent was applied to remaining low-temperature gas cooling equipment in both cases.

With the same total mass gas flow in Cases A2 and EXTC-79, the higher pressure operation in Case A2 requires fewer gasifiers of the same nominal dimensions. The refractory cost is independent of pressure. The metal cost for a single gasifier is proportional to pressure.

The cost of the high temperature gas cooling unit, including particulate removal in Case A2, is very similar to that in Case EXTC-79. The HP steam generator is the most expensive item in the gas cooling unit. Higher mass velocity due to higher pressure and higher allowable pressure drop in the HP steam generator enhances heat transfer in Case A2.

DISCUSSION OF METHANOL COPRODUCTION (B2)

Coal handling, oxidant feed, gasification and ash handling, and particulate removal are the same for both cases (A2 and B2). Salient features of these common units and other dissimilar units are discussed below by process unit.

Oxidant Feed

A cost comparison between a centrifugal compressor and a centrifugal-reciprocating compressor system for oxidant compression is shown below ($1000):

	Centrifugal	Centrifugal-Reciprocating
Compressors, Motor Drivers, and Intercoolers ($1000)	10,920	11,340

As can be seen, the magnitude of the difference in the costs is too small between the two systems to claim any distinct cost advantage.

Gas Cooling

In Case B2, synthesis gas feed to the methanol plant is heated to 656°F against gasifier effluent. The high hydrogen partial pressure and the high temperature mandate the use of alloy shell and tubes. On the other hand, in Case A2, fuel gas to the gas turbine is heated in Unit 51 against HP BFW. The difference in gas cooling costs between Cases A2 and B2 is attributed mainly to the differences in the approach temperatures used in the HP steam generator. It appears that significant capital cost reduction can be realized by optimizing temperature approaches in this exchanger.

Acid Gas Removal

Synthesis gas feed to the ZnO treating unit in the methanol plant contains no more than 5 ppmv sulfur. The Selexol plant processing this synthesis gas, as in Case B2, costs significantly more than a Selexol plant processing fuel gas to remove 90 percent sulfur as in Case A2.

COS Hydrolysis

Inclusion of COS hydrolysis in cases where a high degree of sulfur removal is required appears to have a significant advantage in initial investment. The addition of COS hydrolysis, in Case B2, makes both the acid gas removal and the sulfur recovery units less costly than the case in which a Selexol plant alone is used to remove the H_2S and the COS.

The following table presents a comparison of the costs of the COS hydrolysis unit, acid gas removal unit and the sulfur recovery unit in Case B2, against the acid gas removal unit and the sulfur recovery unit that would have been required in the absence of COS hydrolysis.

PLANT FACILITIES INVESTMENT ESTIMATES IN $1000

	Case B2 Without COS Hydrolysis	Case B2 With COS Hydrolysis
COS Hydrolysis Unit, Including the Feed Heater	-	5,854
Acid Gas Removal Unit	26,610	18,052
	26,610	23,906
Sulfur Recovery Units	27,641	21,633
	54,251	45,539

Sulfur Recovery Units

The cost trends between these units follow the cost trends in the acid gas removal units for both cases.

ZnO Treating

The cost of this unit is proportional to the amount of synthesis gas feed to the methanol plant.

TOTAL CAPITAL REQUIREMENT

The Total Capital Requirement for each case is defined as the sum of plant facilities investment, prepaid royalties, organization and start up costs, working capital, allowance for funds during construction and land costs. The plant facilities investment estimates for both cases studied (A2 and B2) have already been detailed in Table 6-2 and 6-3 (in mid-1978 dollars). The bases for estimating allowances for all of the other capital charges specified above are shown in Table 6-5.

The Total Capital Requirements as defined above for both the A2 and the B2 designs, as well as for the scaled-up B2 design, are shown in Table 6-6, expressed in both mid-1978 dollars and mid-1980 dollars. 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 escalation in installed costs for similar types of equipment in the two year period, mid-1978 to mid-1980).

The constant dollar mid-1980 allowances for funds during construction (AFDC) were calculated as follows (See Table C-1, Appendix C for details):

- Assuming construction to commence in January 1986 and end in December 1989 (4 year period), and assuming annual construction expenditures to be 15%, 25%, 35%, and 25% of the total funds required for the 4 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 (i.e., utility company ownership of both plants analyzed has been assumed).

- The total current dollar investment from Table C-1 (which includes escalated amounts for prepaid royalties, organization and start up 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 6-6.

- 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.

Table 6-5

BASES FOR ESTIMATING CAPITAL CHARGES

Item	Basis
Prepaid Royalties	0.5 percent of the Plant Facilities Investment.
Organization and Start-Up Costs	The organization and start up 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 start up. An allowance of 3 percent of the plant facilities investment should be made to cover organization and start up 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 6-6

TOTAL CAPITAL REQUIREMENT--METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

	Mid-1978 ($1,000)			Mid-1980 ($1,000)*		
	Conventional Texaco Based GCC Power Plant With No Methanol Production A2	Texaco GCC Plant With Once Through Methanol Coproduction (As Designed) B2	Methanol Coproduction Case (Same Design As Case B2), Scaled To Produce Same Quantity Of Electricity As Case A2, B2-Scaled-Up	Conventional Texaco Based GCC Power Plant With No Methanol Production A2	Texaco GCC Plant With Once Through Methanol Coproduction (As Designed) B2	Methanol Coproduction Case (Same Design As Case B2), Scaled To Produce Same Quantity Of Electricity As Case A2, B2-Scaled-Up
Plant Facilities Investment	859,010	806,653	1,101,485	1,090,943	1,024,449	1,398,886
Prepaid Royalties	4,295	4,033	5,507	5,455	5,122	6,994
Organization and Start Up Costs	25,770	24,200	33,045	32,728	30,734	41,967
Working Capital	15,743	15,915	21,733	19,993	20,212	27,600
Land	787	787	1,075	1,000	1,000	1,366
AFDC**	32,364	30,437	41,562	41,102	38,655	52,784
TOTAL CAPITAL REQUIREMENT ($$10^3$)	937,969	882,025	1,204,407	1,191,221	1,120,172	1,529,597
$/kW†	847.68	--	--	1,076.55	--	--
Incremental Total Capital Requirement for Methanol Production, $/FOEB#/Day##	--	--	25,327	--	--	32,165

*In general, the mid-1978 dollar costs were increased by 27 percent to provide the mid-1980 cost estimates. Working capital in mid-1980 dollars was calculated assuming 27 percent escalation in labor and operating costs and 14 percent escalation in coal costs for the two year period.

**See Page 6-8 for explanation of the calculation procedure.

†Based on 100 percent plant capacity factor.

#Barrels of distillate fuel oil (5.85×10^6 BTU/BBL) with higher heating value equivalent to methanol produced.

##Difference between the Total Capital Requirement for the scaled Case B2 and Case A2, divided by the total quantity of methanol produced by the scaled Case B2 design.

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 in Table C-1 are examined, it can be seen that they represent approximately 22.8 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 6-6 shows that for Case A2, the constant dollar mid-1980 plant facilities investment is \$1,090.943 x 10^6.

However, if the escalated December 1989 investment of \$2,282.879 x 10^6 is de-escalated at 10 percent/year for 9-1/2 years, an apparent mid-1980 investment of \$923.108 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 6-6, 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 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 bases for calculating both fixed and variable O&M charges are delineated in Tables 6-7 and 6-8. A summary of both mid-1978 and mid-1980 operating costs for the dedicated GCC plant (Case A2), the "once-through" methanol plant (Case B2) and the scaled-up "once-through" methanol plant (producing the same quantity of electricity as Case A2) is shown in Table 6-9.

Table 6-7

BASES 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 6-8. 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 labor. This overhead charge is 30 percent of the sum of the operating and maintenance labor.
Variable Operating Costs	The variable operating costs are dependent upon the plant capacity factor and are composed of the following charges: • Raw water • Catalysts and chemicals and other consumables • Ash and other waste disposal These items are discussed below:

Table 6-7

BASES FOR CALCULATING OPERATING AND MAINTENANCE COSTS
(Continued)

Item	Basis
Raw Water	The first-year raw water acquisition cost is 50¢/1000 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 non-hazardous wastes only.

Table 6-8

PLANT MAINTENANCE COSTS

Process Unit	Maintenance Cost as a Percent of the Plant Facilities Investment
Coal Handling	3.0
Oxidant Feed	2.0
Gasification and Ash Handling	4.5
Gas Cooling	3.0
Acid Gas Removal and Sulfur Recovery	2.0
Fuel Gas Expansion and Air Compression	3.0
COS Hydrolysis	2.0
Methanol Synthesis	3.0
Steam, Condensate and BFW	1.5
Support Facilities	1.5
Combined Cycle	1.5

Table 6-9

ANNUAL OPERATING AND MAINTENANCE COST ESTIMATES

	Mid-1978 ($1,000)*			Mid-1980 ($1,000)*		
	Conventional Texaco Based GCC Power Plant With No Methanol Production A2	Texaco GCC Plant With Once Through Methanol Coproduction (As Designed) B2	Methanol Coproduction Case (Same Design As Case B2), Scaled To Produce Same Quantity Of Electricity As Case A2, B2-Scaled-Up	Conventional Texaco Based GCC Power Plant With No Methanol Production A2	Texaco GCC Plant With Once Through Methanol Coproduction (As Designed) B2	Methanol Coproduction Case (Same Design As Case B2), Scaled To Produce Same Quantity Of Electricity As Case A2, B2-Scaled-Up
FIXED OPERATING COSTS						
Operating Labor	3,863	4,139	5,651	4,906	5,256	7,177
Maintenance Labor	6,848	6,683	9,126	8,698	8,487	11,590
Maintenance Materials	10,273	10,024	13,688	13,046	12,731	17,384
A&S Labor	3,213	3,247	4,433	4,081	4,123	5,630
Total Fixed Costs	24,197	24,093	32,898	30,731	30,597	41,781
VARIABLE OPERATING COSTS (100% CF)						
Raw Water	1,842	1,680	2,294	2,339	2,134	2,913
Catalysts & Chemicals	4,429	6,073	8,293	5,625	7,713	10,532
Ash Disposal	1,380	1,380	1,885	1,753	1,753	2,395
Total Variable Costs	7,651	9,133	12,472	9,717	11,600	15,840

*O&M costs in mid-1980 dollars were estimated on the basis of the information in Tables 6-7 and 6-8. The mid-1978 dollar values were calculated by assuming the mid-1980 costs to be 27 percent greater than the mid-1978 costs.

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

	Power Production Case A2	Methanol/Power Coproduction Case B2
Control Room Operators	5	5
Field Operators	17	20
Foremen	2	2
Lab and Instrument Technicians	4	4
	28	30

It is anticipated that incorporation of the methanol plant will require additional field operators. In determining labor requirements, modern computer assistance (cathode ray tube consoles) is assumed.

More than 80 percent of the catalyst and chemicals cost is made up of fuel oil (46 percent), corrosion inhibitor (20 percent), and surfactant (15 percent) in the power production Case A2. In the methanol/power coproduction Case B2, the catalyst and chemicals cost is largely made up of fuel oil (30 percent), methanol plant catalyst (25 percent), corrosion inhibitor (15 percent) and surfactant (11 percent). Fuel oil is required for startup of all combustion gas turbines, an average of 4.7 times yearly. For each startup, 24 hours is the allocated time of fuel oil firing. The corrosion inhibitor and surfactant are used exclusively in plant cooling water. Some minor chemical costs are associated with the following operations: process condensate treating, softening of cooling tower blowdown, acid gas removal, demineralization, COS hydrolysis in Case B2, sulfur recovery and tail gas treating. Less than two percent of total catalysts and chemicals cost is contributed by replacement of catalysts in the sulfur recovery and tail gas treating units.

Section 7
FINANCIAL ANALYSIS

Starting with this task, the Advanced Power Systems (APS) Division of EPRI is using a new method of financial analysis to that employed in previous Fluor studies. The new methodology, developed by the APS Division's Engineering and Economic Evaluations Program staff, has been employed by the EPRI Project Manager to prepare the results presented in this section of the report. Therefore, this financial analysis discussion has been written by the EPRI Project Manager.

The methodology used to determine revenue requirements from the methanol coproduction plant (B2) was to first calculate the revenue required for electricity production from the Texaco-based GCC power plant (A2), and then to credit such revenues to the methanol coproduction plant (B2). The remaining revenue required (in excess of the electricity credit) then represents the revenue requirement for the methanol coproduct.

As discussed in earlier sections of this report, both the A2 (GCC) and the B2 (methanol coproduct) plant designs performed by Fluor were based on the same coal feed rates of 10,000 tons/day. This resulted in a GCC power plant (A2) having a capacity of 1106.52 MW and a methanol coproduction system (B2) having an 810.34 MW power production capability. Because of the method of financial analysis described above, it was deemed convenient to scale up the Fluor design for the coproduction plant (B2) such that it would have the same electrical capacity (1106.52 MW) as the GCC plant (A2). Scale-up of the Fluor design was performed on a simple linear basis, i.e., all feedrates, product rates and costs generated by Fluor were multiplied by the ratio 1106.52/810.34. Plant performance characteristics for Case A2 and the scaled-up Case B2 have been presented earlier in this report and are reproduced in Table 7-1 for convenience.

Secondly, it is important to note that the Fluor generated plant facilities investment estimates and operating cost estimates for both plants were prepared in mid-1978 dollars. For the purpose of this study these estimates were escalated

Table 7-1

SYSTEM PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

CASE DESCRIPTION	Conventional Texaco Based GCC Power Plant With No Methanol Production	Texaco GCC Plant With Once-Through Methanol Coproduction (As Designed)	Methanol Coproduction Case (Same Design As Case B2), Scaled To Electricity As Case A2
CASE DESIGNATION	A2	B2	B2-Scaled-UP
GASIFICATION SYSTEM			
Coal Feed Rate, lbs/hr (MF)	798,333	798,333	1,090,124
Oxygen/Coal Feed*, lb/lb m.f.	0.858	0.858	0.858
Oxidant Temperature, °F	287	287	287
Slurry Water/Coal Ratio, lb/lb m.f.	0.503	0.503	0.503
Gasification Section Avg. Pressure, psig	1,000	1,000	1,000
Crude Gas Temp., °F	2,300-2,600	2,300-2,600	2,300-2,600
Crude Gas HHV (Dry Basis), BTU/SCF**	281.1	281.1	281.1
POWER SYSTEM			
Temp. of Fuel Gas to Gas Turbine, °F	339	372	372
Gas Turbine Inlet Temp., °F	2,000	2,000	2,000
Steam Conditions, psig/°F/°F	1450/900/900	1450/900/900	1450/900/900
Surface Condenser Pressure, in. Hg abs.	2.5	2.5	2.5
Stack Gas Exit Temp., °F	290	290	290
Gas Turbine Power#, MW	692.25	519.19	708.95
Steam Turbine Power#, MW	561.43	450.54	615.21
Fuel Gas Expander Power#, MW	36.64	26.37	36.01
Oxygen Plant Power##, MW	1.72	1.72	2.35
Power Consumed, MW	185.52	187.48	256.00
Net System Power, MW	1,106.52	810.34	1,106.52
METHANOL			
Methanol Produced, 10^3 gal/day	--	688.0	939.5
tons/day	--	2,283.4	3,118.0
FOEBδ/day	--	7,705	10,520
10^6 BTU/day	--	45,072	61,546
OVERALL SYSTEM			
Process and Dearerator Makeup Water, gpm	602	1,245††	1,700††
gpm/1000 MW	544	1,536	2,097
Cooling Tower Makeup Water†, gpm	8,299	6,875	9,388
gpm/1000 MW	7,495	8,478	11,577
Cooling Tower Heat Rejection, % of coal HHV	42.9	35.5	35.5
Air Cooler Heat Rejection, % of coal HHV	0.30	1.28	1.28
Net System Heat Rate, BTU/kWh	9,214	--	--
Overall System Efficiency (Coal + Power and Methanol), % of coal HHV	37.0	45.5	45.5
Efficiency of Methanol Production, % of coal HHV	--	68.8%	68.8%

*Dry basis, 100 percent oxygen
#At generator terminals
†Includes condensate from oxidant feed in Unit 11
δFOEB = Barrels of distillate fuel oil (5.8×10^6 BTU/BBL) with higher heating value equivalent to methanol produced
**Excluding HHV of H_2S, COS, NH_3
##From power recovery expander in 11-ME-1
††Includes water for steam injection into gas turbine

to mid-1980 dollars by incrementing them by 27 percent. This high equipment inflation rate for that particular two year period was chosen by EPRI after consultation with Fluor and other major engineering construction companies. Details of these capital and operating cost estimates for both plants can be found in Section 6 of this report. It is important to realize that all of the financial analysis presented in this section are based on the scaled-up Case B2 (coproduction) design and on the mid-1980 dollar capital and operating cost estimates.

As has been mentioned earlier, the cost of the coproduced methanol from Case B2 is determined by first calculating the revenue requirements for electricity production from Case A2. Because a new method of financial analysis is being used and because new financial parameters are being employed, the cost of electricity from plant A2 will first be discussed and compared with other EPRI cost estimates for similar GCC plants as well as a reference coal fired steam plant--all treated in the same manner. Such a comparison will act as a bridge between previous published EPRI results and the results of this study.

COST OF ELECTRICITY ESTIMATES FOR CASE A2

The plant design for Case A2 represents an integrated Texaco-based GCC system using a currently available (2,000°F) gas turbine. It is similar in most respects to two other GCC systems designed under EPRI funding, i.e.,

- A design completed by Fluor in 1980 and published in EPRI Report No. AP-1624 (See Case labeled EXTC-79), November, 1980.
- An independent design completed by R.M. Parsons in 1981 as part of Research Project 986-8 (See Case labeled Configuration A). A report of this effort is scheduled to be published in December, 1981.

The major difference between the above two GCC plant designs and Case A2 is in the selection of the average gasification system pressure. The Case A2 gasification plant has been designed to operate at an average pressure of 1,000 psig whereas the other two GCC systems have average gasification system pressures of 600 psig.

Finally, for comparative purposes, the performance and economic results for the above three Texaco-based GCC power plants will be compared with equivalent results generated for a conventional coal-fired steam plant with flue gas desulfurization. Major design and performance parameters for each of these four plants is shown in Table 7-2.

Table 7-2

DESIGN AND PERFORMANCE PARAMETERS FOR FOUR POWER PLANTS

	Texaco-Based GCC Plant This Report Case A2	Texaco-Based GCC Plant EPRI AP-1624 EXTC-79	Texaco-Based GCC Plant EPRI RP#986-8 Configuration A	Coal Fired Steam Plant with FGD EPRI AP-1725
Contractor	Fluor	Fluor	R. M. Parsons	Fluor/Bechtel
Net System Power, MW	1,106.52	1,095.75	1,080.77	987.18
Net System Heat Rate, BTU/kWh	9,214	9,404	9,534	9,981
Net System Thermal Efficiency, %	37.0	36.29	35.80	34.19
Slurry Water/Coal Ratio, lbs/lb m.f.	0.503	0.503	0.504	NA
Average Gasification Pressure, psig	1,000	600	600	NA
Gas Turbine Inlet Temperature, °F	2,000	2,000	2,000	NA
Steam Conditions, psig/°F/°F	1450/900/900	1450/900/900	1450/900/900	2400/1000/1000
Condenser Pressure, inches Hg	2.5	2.5	2.5	2.5
Fuel Gas Expander Used	Yes	No	No	NA

Costs of electricity for all four power plants described in Table 7-2 were calculated on the basis of criteria outlined in Table 7-3.

Detailed cost of electricity calculations for the Case A2 system are reproduced in Appendix C. The four tables reproduced in the appendix present a detailed capital outlay schedule, an annual capital recovery schedule, an annual revenue requirements schedule and an annual cash flow schedule.

Table 7-4 presents the costs of electricity results for Case A2 as well as a comparison of the Case A2 results with the other three designs.

The results of Table 7-4 lead to a number of interesting conclusions. First it is encouraging to see that the capital and operating cost estimates for the three Texaco-based GCC power plants are quite similar. The range of total capital requirement estimates for the three GCC configurations ($1,009/kW to $1,085/kW) is too small to indicate any significant differences between the three system designs. The conclusion to be drawn from this is that there appears to be no impact of operating pressure (in the range 600 psi to 1,000 psi) on system capital cost.

A second major conclusion to be reached based on the information presented in Table 7-4 is that the capital requirements for the GCC systems appear to be essentially the same as the capital required for the conventional coal fired steam plant. On the other hand, the cost of electricity estimates for the GCC plants indicate a 10 percent reduction over that estimated for the coal fired steam plant. This cost saving attributed to the GCC systems is due primarily to lower coal and O&M costs.

It has been stated previously that the financial calculations described in this section of the report are based on a new methodology and a new set of financial criteria developed by the APS Division. The new criteria and methodology were developed to facilitate the analysis and comparison of systems producing or coproducing products other than electricity (i.e., primarily liquid and gaseous fuels). EPRI has developed another set of financial criteria for analyzing and comparing costs of electricity generated in power plants. These criteria are to be published shortly in the 1981 Technical Assessment Guide (1981 TAG).

Table 7-3

FINANCIAL CRITERIA USED FOR INVESTOR OWNED UTILITY REVENUE REQUIREMENT CALCULATIONS

Plant Location	• Southern Illinois
Post-1980 General Inflation Rate	• 10 percent/Year
Plant Start Up	• 1990
Design and Construction Period	• 4 Years for GCC Plants 6 Years for Coal-Fired Plant
Project Book Life	• 30 Years
Project Tax Life	• 16 Years for GCC Plants 22 Years for Coal-Fired Plant
Tax Depreciation Method	• Sum-of-the-Year Digits
Net Plant Salvage Value	• 10 percent of PFI
Delivered Coal Cost in 1980$	• $1.30/10^6$ BTU
Real Coal Price Escalation (Above General Inflation)	• 1 percent/Year
Property Tax Rate	• 2 percent/Year of Escalated PFI
Insurance Rate	• 1 percent/Year of Escalated PFI
Federal Income Tax Rate	• 46 percent
State Income Tax Rate	• 6 percent
Investment Tax Credit	• 10 percent of Escalated PFI Normalized Over Period of Commercial Operation
Project Financing:	
Common Equity	• 35 percent at 16 percent/Year After Tax Return
Preferred Stock	• 15 percent at 12.75 percent/Year Dividend
Debt	• 50 percent at 12.25 percent/Year Interest

Table 7-4

COST OF ELECTRICITY RESULTS-70% CAPACITY FACTOR
TEXACO-BASED GCC PLANTS AND COAL FIRED STEAM PLANTS #

	Texaco-Based GCC Plants (2,000°F Gas Turbine)						Coal Fired Steam Plant with FGD From EPRI Report AP-1725*	
	Case A2 From This Report. (Fluor)		Case EXTC-79 From EPRI Report AP-1624 (Fluor)*		Configuration A From EPRI RP 9d6-8 (R. M. Parsons)			
Net System Capacity, MW	1,106.52		1,095.75		1,080.80		987.18	
Net System Heat Rate, BTU/kWh	9,214		9,404		9,534		9,981	
Net System Efficiency, %	37.04		36.29		35.80		34.19	
	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars
Total Capital Requirement for Start-Up in 1990, $/kW	2,662	1,077	2,496	1,009	2,683	1,085	2,547	1,030
Cost of Electricity, Mills/kWh								
First Year (1990)	128.84	47.36	123.63	45.45	129.14	47.47	143.19	52.64
Fifth Year (1994)	154.42	38.77	149.10	37.44	154.64	38.83	167.66	42.10
Tenth Year (1999)	204.65	31.90	199.20	31.05	204.77	31.92	219.15	34.17
Last Year (2019)	1005.60	23.30	1003.02	23.24	1006.95	23.33	1139.69	26.41
Levelized**	193.78	33.14	188.14	32.23	193.9[illegible]	33.17	211.38	36.15

*Plant Facilities Investment Estimate and Operating Cost Estimates from these reports, presented in mid-1978 dollars were escalated by 27 percent to bring them to mid-1980 dollars.

**A levelized revenue requirement is one which if held constant will yield the same return on common equity as the varying year-by-year values.

#Cost of electricity results for all four of these cases based on financial parameters outlined in the EPRI 1981 Technical Assessment guide can be found in Appendix B, Table B-4.

As this section of the report discusses the cost of electricity from a variety of electric power plants, capital and operating costs for all of these systems have been determined using the financial criteria outlined in the 1981 TAG. Details of these financial analyses can be found in Appendix B, Table B-4.

COST ESTIMATES FOR METHANOL COPRODUCTION (CASE B2)

It has been stated elsewhere that the cost of methanol coproduced in the Case B2 design would be estimated by first assessing annual revenues for the production of 1106.52 MW of electricity from Case A2 and then crediting those revenues to the total revenue requirements for the scaled-up Case B2 plant which has a capacity of 1106.52 MW plus a capability of producing 939.5 x 10^3 gallons/day of methanol. The net annual revenue requirements calculated in this manner represent the incremental revenues directly attributable to methanol production.

Performance characteristics of the scaled-up methanol coproduction plant (Case B2) are shown in Table 7-1. Capital and operating and maintenance costs for this plant (in both mid-1978 dollars and mid-1980 dollars) are shown in Tables 6-6 and 6-9 respectively.

The cost of coproducing methanol from the scaled-up Case B2 plant was determined on the basis of financial criteria shown in Table 7-3. Detailed cost of electricity and cost of methanol calculations are reproduced in Appendix D. The four tables for each case present detailed capital outlay schedules, annual capital recovery schedules, annual revenue requirement schedules and annual cash flow schedules.

Table 7-5 presents the estimated costs of electricity and methanol for Case A2 and Case B2.

Some points that must be kept in mind when examining the methanol costs shown in Table 7-5 are:

- The projected costs of methanol are representative of what would be anticipated for a coproduction plant (Case B2) owned, operated and financed by a regulated private utility company.
- It has been assumed that a methanol/electricity coproduction plant of the type of Case B2 would be operated at a capacity factor that is greater than that traditionally assumed for baseload fossil plants due to the high value of the coproduced methanol fuel. For this analysis, a 90 percent capacity factor has been assumed.

Table 7-5

COST OF ELECTRICITY AND COST OF METHANOL RESULTS - 90% CAPACITY FACTOR

	Texaco-Base GCC Plant Case A2	Texaco-Based Methanol Coproduction Plant Case B2 (Scaled-Up)
Net Electricity Capacity*, MW	1,106.52	1,106.52
Methanol Capacity*:		
10^3 gal/day	--	939.5
FOEB[a]/day	--	10,520
10^6 BTU/day	--	61,546

	Current Dollars Mills/kWh	Mid-1980 Dollars Mills/kWh	Current Dollars# $/$10^6$ BTU	Mid-1980 Dollars# $/$10^6$ BTU	Mid-1980 Dollars# ¢/gallon	Mid-1980 Dollars# $/FOEB[a]
Cost of Electricity and Cost of Methanol						
First Year (1990)	108.41	39.85	15.18	5.58	36.6	32.65
Fifth Year (1994)	132.57	33.28	19.10	4.80	31.4	28.08
Tenth Year (1999)	180.20	28.09	26.83	4.18	27.4	24.45
Last Year (2019)	952.75	22.08	153.23	3.55	23.3	20.77
Levelized**	170.26	29.11	25.24	4.32	28.3	25.27

*At 100% design capacity.

[a]Barrels of distillate fuel oil (5.85×10^6 BTU/BBL) with higher heating value equivalent to methanol produced.

**A levelized cost is one which if held constant will yield the same return on common equity as the varying year-by-year values.

#These cost estimates for methanol from the Case B2 Coproduction Plant reflect a credit for electricity produced and sold at the year-by-year prices shown in the two left-hand columns of this Table.

In order to compare the costs of methanol produced in the "once-through" mode shown in Table 7-5, with the alternative of producing methanol from coal in a dedicated mode (i.e., not coproduced with electricity), the following analysis has been performed. Recently, Fluor prepared a cost estimate for a dedicated coal-to-methanol plant (based on Texaco gasification of Illinois No. 6 coal) for EPRI. Details of the Fluor design and cost estimate for this dedicated methanol plant can be found in EPRI report AP-1962, August, 1981.

Major design and performance characteristics of the dedicated coal to methanol plant design are summarized and compared with similar characteristics of the "once-through" methanol section of the Case B2 design in Table 7-6. The most interesting difference shown in this table is the fact that the thermal efficiency of the "once-through" methanol concept is projected to be substantially higher than the efficiency of the dedicated coal-to-methanol plant. This increase in efficiency is due primarily to the elimination of losses created by the requirements for shift conversion, CO_2 removal and gas recycle in the dedicated methanol plant.

A financial analysis was conducted to determine the cost of producing methanol from coal in the dedicated methanol plant described above. Detailed cash outlay schedules and annual cash flow schedules are shown in Appendix E. It is most important to realize that the financial parameters employed for the dedicated methanol production case are representative of those used by a nonregulated private company. Differences between regulated and nonregulated financial parameters used for this study are shown in Table 7-7.

Table 7-8 presents a comparison of the cost of producing methanol by the "once-through" technique in a regulated utility owned facility with the anticipated minimum selling price of methanol produced in a nonregulated company owned dedicated coal-to-methanol plant.

A fundamental difference in cost/pricing policies between regulation and non-regulated companies must be described before the methanol costs/prices shown in Table 7-8 can be compared.

A regulated utility company is required to sell its regulated product at actual year-by-year costs of production. These costs include year-by-year fuel cost, operating and maintenance costs and fixed charges. Fixed charges are capital

Table 7-6

DESIGN AND PERFORMANCE CHARACTERISTICS OF
DEDICATED AND "ONCE-THROUGH" METHANOL PLANTS

	Dedicated Methanol Plant Fluor Design EPRI AP-1962 August, 1981	Once-Through Methanol Design Case B2 (Scaled-Up) This Report
Gasifiers	Texaco	Texaco
Coal Type	Illinois #6	Illinois #6
Coal Slurry Solids, %	?	66.7
Shift Conversion Employed	Yes	No
CO_2 Removal	Yes	No
Methanol Synthesis	ICI	Chem Systems Liquid Phase
Gas Recycle Employed	Yes	No
Total Coal Feed Rate, lbs/hr (MF)	1,204,000	1,090,124
Net Electricity Produced, MW	0	1,106.52
Methanol Produced, 10^3 gal/day	3,228.6	939.5
tons/day	10,927	3,118.0
FOEB[a]/day	36,154	10,520
10^6 BTU/day	211,500	61,546
Efficiency of Methanol Production, % of Coal HHV	57.86	68.80

[a]Barrels of distillate fuel oil (5.85×10^6 BTU/BBL) with higher heating value equivalent to methanol produced.

Table 7-7

FINANCIAL PARAMETERS USED TO DETERMINE METHANOL PRICES AND COSTS

	Regulated Utility Owned Methanol Electricity Coproduction Plant, Case B2	Nonregulated Company Owned Dedicated Coal-to-Methanol Plant
Plant Location	Southern Illinois	Southern Illinois
Post-1980 General Inflation Rate	10%/Year	10%/Year
Plant Start Up	1990	1990
Design and Construction Period	4 Years	5 Years
Project Book Life	30 Years	20 Years
Project Tax Life	16 Years	13 Years
Net Plant Salvage Value	10% of PFI	10% of PFI
Delivered Coal Cost in 1980$	$1.30/10^6 BTU	$1.30/10^6 BUT
Real Coal Price Escalation	1%/Year	1%/Year
Annual Property Tax Rate	2% of PFI	2% of PFI
Annual Insurance Rate	1% of PFI	1% of PFI
Federal Income Tax Rate	46%	46%
State Income Tax Rate	6%	6%
Investment Tax Credit	10% of PFI Normalized Over Period of Commercial Operation	10% of PFI Taken During the Construction Period
Common Equity	35% at 16%/Year After Tax Return	100% at 20%/Year After Tax Return
Preferred Stock	15% at 12.75%/Year Dividend	0
Debt	50% at 12.25%/Year Interest	0

Table 7-8

PRODUCTION COST AND SELLING PRICE ESTIMATES
FOR METHANOL PRODUCTION-90% CAPACITY FACTOR

	Regulated Utility Owned Methanol Electricity Coproduction Plant, Case B2		Nonregulated Company Owned Dedicated Coal-to-Methanol Plant	
	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars
Total Capital Requirement for 1990 Start Up, \$/FOEB[a]/day	79,545	32,165	103,142	41,666
Methanol Cost/Price				
First Year (1990) \$/10^6 BTU	15.18	5.58	21.41	7.87
¢/gallon	99.4	36.6	140.3	51.6
Fifth Year (1994) \$/10^6 BTU	19.10	4.80	31.35	7.87
¢/gallon	125.1	31.4	205.4	51.6
Tenth Year (1999) \$/10^6 BTU	26.83	4.18	50.49	7.87
¢/gallon	175.8	27.4	330.8	51.6
Twentieth Year (2009) \$/10^6 BTU	60.30	3.62	130.96	7.87
¢/gallon	395.0	23.7	857.9	51.6
Levelized* \$/10^6 BTU	25.24	4.32	36.26	7.87
¢/gallon	165.4	28.3	237.5	51.6

[a]Barrels of distillate fuel oil (5.85 x 10^6 BTU/BBL) with higher heating value equivalent to methanol produced.

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

related. As time goes by, the unrecovered capital investment decreases and therefore it should be anticipated that the cost of a regulated product will decrease in constant dollars. The second column in Table 7-8 (methanol cost in constant mid-1980 dollars) shows this characteristic, i.e., the constant dollar cost of producing the "once-through" methanol decreases from $\$5.58/10^6$ BTU in 1990 to $\$3.62/10^6$ BTU in 2009 due to a dramatic decrease in the unrecovered capital investment for the plant.

On the other hand, when a nonregulated private company produces a product, that product is generally sold at the "competitive market price." Therefore, the prices in the last two columns of Table 7-8 have been calculated as follows: A first year (1990) selling price was determined such that if it were escalated at the general inflation rate 10 percent per year for the life of the project (20 years), the net after tax discounted cash flow rate of return to the equity owner would be exactly 20 percent. The selling prices indicated in Table 7-8 imply that the equity holder will be satisfied with a 20 percent return (9.09 percent above inflation) on investment. If the return on equity requirement is higher, the initial selling price of methanol will increase significantly (as will be shown by the sensitivity analysis presented later).

Comparing the cost of producing methanol in a utility owned methanol/electricity coproduction plant with the anticipated selling price of methanol produced in a nonutility owned dedicated coal to methanol facility, a number of interesting features emerge:

- The first year cost of "once-through" methanol has the potential to be 30 percent lower than the expected minimum selling price of methanol produced by a nonregulated company. Such a saving translates into a saving of \$50 million/year for a utility consuming 10,000 bbl/day of liquid fuel.
- After the first year of operation, the cost of methanol produced by a utility in a methanol/electricity coproduction plant decreases (in constant dollars) with time from $\$5.58/10^6$ BTU in 1990 to $\$3.62/10^6$ BTU (both in mid-1980 dollars) in 2009. The nonutility produced methanol, however, will, at best, maintain its constant dollar price of $\$7.87/10^6$ BTU for the twenty year period. If liquid fuels escalate in price at a rate higher than the general inflation rate, the constant dollar $\$7.87/10^6$ BTU will increase proportionately.
- The average (levelized) constant dollar cost of the coproduced methanol ($\$4.32/10^6$ BTU) represents a saving of 45 percent over the average constant dollar selling price of $\$7.87/10^6$ BTU for nonutility produced methanol.

The final conclusion to be derived from this analysis is that the potential benefits to the utility industry to coproducing "once-through" methanol and electricity could be extremely large. It is critical to keep in mind, however, the fact that the "once-through" methanol process described in this report only exists at a small experimental level. This work simply demonstrates the potential economic benefits that could be realized if the Chem Systems "once-through" methanol process could be successfully developed at commercial scale, or if one of the currently existing commercial methanol synthesis processes could be modified to operate in the "once-through" mode. EPRI is currently investigating this second alternative.

SENSITIVITY ANALYSES

The costs of methanol presented in Table 7-8 must be considered to be speculative at this point as none of the plants being evaluated has ever been constructed and successfully operated at any scale. It is also important to realize that current political and economic uncertainties make it impossible to project, with any degree of accuracy, what financial factors (i.e., inflation rates, interest rates, returns on equity, governmental assistance, etc.) will exist in 1990 when all of these systems have been assumed to start operation. Therefore, a number of sensitivity studies have been conducted to determine the impact of different technical and economic conditions on the estimates of methanol and electricity prices presented previously.

Table 7-9 presents sensitivities of <u>first year</u> product costs to design and financial factors whereas Table 7-10 presents the sensitivities of <u>levelized</u> (or life cycle) costs to these factors. Looking at Table 7-9, sensitivities of <u>first year</u> product costs, the following conclusions can be reached:

- A lower than 10 percent inflation rate will tend to decrease the first year constant dollar cost of methanol coproduced by a regulated utility significantly, whereas it would have no impact on the first year constant dollar price of dedicated methanol produced by a nonregulated company.
- A two year startup delay would significantly increase the first year constant dollar price of nonregulated methanol production and would have essentially no impact on the first year constant dollar cost of regulated methanol coproduced with electricity.
- The base case first year methanol cost/price results indicate that regulated company coproduction costs would be approximately 30 percent lower than nonregulated company selling prices for dedicated methanol production. The results of Table 7-9 show that

40 percent of the 30 percent reduction is due to the different methods of financing the two plants (i.e., regulated vs non-regulated financing) whereas the other 60 percent of the 30 percent cost reduction is due to more efficient conversion of the coal to methanol via the "once-through" route.

- Finally, it is of great interest to note that the first year constant dollar selling price of dedicated methanol is acutely sensitive to the return on equity required by the nonregulated producer, i.e., changing the required return on equity from 20 percent to 30 percent would increase the required selling price of methanol from $\$7.87/10^6$ BTU to $\$12.72/10^6$ BTU.

Table 7-9

SENSITIVITIES OF FIRST YEAR PRODUCT COSTS TO DESIGN AND FINANCIAL FACTORS
CONSTANT MID-1980 DOLLARS

	Texaco-Based GCC Power Plant-Case A2 This Report. Electricity Cost		"Once-Through" Methanol/Electricity Coproduction Case B2 This Report. Methanol Case		Texaco-Based Dedicated Coal to Methanol Plant EPRI AP-1962. Methanol Price	
	First Year Mills/kWh	Percent Change From Base	First Year $/10^6 BTU	Percent Change From Base	First Year $/10^6 BTU	Percent Change From Base
Base Case Results*	39.85	Base	5.58	Base	7.87	Base
70% Capacity Factor Operation	47.36	+18.9	6.53	+17.0	9.35	+18.8
3% Escalation in Real Coal Price@	42.59	+ 6.9	6.01	+ 7.7	8.90	+13.1
35% Increase in Plant Facilities Investment	47.52	+19.3	6.52	+16.9	9.30	+18.2
5% Annual Inflation Rate‡	31.53	-20.9	4.58	-17.9	7.84	- 0.4
2 Year Start-Up Delay	40.52	+ 1.7	5.67	+ 1.6	8.86	+12.6
Regulated Utility Financing#	--		--		6.94	-11.8
Thermal Efficiency Decreased by 10%**	--		--		8.75	+11.2
25% Investment Tax Credit	--		--		7.13	- 9.4
Loan Guarantee (75% Debt at 12.25%/Year)	--		--		5.75	-27.3
30% Return on Common Equity	--		--		12.72	+61.6

See Table 7-11 for explanatory notes

Table 7-10

SENSITIVITIES OF LEVELIZED PRODUCT COSTS TO DESIGN AND FINANCIAL FACTORS
CONSTANT MID-1980 DOLLARS

	Texaco-Based GCC Power Plant-Case A2 This Report. Electricity Cost		"Once-Through" Methanol/Electricity Coproduction Case B2 This Report. Methanol Case		Texaco-Based Dedicated Coal to Methanol Plant EPRI AP-1962. Methanol Price	
	Levelized Mills/kWh	Percent Change From Base	Levelized $\$/10^6$ BTU	Percent Change From Base	Levelized $\$/10^6$ BTU	Percent Change From Base
Base Case Results*	29.11	Base	4.32	Base	7.87	Base
70% Capacity Factor Operation	33.14	+13.8	4.84	+12.0	9.35	+18.8
3% Escalation in Real Coal Price@	36.73	+26.2	5.52	+27.8	8.90	+13.1
35% Increase in Plant Facilities Investment	32.69	+12.3	4.75	+10.0	9.30	+18.2
5% Annual Inflation Rate†	29.26	+ 0.5	4.35	+ 0.7	7.84	- 0.4
2 Year Start-Up Delay	29.59	+ 1.7	4.39	+ 1.6	8.86	+12.6
Regulated Utility Financing#	--		--		5.35	-32.0
Thermal Efficiency Decreased by 10%**	--		--		8.75	+11.2
25% Investment Tax Credit	--		--		7.13	- 9.4
Loan Guarantee (75% Debt at 12.25%/Year)	--		--		5.75	-27.3
30% Return on Common Equity	--		--		12.72	+61.6

See Table 7-11 for explanatory notes

Table 7-11

NOTES FOR TABLES 7-9 and 7-10

* Base Case parameters used for Cases A2 and B2 are itemized in Table 7-3 (Regulated Utility Financing) and those used for the dedicated coal to methanol plant are itemized in Table 7-7 (Nonregulated Private Company).

∂ A 3 percent inflation free coal price escalation rate is equivalent to a 13.3 percent actual escalation rate if general inflation is 10 percent/year.

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

	Cases A2, B2	Dedicated Coal-to-Methanol Plant
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 Rate	6.05%	6.05%

≠ For this analysis, the criteria shown in Table 7-3 were applied to the dedicated coal-to-methanol plant.

** The thermal efficiency of the dedicated coal-to-methanol plant was changed from 57.86 percent to 52.07 percent.

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Appendix A

COMBINED-CYCLE SYSTEM DETAILS

GENERAL

Two similar combined-cycle designs are used for the cases in this study. Performance for the combustion gas turbines, including net power generation and complete heat and material balance data, were supplied by EPRI computer calculations based on fuel gas flows, compositions, and temperatures determined by Fluor.

The basic steam cycle conditions were specified by EPRI, but other steam levels were selected by Fluor as appropriate to the process. Where possible, consistency was maintained with the Westinghouse designs in EPRI report AF-642.

Each of the combined-cycle systems consists of gas turbogenerators with their associated heat recovery steam generators (HRSGs), a steam turbogenerator, and auxiliary pumps, steam drums, and deaerator. All steam cycle equipment was sized to interface with the coal gasification process. Both cases employ an auxiliary generator driven by a low-pressure steam turbine. HRSG approach temperatures (40°F), pressure losses, and blade loadings used in the calculations all reflect current utility criteria for lowest cost of power. Design parameters have been selected to demonstrate that the heat balance reflects performance of equipment currently available for order.

A summary of the calculated power output at the generator terminals and heat rejected to the process and power plant cooling towers is in Table A-1. The calculated steam turbine power outputs include a two percent deduction for estimated mechanical and electrical losses as well as a radiation loss of three Btu/lb of the gas.

TECHNICAL CRITERIA

Process Interface

Flow rates, compositions, pressures, and temperatures of the fluids to the power block are based on the design of the process fuel plant and the methanol plant. Heat integration between the process units and the power block is used wherever possible for cost-effective utilization of energy.

While most of the gas and steam turbine parameters were held constant, the resulting integrated plants differ when available heat is utilized. The quantity and quality of waste heat provided to the power block are related to the fuel plant and methanol plant process.

Prime Cycle

In both cases, the fuel gas from the gasification process is delivered to the gas turbine valve at 245 psig. In Case B2, 295 psig steam is injected prior to combustion to limit the formation of NO_x.

Steam Bottoming Cycle

Steam Conditions. Steam conditions used for the three-section steam turbine are:

Turbine Throttle

Cases A2 and B2	1450 psig, 900°F superheat 385 psig, 900°F reheat
Condenser	2.5 inches Hg abs

Steam Generation. Other steam pressure levels are 445 psig, 115 psig, and 50 psig. The two major sources of heat, producing all 1450 psig steam, are the sensible heats from raw gasifier effluent and gas turbine exhaust.

Heat Recovery Steam Generator (HRSG) Conditions

Each gas turbine is coupled to two identical heat recovery steam generators. Low level heat recovery from the HRSG was calculated by maintaining flue gas stack temperature at 290°F. For heat balance purposes, boiler blowdown was neglected. The low-pressure flash gases from the process plant are used as reducing gases in the Beavon unit in all cases. In some cases, there is more flash gas available than the Beavon unit requires. These excess gases are burned as supplementary fuel in the HRSG.

For the detailed HRSG flow scheme, see the Units 50 and 51 flow diagrams and combined-cycle descriptions in the body of the report.

<u>Steam Driver</u>

A medium-pressure condensing turbine is used in each case for the main HP boiler feedwater pump. The spare HP boiler feedwater pump is motor driven.

COMPONENT DESCRIPTION

<u>Gas Turbine (50-1-GT-1)</u>. For both rotating blades and static parts, the gas turbine utilizes compressor bleed air that is cooled against IP feedwater producing IP steam. The following materials are used in high-temperature service for this machine with 2000°F nominal inlet temperature:

Combustion Liner and Transition Piece	- Hastelloy-X
Turbine Nozzles	- FSX-414
Turbine Buckets (Blades)	- IN738 (stages 1 and 2) U500 (stage 3)

The first-stage turbine buckets are coated to provide protection against oxidation and corrosion. The combustor outlet temperature of 2000°F is, within the design limits, for currently available turbines in peaking service. Some minor upgrading of rotor materials in the expander section may be necessary for baseload operation at this temperature.

<u>Gas Turbine Generator (50-1-G-1)</u>. Each gas turbine drives a suitably rated, 0.9 power factor, 0.58 short circuit ratio, three-phase, 60 hertz, 13.8 kV, 3600 rpm open-ventilated air-cooled generator. This generator employs Micapal and Mica Mat insulation systems for stator windings, and Nomex insulation for the rotor windings. These insulation systems are designed to operate at higher temperatures than conventional insulating materials. The generator utilizes conductor cooling in the rotor windings, wherein air flows through radial holes in the windings, so the cooling air is in direct contact with the copper conductor. By eliminating the thermal resistance of insulation and steel, this provides much more effective cooling than the conventional ventilating duct technique.

A tabulation of gas turbine performance and generator output is given in Table A-2.

Steam Bottoming Cycle

HRSG (51-1-B-1). Two HRSGs 51-1-B-1 are coupled with each gas turbine 50-1-GT-1, to recover heat from turbine exhaust gases. In addition to superheating and reheating duties, the HRSG generates saturated steam at two or three pressure levels, depending upon the case. HP steam at 1500 psig, MP steam at 115 psig, and VLP steam at 15 psig are always produced.

The HP saturated steam generated in the HP evaporator is heated to 900°F in the superheater E-1. HP saturated steam available from the process is combined with the HP steam from the HRSG before entering the superheater.

Saturated steam produced in the various IP steam generators is combined with cold reheat steam from the high-pressure steam turbine and superheated to 900°F by passing through the reheater. Steam from the MP generator is added to IP steam turbine exhaust, then enters the condensing turbines.

One common tray-type deaerator, operating at 15 psia, serves the multiple HRSGs and process steam system. Each HRSG is provided with its own MP and HP steam drums and corresponding boiler feedwater circulation pumps.

HRSG design is based on vertical finned tubes. Modular construction permits shipping in sections and minimizes installation costs. The HRSG exhaust gas temperature of 290°F, allows the gas side surface of the final coil to operate safely above the sulfur dioxide dew point. The performance of the HRSGs for each case is summarized in Table A-3.

Steam Turbine (51-T-1A&B and 51-T-2). A tandem compound, reheat turbine system, consisting of HP and IP stages 51-T-1A&B and two split-case MP ends 51-T-2, is used for both cases. This system is unconventional in that steam is not extracted for feedwater heating.

The HP end receives superheated steam at 1450 psig, 900°F and exhausts to the IP steam header at approximately 445 psig. The IP steam, after combining with gas turbine air cooler steam and sulfur plant steam, is reheated to 900°F in the HRSG reheaters and flows to the IP stage with an inlet condition of 385 psig, 900°F. The exhaust steam is at 93.8 psig.

The LP end 51-T-2 is a condensing type unit receiving steam at 93.8 psig and exhausts at 2-1/2 inches Hg absolute. The main surface condenser associated with 51-T-2 is designed for cooling water flow in two tube side passes with 80°F cooling water inlet temperature and 20°F temperature rise.

Low-pressure steam turbogenerator 51-T-3 has been provided to recover additional low-temperature process heat.

Generator (51-1-G-1). The primary steam turbine system consisting of 51-T-1A, 51-T-1B, and 51-T-2 drives a suitably rated generator: consisting of, 0.9 power, 0 to 58 short circuit ratio, three-phase 60 hertz, 24 kV 3600 rpm outdoor type.

Table A-1

POWER BLOCK PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANT

Methanol Coproduction	None	Yes
Case Designation	A2	B2
GENERATION		
Gas Turbine, MW	692.25	519.19
HP Steam Turbine, MW	546.90	434.19
LP Steam Turbine, MW	14.53	16.35
Fuel Gas Expanders, MW	36.64	26.37
Total, Power Block, MW	1,290.32	996.10
HEAT REJECTION TO TOWERS		
Process Cooling, 10^6 Btu/hr*	554.3	570.0
Power Block Heat Rejection, 10^6 Btu/hr	3,828.1	3,053.5
Total Heat Rejection, 10^6 Btu/hr	4,382.4	3,623.5

*Includes mechanical and electrical losses to cooling water

Table A-2

GAS TURBINE PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes
Case Designation	A2	B2
Compressor Suction Pressure, psia	14.4	14.4
Compressor Discharge Temperature, °F	700.5	700.5
Rotor Coolant Temperature, °F	450.0	450.0
Turbine Exhaust Pressure, psia	15.5	15.5
Compressor Air Flow, lb/s	5,702.7	3,836.8
Fuel Flow, lb/s	400.4	431.0
Turbine Exhaust Temperature, °F	966.9	982.4
Rotor Cooling Air Cooler Duty, 10^6 Btu/hr	52.1	35.0
Power Output, kW*	692,250	519,190
Total Exhaust Gas Flow, lb/s	6,103.1	4,267.8

*At generator terminals

Table A-3

HRSG PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes
Case Designation	A2	B2
Exhaust Gas Flow, lb/s	6,103.1	4,267.8
SH AND RH SECTIONS		
Exhaust Gas Temperature In, °F	959.6	979.0
SH Flow, lb/s	858.0	721.8
SH Temperature In, °F	598.0	598.0
SH Enthalpy In, Btu/lb	1,167.6	1,167.6
SH Temperature Out, °F	900.0	900.0
SH Pressure Out, psig	1,505.0	1,505.0
SH Enthalpy Out, Btu/lb	1,429.9	1,429.9
SH Duty, 10^6 Btu/hr	810.2	681.7
RH Flow, lb/s	933.7	746.4
RH Temperature In, °F	607.0	616.0
RH Enthalpy In, Btu/lb	1,305.6	1,311.2
RH Temperature Out, °F	900.0	900.0
RH Pressure Out, psig	385.0	385.0
RH Enthalpy Out, Btu/lb	1,470.1	1,470.1
RH Duty, 10^6 Btu/hr	553.0	426.9
Exhaust Gas Temperature, °F	728.5	714.0
HP EVAPORATOR SECTION		
Water Enthalpy In, Btu/lb	614.0	612.7
HP Steam Evap., lb/s	263.3	156.1
HP Steam from Gasifier WHB, lb/s	594.8	565.7
HP Drum Temperature, °F	598.0	598.0
HP Drum Pressure, psia	1,520.0	1,520.0
HP Steam Enthalpy Out, Btu/lb	1,167.6	1,167.6
HP Evap. Duty, 10^6 Btu/hr	524.2	311.8
Exhaust Gas Temperature, °F	638.0	638.0
HP ECONOMIZER SECTION A		
Water Flow, lb/s	1,003.1	721.8
Water Enthalpy In, Btu/lb	394.4	399.8
Water Flow to Unit 20, lb/s	594.8	565.7
Water Enthalpy Out, Btu/lb	614.0	612.7
Duty, 10^6 Btu/hr	793.0	553.3
Exhaust Gas Temperature, °F	499.0	501.0

Table A-3 (Continued)

HRSG PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes
Case Designation	A2	B2
IP ECONOMIZER SECTION		
Water Flow, lb/s	65.8	111.2
Water Enthalpy In, Btu/lb	319.2	319.2
Water Enthalpy Out, Btu/lb	440.0	397.4
Duty, 10^6 Btu/hr	28.6	31.3
Exhaust Gas Temperature, °F	494.0	493.0
HP ECONOMIZER SECTION B		
Water Flow, lb/s	1,003.1	721.8
Water Enthalpy In, Btu/lb	321.4	321.5
Water Enthalpy Out, Btu/lb	394.4	399.8
Duty, 10^6 Btu/hr	263.6	203.5
Exhaust Gas Temperature, °F	446.6	442.0
MP EVAPORATOR SECTION		
Water Enthalpy In, Btu/lb	319.2	319.2
MP Steam Evap., lb/s	87.3	49.9
MP Drum Temperature, °F	347.0	347.0
MP Drum Pressure, psia	129.4	129.4
MP Steam Enthalpy Out, Btu/lb	1,192.4	1,192.4
MP Evap. Duty, 10^6 Btu/hr	274.5	156.8
Exhaust Gas Temperature, °F	397.6	402.0
MP ECONOMIZER SECTION		
Water Flow, lb/s	166.4	168.0
Water Enthalpy In, Btu/lb	219.9	219.9
Water Flow to Process, lb/s		
Water Enthalpy Out, Btu/lb	319.2	319.2
Duty, 10^6 Btu/hr	59.5	60.0
Exhaust Gas Temperature, °F	387.0	387.0
HP ECONOMIZER SECTION C		
Water Flow, lb/s	858.0	721.8
Water Enthalpy In, Btu/lb	224.1	224.1
Water Enthalpy Out, Btu/lb	321.5	321.5
Duty, 10^6 Btu/hr	300.8	253.1
Exhaust Gas Temperature, °F	332.7	323.0
LP EVAPORATOR AND DEAERATOR		
Deaerator Temperature, °F	250.0	250.0
Deaerator Pressure, psia	29.4	29.4
LP Steam Flow In, lb/s	3.2	1.9
LP Steam Enthalpy In, Btu/lb	1,179.4	1,179.4

Table A-3 (Continued)

HRSG PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes
Case Designation	A2	B2
LP EVAPORATOR AND DEAERATOR Continued)		
Condensate Flow In, lb/s	1,145.3	1,021.0
VLP Steam Evaporated, lb/s	69.2	37.3
VLP Evaporator Feedwater Enthalpy In, Btu/lb	217.8	217.8
VLP Steam Enthalpy Out, Btu/lb	1,163.9	1,163.9
Evaporator Duty, 10^6 Btu/hr	235.7	127.2
Exhaust Gas Temperature Out, °F	290.0	290.0

Table A-4

HP/IP/MP STEAM TURBINE PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes
Case Designation	A2	B2
HP BACK PRESSURE ELEMENT		
Throttle Conditions	1450/900	1450/900
Inlet Enthalpy, Btu/lb	1,429.9	1,429.9
Throttle Flow from HRSG, lb/s	858.0	721.8
Throttle Flow from Process, lb/s	0.0	0.0
Total Throttle Flow, lb/s	858.0	721.8
Shaft Seal Bypass, lb/s	4.8	4.1
Exhaust Enthalpy, Btu/lb	1,315.4	1,315.4
IP BACK PRESSURE ELEMENT		
Reheat Conditions	385/900	385/900
Reheat Flow, lb/s	933.7	746.4
Shaft Seal Bypass Flow, lb/s	1.1	0.9
Inlet Enthalpy, Btu/lb	1,470.1	1,470.1
Exhaust Enthalpy, Btu/lb	1,329.1	1,329.1
MP CONDENSING ELEMENT		
Inlet Conditions, psig/°F	93.8/572	93.8/580
Inlet Enthalpy, Btu/lb	1,314.4	1,318.6
Inlet Flow, lb/s	1,024.4	790.5
Flow to Condensers, lb/s	1,030.4	795.4
Exhaust Enthalpy, Btu/lb	1,022.5	1,025.1
Condensers Cooling Water Flow, gpm		
Total Power Output, kW*	546,900	434,191

*At generator terminals

Table A-5

LP CONDENSING STEAM TURBINE PERFORMANCE SUMMARY - METHANOL COPRODUCTION
OXYGEN-BLOWN TEXACO-BASED GCC PLANTS

Methanol Coproduction	None	Yes
Case Designation	A2	B2
Inlet Conditions,psig/°F	50/298	50/298
Inlet Enthalpy, Btu/lb	1,179.4	1,179.4
Inlet Flow, lb/s	73.9	83.1
Flow to Condenser, lb/s	73.9	83.1
Exhaust Enthalpy, Btu/lb	989.3	989.3
Condenser Cooling Water Flow, gpm	24,270	27,300
Power Output, kW*	14,525	16,347

*At generator terminals

Appendix B

COSTS OF ELECTRICITY CALCULATED ON THE BASIS OF THE 1981 TAG CRITERIA

The financial criteria used to generate cost of product estimates in this report were developed to cover a wide range of products (electricity, liquid and gaseous fuels) produced by a variety of different organizations (regulated utility companies, nonregulated private corporations).

EPRI, which is the research and development arm of the electric utility industry, is an organization which concentrates the bulk of its activities in the area of power generation by regulated utility companies. The Institute has developed a set of consistent financial criteria for the estimation of the cost of electric power generated by electric utility companies.

As this study contains capital and operating cost estimates for a number of GCC power generating systems, cost of electricity estimates are being presented in this appendix on the basis of the formal EPRI financial criteria for evaluating power plants. These EPRI criteria are detailed in the 1981 Technical Assessment Guide (TAG) to be published later this year.

Table B-1 presents a comparison of the financial parameters used to determine the cost of capital specified by the 1981 TAG with similar parameters used for the estimates presented in this report. Tables B-2 and B-3 present similar comparisons of criteria used to determine additional capital requirements and operating costs.

Capital requirements and 30 year levelized costs of electricity for the three GCC plants and the conventional coal fired plant discussed in this report have been calculated on the basis of the 1981 TAG criteria and are presented in Table B-4. These costs are equivalent to those presented in Tables 6-6 and 7-4 in the body of the report.

Table B-1

COMPARISON OF COST OF CAPITAL CRITERIA

	1981 TAG Criteria	APS Division Criteria (See Table 7-3)
	Quantity Used	Quantity Used
Debt/Equity Ratio	50/50	50/50
Common Equity	35%	35%
Preferred Stock	15%	15%
Debt	50%	50%
General Inflation Rate, %/Yr.	8.5	10.0
Common Equity Cost, %/Yr.	15.3	16.0
Preferred Stock Cost, %/Yr.	11.5	12.75
Debt Cost, %/Yr.	11.0	12.25
Weighted Cost of Capital, %/Yr.	12.5	13.64
Federal + State Income Tax Rate	50%	49.24%
Property Taxes & Insurance, %/Yr.	2	3
Investment Tax Credit, %	10	10
Plant Book Life, Years	30	30
Plant Salvage Value, %	0	10
Plant Tax Life		
Steam Plants	22.5	22
Combined Cycle Plants	--	16
30 Year Levelized Fixed Charge Rate, %/Yr.	16.1*	18.5-19.5**

* Based on using the before tax "weighted cost of capital" as the discount rate.

**Based on using the after tax return on common equity as the discount rate.

Table B-2

COMPARISON OF ADDITIONAL CAPITAL REQUIREMENTS CRITERIA

	1981 TAG Criteria	APS Division Criteria
Royalties	0.5% of Process Capital	0.5% of (Process Capital + General Facilities + Eng. & Home Office Costs + Initial Catalyst & Chemicals + Contingencies)
Preproduction Costs	1 Mo. Fixed Operating Costs	--
	1 Mo. Variable Operating Costs at Full Capacity	--
	25% of 1 Mo. Fuel Cost At Full Capacity	--
	2% of Plant Investment	3% of Plant Investment
Working Capital	1 Mo. of Fuel at 100% CF	1 Mo. of Fuel at 100% CF
	1 Mo. of Other Consumables	3 Mo. of Labor Costs
	--	1 Mo. of All Other Operating Costs at 100% CF
	--	Contingency of 25% of Sum of Above
Land (Dec. 1980$/Acre)	5500	5208
AFDC Charges	Determined at the Weighted Cost of Capital	Determined at the Interest Rate

Table B-3

COMPARISON OF OPERATING COST CRITERIA

	1981 TAG Criteria	APS Division Criteria
	Quantity Used (Dec. 1980$)	Quantity Used (July 1980$)
Operating Labor Rate, $/hr	16.43	20.00
Illinois No. 6 Coal (Deliv) $\$/10^6$ BTU	1.65	1.30
Coal Escalation Rate (Inflation Free), %/Year	0.8	1.0
Raw Water, ¢/1000 gal.	50	50
Dry Solids Disposal, $/ton	5.00	5.00
Sulfur Credit, $/long ton	60	0
Admin. & Support Labor, % of Operating & Maintenance Labor	30%	30%
Design Capacity Factor	65%	70%
Plant Start Up	Jan., 1981	Jan., 1990

It can be seen from the data in Tables 6-6 and B-4, that the total capital requirement for the Case A2 GCC plant is essentially the same, i.e., \$1149.3/kW in December 1980 dollars from Table B-4 and \$1076.55/kW in mid-1980 dollars (or \$1130.38/kW in December 1980) from Table 6-6.

The major difference between these two sets of criteria appears in the levelized cost of electricity calculation. For the Case A2 GCC plant, the 30 year current dollar levelized cost of electricity calculated using the 1981 TAG criteria shown in Table B-4 is 82.06 mills/kWh, whereas that shown in Table 7-4 is 193.78 mills/kWh. The primary reason for this large difference is the fact that the TAG criteria produce an "average" (levelized) current dollar cost in the time period 1981-2010 (i.e., for a plant starting up in 1981) whereas the costs presented in Table 7-4 are "average" (levelized) current dollar costs in the time period 1990-2020 (i.e., for a plant starting up in 1990). The important feature of these two different sets of financial criteria is that they result in the same relative differences in costs of electricity between systems, i.e.,

	Electricity Costs Presented in Table 7-4 In The Body Of This Report		Electricity Costs Presented in Table B-4 Based On The 1981 TAG Criteria	
	Levelized Cost of Electricity Mills/kWh	Relative Difference With Respect To Coal-Fired Plant	Levelized Cost of Electricity Mills/kWh	Relative Difference With Respect To Coal-Fired Plant
Case A2	193.78	- 8.33%	82.06	- 9.45%
Case EXTC-79	188.50	-10.82%	80.16	-11.54%
Configuration A	193.99	- 8.23%	81.25	-10.34%
Coal-Fired Plant	211.38	--	90.62	--

Table B-4

CAPITAL COSTS AND COST OF ELECTRICITY ESTIMATES BASED ON THE 1981 TAG* CRITERIA**

	December 1980 Dollars Texaco Based GCC Plants (2,000°F Gas Turbine)			
	Case A2 From This Report Fluor	Case EXTC-79 From AP-1624 Fluor	Configuration A From RP 986-8 R. M. Parsons	Coal Fired Steam Plant With FGD From EPRI AP-1725
Net System Capacity, MW	1,106.52	1,095.75	1,080.80	987.18
Net System Heat Rate, BTU/kWh	9,214	9,404	9,534	9,981
Net System Efficiency, %	37.04	36.29	35.80	34.19
CAPITAL INVESTMENT ($/kW)				
Total Plant Investment	1,041.837	974.932	1,022.830	973.522
Allowance for Funds During Construction	59.384	55.571	58.301	74.962
Royalties	4.072	3.768	3.901	0
Preproduction Costs	26.680	25.196	25.959	27.024
Inventory Capital	11.678	11.869	11.830	13.693
Initial Catalysts & Chemicals	4.695	4.973	0.786	1.229
Land	0.994	1.004	1.400	4.457
TOTAL CAPITAL REQUIREMENT	1,149.340	1,077.313	1,125.007	1,094.886
30 YEAR LEVELIZED COSTS (Mills/kWh at 65% Capacity Factor)				
Fixed Operating Costs	7.33	6.82	6.62	8.35
Variable Operating Costs	2.57	2.39	2.32	2.92
Consumables	2.40	2.26	1.72	6.07
Sulfur Credit	(1.81)	(1.65)	(1.65)	0
Coal Cost ($1.65/$10^6$ BTU)	39.07	39.88	40.43	42.32
Levelized Fixed Charges	32.50	30.46	31.81	30.96
LEVELIZED COST OF ELECTRICITY, MILLS/kWh	82.06	80.16	81.25	90.62

* *EPRI Technical Assessment Guide*, 1981.

** Equivalent data to those presented in this table can be found in Table 6-6 and 7-4.

Appendix C.

FINANCIAL ANALYSIS OF GCC PLANT (CASE A2) OPERATING AT A 70% CAPACITY FACTOR. REGULATED UTILITY OWNERSHIP.

1106 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/23/81.
70 PCT CF, -- 10 PCT INFL, -- 1990 STARTUP.

TABLE C3

CAPITAL OUTLAY SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

DESIGN/ CONSTRUCTION PERIOD (YEAR)	CALENDAR YEAR	PLANT FACILITIES INVESTMENT: IN MID-1986 DOLLARS	AMOUNT OF ESCALATION	ESCALATED INVESTMENT	ALLOWANCE FOR FUNDS DURING CONSTRUCTION: EQUITY	INTEREST	OTHER OUTLAYS*	TOTAL OUTLAY	GRANTS IN AID OF CONSTRUCTION	INVESTMENT TAX CREDITS	OTHER INCOME TAX OFFSETS	NET OUTLAY FOR PLANT
1.	1986.	163635.	126254.	289889.	--	--	1689.	291579.	0.	0.	0.	291579.
2.	1987.	272725.	258739.	531464.	--	--	0.	531464.	0.	0.	0.	531464.
3.	1988.	381815.	436639.	818454.	--	--	0.	818454.	0.	0.	0.	818454.
4.	1989.	272725.	370346.	643071.	--	--	140104.	793175.	0.	0.	0.	793175.
	TOTALS	1090900.	1191979.	2282879.	0.	521261.	141793.	2424672.	0.	224142.**	0.	2424672.

GROSS DEPRECIABLE INVESTMENT = 2894693.
NET NON-DEPRECIABLE PLANT OUTLAY = 51133.
TOTAL NON-DEPRECIABLE INVESTMENT = 51133.
TOTAL INVESTMENT = 2945932.

* PREPAID ROYALTIES, LAND, ORGANIZATION AND STARTUP EXPENSES, AND WORKING CAPITAL

** TO BE NORMALIZED OVER PERIOD OF COMMERCIAL OPERATION

GROSS DEPRECIABLE INVESTMENT = ESCALATED PLANT FACILITIES INVESTMENT LESS GRANTS-IN-AID OF CONSTRUCTION PLUS ALLOWANCE FOR FUNDS DURING CONSTRUCTION PLUS PRE-PAID ROYALTIES PLUS ORGANIZATION AND STARTUP EXPENSES

PLANT FINANCING:

COMMON EQUITY	= (.350)X(2945932.) =	1031076.
PREFERRED STOCK	= (.150)X(2945932.) =	441890.
DEBT	= (.500)X(2945932.) =	1472966.

		2945932.

1106 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/20/81.
70 PCT CF, -- 10 PCT INFL, -- 1990 STARTUP.

TABLE C2

CAPITAL RECOVERY SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

PERIOD OF COMMERCIAL OPERATION (YEAR)	CALENDAR YEAR	DEBT BALANCE (BEGINNING OF YR.)	DEBT PRINCIPAL PAYMENT	PREFERRED STOCK BALANCE (BEGINNING OF YEAR)	RECOVERY OF PREFERRED	COMMON EQUITY OUTSTANDING (BEGINNING OF YEAR)	ANNUAL RECOVERY OF COMMON EQUITY THROUGH BOOK DEPRECIATION	ANNUAL RECOVERY OF COMMON EQUITY OTHER
1.	1990.	1472966.	49099.	441893.	14730.	1031076.	23015.	0.
2.	1991.	1423867.	49099.	427163.	14730.	1008061.	23015.	0.
3.	1992.	1374768.	49099.	412431.	14730.	985045.	23015.	0.
4.	1993.	1325669.	49099.	397701.	14730.	962030.	23015.	0.
5.	1994.	1276571.	49099.	382971.	14730.	939014.	23015.	0.
6.	1995.	1227472.	49099.	368242.	14730.	915999.	23015.	0.
7.	1996.	1178373.	49099.	353512.	14730.	892983.	23015.	0.
8.	1997.	1129274.	49099.	338782.	14730.	869968.	23015.	0.
9.	1998.	1080175.	49099.	324053.	14730.	846953.	23015.	0.
10.	1999.	1031076.	49099.	309323.	14730.	823937.	23015.	0.
11.	2000.	981977.	49099.	294593.	14730.	800922.	23015.	0.
12.	2001.	932878.	49099.	279864.	14730.	777906.	23015.	0.
13.	2002.	883780.	49099.	265134.	14730.	754891.	23015.	0.
14.	2003.	834681.	49099.	250404.	14730.	731875.	23015.	0.
15.	2004.	785582.	49099.	235675.	14730.	708860.	23015.	0.
16.	2005.	736483.	49099.	220945.	14730.	685844.	23015.	0.
17.	2006.	687384.	49099.	206215.	14730.	662829.	23015.	0.
18.	2007.	638285.	49099.	191486.	14730.	639813.	23015.	0.
19.	2008.	589186.	49099.	176756.	14730.	616798.	23015.	0.
20.	2009.	540088.	49099.	162026.	14730.	593783.	23015.	0.
21.	2010.	490989.	49099.	147297.	14730.	570767.	23015.	0.
22.	2011.	441890.	49099.	132567.	14730.	547752.	23015.	0.
23.	2012.	392791.	49099.	117837.	14730.	524736.	23015.	0.
24.	2013.	343692.	49099.	103108.	14730.	501721.	23015.	0.
25.	2014.	294593.	49099.	88378.	14730.	478705.	23015.	0.
26.	2015.	245494.	49099.	73648.	14730.	455690.	23015.	0.
27.	2016.	196395.	49099.	58919.	14730.	432674.	23015.	0.
28.	2017.	147297.	49099.	44189.	14730.	409659.	23015.	0.
29.	2018.	98198.	49099.	29459.	14730.	386643.	23015.	0.
30.	2019.	49099.	49099.	14730.	14730.	363628.	23015.	0.
31.	2020.	0.	0.	0.	0.	340612.	0.	0.

1106 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/27/91.
70 PCT CF, -- 10 PCT INFL, -- 1990 STARTUP.

TABLE C3

REVENUE REQUIREMENTS SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

YR	CALENDAR YEAR	RETURN ON COMMON EQUITY	PREFERRED STOCK DIVIDENDS	INTEREST ON DEBT	INCOME TAXES	OTHER TAXES AND INSURANCE	RECOVERY OF CAPITAL: BOOK DEPRECIATION	RECOVERY OF CAPITAL: OTHER	FUEL/RAW MATERIAL COST	OPERATING AND MAINTENANCE COSTS	TOTAL REVENUE REQUIRED	REVENUE FROM BY-PRODUCTS	REVENUE FROM PRINCIPAL PRODUCT: TOTAL	MILLS PER KWH	MILLS PER KWH EXPRESSED IN MID-1990 DOLLARS
1	1990.	164972.	56341.	180438.	-13113.	68486.	86844.	0.	232864.	97351.	874134.	0.	874184.	125.84	47.36
2	1991.	161290.	54463.	174424.	76.	68486.	86844.	0.	258712.	107086.	911341.	0.	911341.	134.32	44.85
3	1992.	157607.	52585.	168459.	13265.	68486.	86844.	0.	287429.	117794.	952420.	0.	952420.	142.37	42.64
4	1993.	153925.	50707.	162395.	26454.	68486.	86844.	0.	319334.	129574.	997719.	0.	997719.	147.35	40.51
5	1994.	150242.	48829.	156360.	39644.	68486.	86844.	0.	354780.	142531.	1047735.	0.	1047736.	154.42	38.77
6	1995.	146560.	46951.	150365.	52833.	68486.	86844.	0.	394161.	156784.	1102984.	0.	1102984.	162.56	37.10
7	1996.	142577.	45073.	144351.	66022.	68486.	86844.	0.	437912.	172463.	1164025.	0.	1164028.	171.56	35.63
8	1997.	139195.	43195.	138336.	79211.	68486.	86844.	0.	486521.	189709.	1231495.	0.	1231496.	181.51	34.24
9	1998.	135512.	41317.	132321.	92400.	68486.	86844.	0.	540524.	208680.	1306635.	0.	1306635.	192.45	33.31
10	1999.	131630.	39439.	126307.	105569.	68486.	86844.	0.	600523.	229548.	1388555.	0.	1388565.	204.65	31.50
11	2000.	128147.	37561.	120292.	118778.	68486.	86844.	0.	667181.	252503.	1479792.	0.	1479792.	218.10	30.91
12	2001.	124465.	35683.	114278.	131967.	68486.	86844.	0.	741239.	277753.	1580713.	0.	1580713.	232.97	30.22
13	2002.	120783.	33805.	108263.	145157.	68486.	86844.	0.	823515.	305528.	1692393.	0.	1692386.	249.43	29.22
14	2003.	117100.	31927.	102248.	159346.	68486.	86844.	0.	914925.	336081.	1815957.	0.	1815957.	267.64	28.53
15	2004.	113418.	30049.	96233.	171535.	68486.	86844.	0.	1016482.	369689.	1952735.	0.	1952736.	287.[illegible]	27.86
16	2005.	109735.	28170.	90219.	184724.	68486.	86844.	0.	1129311.	406658.	2134145.	0.	2134145.	314.12	27.29
17	2006.	106053.	26292.	84205.	197913.	68486.	86844.	0.	1254665.	447324.	2271732.	0.	2271782.	334.82	26.79
18	2007.	102370.	24414.	78190.	192519.	68486.	86844.	0.	1393953.	492055.	2438813.	0.	2438813.	359.44	26.14
19	2008.	98688.	22536.	72175.	187125.	68486.	86844.	0.	1548659.	541262.	2625775.	0.	2625776.	386.99	25.57
20	2009.	95005.	20658.	66161.	181731.	68486.	86844.	0.	1720560.	595388.	2834834.	0.	2834834.	417.81	25.11
21	2010.	91323.	18780.	60146.	176337.	68486.	86844.	0.	1911543.	654927.	3068336.	0.	3068336.	452.23	24.71
22	2011.	87640.	16902.	54132.	170943.	68486.	86844.	0.	2123724.	720419.	3329091.	0.	3329091.	491.65	24.37
23	2012.	83958.	15024.	48117.	165549.	68486.	86844.	0.	2359457.	792461.	3619897.	0.	3619897.	532.51	24.09
24	2013.	80275.	13146.	42102.	160155.	68486.	86844.	0.	2621357.	871707.	3944074.	0.	3944074.	581.29	23.86
25	2014.	76593.	11268.	36088.	154761.	68486.	86844.	0.	2912328.	958878.	4305246.	0.	4305246.	634.52	23.65
26	2015.	72910.	9390.	30073.	149367.	68486.	86844.	0.	3235596.	1054766.	4727433.	0.	4727433.	693.79	23.54
27	2016.	69228.	7512.	24058.	143973.	68486.	86844.	0.	3594747.	1160243.	5155092.	0.	5155092.	758.77	23.43
28	2017.	65545.	5634.	18044.	138579.	68486.	86844.	0.	3993764.	1276267.	5653154.	0.	5653154.	833.18	23.36
29	2018.	61863.	3756.	12029.	133185.	68486.	86844.	0.	4437772.	1403893.	6207129.	0.	6207129.	914.82	23.32
30	2019.	58180.	1878.	6015.	127791.	68486.	86844.	0.	4929597.	1544283.	6823764.	0.	6823764.	1005.[illegible]	23.31
											LEVELIZED			192.79 *	33.14 **

LEVELIZED FIXED CHARGE RATE IN CURRENT DOLLARS = .195675

* LEVELIZED USING RETURN ON EQUITY OF 15.000 PCT./YEAR

** LEVELIZED USING RETURN ON EQUITY OF 5.455 PCT./YEAR

1106 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/29/81.
70 PCT CF, -- 10 PCT INFL, -- 1990 STARTUP.

TABLE C4

PROJECT CASH FLOW SCHEDULE FOR AN INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

		TOTAL REVENUE AT:		TAXES ON INCOME WITH REVENUE AT:									CASH FLOW TO COMMON EQUITY WITH REVENUE AT:	
YR	CALENDAR YEAR	PRICES NOT LEVELIZED	LEVELIZED PRICE *	PRICES NOT LEVELIZED	LEVELIZED PRICE *	OTHER TAXES AND INSURANCE	PREFERRED STOCK COST +	DEBT PRINCIPAL AND INTEREST	FUEL/RAW MATERIAL COST	OPERATING AND MAINTENANCE COSTS	COMMON EQUITY PORTION OF RECURRING INVESTMENT	PLANT SALVAGE VALUE, WORKING CAPITAL, AND LAND	PRICES NOT LEVELIZED	LEVELIZED PRICE *
1	1990.	874184.	1314831.	-13113.	203952.	69486.	71771.	225537.	232864.	97551.	0.	0.	137938.	411660.
2	1991.	911381.	1314831.	76.	198735.	68486.	69193.	223523.	258712.	107566.	0.	0.	184305.	389097.
3	1992.	952420.	1314831.	13265.	191716.	68486.	67515.	217538.	287429.	117794.	0.	0.	186623.	364563.
4	1993.	997719.	1314831.	26454.	182601.	68486.	65437.	211493.	319334.	129574.	0.	0.	175940.	337927.
5	1994.	1047735.	1314831.	39644.	171161.	68486.	63558.	205479.	354780.	142531.	0.	0.	173259.	306835.
6	1995.	1102984.	1314831.	52833.	157147.	68486.	61687.	199459.	394161.	156784.	0.	0.	159575.	277109.
7	1996.	1164528.	1314831.	66022.	140277.	68486.	59842.	193457.	437912.	172463.	0.	0.	155893.	242441.
8	1997.	1231495.	1314831.	79211.	122215.	68486.	57924.	187435.	486521.	189789.	0.	0.	162210.	204511.
9	1998.	1296185.	1314831.	92400.	96737.	68486.	56046.	181423.	541524.	208560.	0.	0.	159528.	162968.
10	1999.	1368565.	1314831.	105589.	69283.	68486.	54168.	175435.	600523.	229546.	0.	0.	154845.	117418.
11	2000.	1479792.	1314831.	118775.	37552.	68486.	52290.	169391.	667181.	252563.	0.	0.	151163.	67429.
12	2001.	1580713.	1314831.	131967.	1947.	68486.	50412.	153376.	741258.	277753.	0.	0.	147490.	12519.
13	2002.	1692381.	1314831.	145157.	-40746.	68486.	46534.	157362.	823515.	305520.	0.	0.	143778.	-47846.
14	2003.	1815957.	1314831.	158346.	-66429.	68486.	46656.	151347.	914925.	336301.	0.	0.	143116.	-114256.
15	2004.	1952736.	1314831.	171535.	-142559.	68486.	44778.	145335.	1016402.	369589.	0.	0.	136433.	-187367.
16	2005.	2104148.	1314831.	184724.	-213936.	68486.	42900.	139318.	1129311.	406556.	0.	0.	132751.	-267907.
17	2006.	2271782.	1314831.	197913.	-273259.	68486.	41022.	133303.	1254665.	447324.	0.	0.	129058.	-356600.
18	2007.	2436813.	1314831.	192519.	-367929.	68486.	39144.	127289.	1392933.	492056.	0.	0.	125386.	-445147.
19	2008.	2625776.	1314831.	187125.	-459394.	68486.	37266.	121274.	1548659.	541262.	0.	0.	121703.	-543732.
20	2009.	2834534.	1314831.	161731.	-566716.	68486.	35388.	115260.	1720560.	595388.	0.	0.	118021.	-653533.
21	2010.	3068395.	1314831.	176337.	-687113.	68486.	33510.	109245.	1911543.	654927.	0.	0.	114339.	-775766.
22	2011.	3329391.	1314831.	172943.	-826276.	68486.	31632.	103231.	2123724.	720419.	0.	0.	110656.	-911782.
23	2012.	3619697.	1314831.	165549.	-969465.	68486.	29754.	97216.	2359457.	792461.	0.	0.	106973.	-1063078.
24	2013.	3944074.	1314831.	160155.	-1134434.	68486.	27876.	91201.	2621357.	871707.	0.	0.	103291.	-1231313.
25	2014.	4305246.	1314831.	154761.	-1317719.	68486.	25998.	85187.	2912328.	958978.	0.	0.	99608.	-1418326.
26	2015.	4707433.	1314831.	149367.	-1521150.	68486.	24120.	79172.	3235596.	1054766.	0.	0.	95925.	-1626159.
27	2016.	5155092.	1314831.	143973.	-1746971.	68486.	22242.	73157.	3594747.	1160243.	0.	0.	92243.	-1857073.
28	2017.	5653164.	1314831.	138579.	-1997516.	68486.	20364.	67143.	3993764.	1276267.	0.	0.	88561.	-2113577.
29	2018.	6207129.	1314831.	133185.	-2275792.	68486.	18486.	61128.	4437072.	1403893.	0.	0.	84878.	-2398452.
30	2019.	6823964.	1314831.	127791.	-2584453.	68486.	16608.	55113.	4929587.	1544263.	0.	340612.	421806.	-2374171.

PRESENT VALUE AT BEGINNING OF 1990 OF CASH FLOWS TO COMMON EQUITY DISCOUNTED AT 16.00 PCT./YEAR:

WITH REVENUE AT LEVELIZED PRICE = $ 1031976250.
WITH REVENUE AT PRICES NOT LEVELIZED = $ 1031976250.

COMMON EQUITY OUTSTANDING AT BEGINNING OF 1990 = $ 1031976250.

* ONLY PRINCIPAL PRODUCT PRICE IS LEVELIZED, USING RETURN ON EQUITY OF 16.00 PCT./YEAR
\+ RECOVERY AND DIVIDENDS

1106 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/23/81,
BASE CASE -- 90 PCT CF. -- 10 PCT INFL. -- 1990 STARTUP.

TABLE D1

CAPITAL OUTLAY SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

DESIGN/ CONSTRUCTION PERIOD (YEAR)	CALENDAR YEAR	PLANT FACILITIES INVESTMENT: IN MID-1980 DOLLARS	PLANT FACILITIES INVESTMENT: AMOUNT OF ESCALATION	PLANT FACILITIES INVESTMENT: ESCALATED INVESTMENT	ALLOWANCE FOR FUNDS DURING CONSTRUCTION: EQUITY	ALLOWANCE FOR FUNDS DURING CONSTRUCTION: INTEREST	OTHER OUTLAYS*	TOTAL OUTLAY	GRANTS IN AID OF CONSTRUCTION	INVESTMENT TAX CREDITS	OTHER INCOME TAX OFFSETS	NET OUTLAY FOR PLANT
1.	1986.	163635.	126254.	289889.	--	--	1689.	291579.	0.	0.	0.	291579.
2.	1987.	272725.	258739.	531464.	--	--	0.	531464.	0.	0.	0.	531464.
3.	1988.	381815.	436639.	818454.	--	--	0.	818454.	0.	0.	0.	818454.
4.	1989.	272725.	370346.	643071.	--	--	140104.	783175.	0.	0.	0.	783175.
	TOTALS	1090900.	1191979.	2282879.	0.	521261.	141793.	2424672.	0.	224042.**	0.	2424672.

GROSS DEPRECIABLE INVESTMENT = 2894800.
NET NON-DEPRECIABLE PLANT OUTLAY = 51133.
TOTAL NON-DEPRECIABLE INVESTMENT = 51133.
TOTAL INVESTMENT = 2945932.

* PREPAID ROYALTIES, LAND, ORGANIZATION AND STARTUP EXPENSES, AND WORKING CAPITAL

** TO BE NORMALIZED OVER PERIOD OF COMMERCIAL OPERATION

GROSS DEPRECIABLE INVESTMENT = ESCALATED PLANT FACILITIES INVESTMENT LESS GRANTS-IN-AID OF CONSTRUCTION PLUS ALLOWANCE FOR FUNDS DURING CONSTRUCTION PLUS PRE-PAID ROYALTIES PLUS ORGANIZATION AND STARTUP EXPENSES

PLANT FINANCING:

COMMON EQUITY	= (.350)X(2945932.)	=	1031076.
PREFERRED STOCK	= (.150)X(2945932.)	=	441890.
DEBT	= (.500)X(2945932.)	=	1472966.

			2945932.

1105 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/23/81.
BASE CASE -- 70 PCT CF. -- 10 PCT INFL. -- 1990 STARTUP.

TABLE D2

CAPITAL RECOVERY SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

PERIOD OF COMMERCIAL OPERATION (YEAR)	CALENDAR YEAR	DEBT BALANCE (BEGINNING OF YR.)	DEBT PRINCIPAL PAYMENT	PREFERRED STOCK BALANCE (BEGINNING OF YEAR)	RECOVERY OF PREFERRED	COMMON EQUITY OUTSTANDING (BEGINNING OF YEAR)	ANNUAL RECOVERY OF COMMON EQUITY: THROUGH BOOK DEPRECIATION	ANNUAL RECOVERY OF COMMON EQUITY: OTHER
1.	1990.	1472966.	49099.	441890.	14730.	1031076.	23015.	0.
2.	1991.	1423867.	49099.	427160.	14730.	1008061.	23015.	0.
3.	1992.	1374768.	49099.	412431.	14730.	985045.	23015.	0.
4.	1993.	1325669.	49099.	397701.	14730.	962030.	23015.	0.
5.	1994.	1276571.	49099.	382971.	14730.	939014.	23015.	0.
6.	1995.	1227472.	49099.	368242.	14730.	915999.	23015.	0.
7.	1996.	1178373.	49099.	353512.	14730.	892983.	23015.	0.
8.	1997.	1129274.	49099.	338782.	14730.	869968.	23015.	0.
9.	1998.	1080175.	49099.	324053.	14730.	846953.	23015.	0.
10.	1999.	1031076.	49099.	309323.	14730.	823937.	23015.	0.
11.	2000.	981977.	49099.	294593.	14730.	800922.	23015.	0.
12.	2001.	932879.	49099.	279864.	14730.	777906.	23015.	0.
13.	2002.	883780.	49099.	265134.	14730.	754891.	23015.	0.
14.	2003.	834681.	49099.	250404.	14730.	731875.	23015.	0.
15.	2004.	785582.	49099.	235675.	14730.	708860.	23015.	0.
16.	2005.	736483.	49099.	220945.	14730.	685844.	23015.	0.
17.	2006.	687384.	49099.	206215.	14730.	662829.	23015.	0.
18.	2007.	638285.	49099.	191486.	14730.	639813.	23015.	0.
19.	2008.	589186.	49099.	176756.	14730.	616798.	23015.	0.
20.	2009.	540088.	49099.	162026.	14730.	593783.	23015.	0.
21.	2010.	490989.	49099.	147297.	14730.	570767.	23015.	0.
22.	2011.	441890.	49099.	132567.	14730.	547752.	23015.	0.
23.	2012.	392791.	49099.	117837.	14730.	524736.	23015.	0.
24.	2013.	343692.	49099.	103108.	14730.	501721.	23015.	0.
25.	2014.	294593.	49099.	88378.	14730.	478705.	23015.	0.
26.	2015.	245494.	49099.	73648.	14730.	455690.	23015.	0.
27.	2016.	196395.	49099.	58919.	14730.	432674.	23015.	0.
28.	2017.	147297.	49099.	44189.	14730.	409659.	23015.	0.
29.	2018.	98198.	49099.	29459.	14730.	386643.	23015.	0.
30.	2019.	49099.	49099.	14730.	14730.	363628.	23015.	0.
31.	2020.	0.	0.	0.	0.	340612.	0.	0.

1106 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/20/81.
BASE CASE -- 90 PCT CF. -- 10 PCT INFL. -- 1990 STARTUP.

TABLE D4

PROJECT CASH FLOW SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

YR	CALENDAR YEAR	TOTAL REVENUE AT: PRICES NOT LEVELIZED	TOTAL REVENUE AT: LEVELIZED PRICE *	TAXES ON INCOME WITH REVENUE AT: PRICES NOT LEVELIZED	TAXES ON INCOME WITH REVENUE AT: LEVELIZED PRICE *	OTHER TAXES AND INSURANCE	PREFERRED STOCK COST +	DEBT PRINCIPAL AND INTEREST	FUEL/RAW MATERIAL COST	OPERATING AND MAINTENANCE COSTS	COMMON EQUITY PORTION OF RECURRING INVESTMENT	PLANT SALVAGE VALUE, WORKING CAPITAL, AND LAND	CASH FLOW TO COMMON EQUITY WITH REVENUE AT: PRICES NOT LEVELIZED	CASH FLOW TO COMMON EQUITY WITH REVENUE AT: LEVELIZED PRICE *
1	1990.	945757.	1465259.	-13113.	252538.	68486.	71671.	229537.	299357.	102391.	0.	0.	187958.	461839.
2	1991.	990843.	1465259.	76.	243526.	68486.	69193.	223523.	332639.	112530.	0.	0.	184305.	435270.
3	1992.	1040542.	1465259.	13265.	232194.	68486.	67315.	217598.	369552.	123094.	0.	0.	180623.	406310.
4	1993.	1095565.	1465259.	26454.	210230.	68486.	65437.	211493.	410572.	136283.	0.	0.	175940.	374698.
5	1994.	1156482.	1465259.	39544.	201533.	68486.	63558.	205479.	456146.	149911.	0.	0.	173258.	340145.
6	1995.	1223719.	1465259.	52833.	161615.	68486.	61680.	199464.	505778.	164902.	0.	0.	169575.	302333.
7	1996.	1296076.	1465259.	66022.	155191.	68486.	59802.	193459.	563030.	181393.	0.	0.	165893.	261947.
8	1997.	1360325.	1465259.	79211.	136950.	68486.	57924.	187435.	625526.	199532.	0.	0.	162210.	215475.
9	1998.	1471326.	1465259.	92400.	99251.	68486.	56046.	181420.	694961.	219485.	0.	0.	158528.	165600.
10	1999.	1572929.	1465259.	105539.	62354.	68486.	54168.	175406.	772101.	241433.	0.	0.	154845.	119801.
11	2000.	1683489.	1465259.	118778.	21171.	68486.	52290.	169391.	857864.	265577.	0.	0.	151163.	59541.
12	2001.	1806577.	1465259.	131967.	-26398.	68486.	50412.	163376.	953523.	292134.	0.	0.	147480.	-15773.
13	2002.	1943490.	1465259.	145157.	-80476.	68486.	48534.	157362.	1058605.	321348.	0.	0.	143798.	-88800.
14	2003.	2094765.	1465259.	158345.	-141776.	68486.	46655.	151347.	1176352.	353483.	0.	0.	140116.	-169270.
15	2004.	2262361.	1465259.	171525.	-211031.	68486.	44778.	145333.	1306955.	388831.	0.	0.	136433.	-257994.
16	2005.	2447865.	1465259.	184724.	-289263.	68486.	42900.	139318.	1451972.	427714.	0.	0.	132751.	-355868.
17	2006.	2653419.	1465259.	197913.	-377239.	68486.	41022.	133303.	1613141.	470486.	0.	0.	129068.	-463896.
18	2007.	2862557.	1465259.	192519.	-485663.	68486.	39144.	127289.	1792159.	517534.	0.	0.	125386.	-573731.
19	2008.	3096275.	1465259.	187123.	-606140.	68486.	37266.	121274.	1991133.	569287.	0.	0.	121703.	-696049.
20	2009.	3357251.	1465259.	181731.	-746038.	68486.	35388.	115260.	2212149.	626216.	0.	0.	118021.	-832203.
21	2010.	3648452.	1465259.	176337.	-888819.	68486.	33510.	109245.	2457698.	688838.	0.	0.	114338.	-983699.
22	2011.	3973171.	1465259.	170943.	-1054165.	68486.	31632.	103230.	2730502.	757722.	0.	0.	110656.	-1152209.
23	2012.	4335066.	1465259.	165549.	-1237593.	68486.	29754.	97216.	3033580.	833494.	0.	0.	106973.	-1339506.
24	2013.	4738168.	1465259.	160155.	-1441578.	68486.	27876.	91201.	3370316.	916343.	0.	0.	103291.	-1547886.
25	2014.	5186989.	1465259.	154761.	-1667971.	68486.	25998.	85187.	3744421.	1008527.	0.	0.	99608.	-1779390.
26	2015.	5686503.	1465259.	149357.	-1919326.	68486.	24120.	79172.	4160052.	1109380.	0.	0.	95926.	-2036626.
27	2016.	6242238.	1465259.	143973.	-2198363.	68486.	22242.	73157.	4621618.	1220318.	0.	0.	92243.	-2322399.
28	2017.	6860322.	1465259.	138579.	-2506102.	68486.	20364.	67143.	5134840.	1342350.	0.	0.	69561.	-2639821.
29	2018.	7547555.	1465259.	133185.	-2851890.	68486.	18486.	61128.	5704807.	1476585.	0.	0.	84878.	-2992343.
30	2019.	8311478.	1465259.	127791.	-3233439.	68486.	16608.	55113.	6338040.	1624244.	0.	340612.	421808.	-3043181.

PRESENT VALUE AT BEGINNING OF 1990 OF CASH FLOWS TO COMMON EQUITY DISCOUNTED AT 16.00 PCT./YEAR:

WITH REVENUE AT LEVELIZED PRICE = $ 1031076250.
WITH REVENUE AT PRICES NOT LEVELIZED = $ 1031076250.

COMMON EQUITY OUTSTANDING AT BEGINNING OF 1990 = $ 1031076250.

* ONLY PRINCIPAL PRODUCT PRICE IS LEVELIZED, USING RETURN ON EQUITY OF 16.00 PCT./YEAR
\+ RECOVERY AND DIVIDENDS

1106 MW TEXACO GCC PLANT -- ILLINOIS LOCATION -- IOU -- CASE A2 -- 7/20/81.
BASE CASE -- 51 PCT CF. -- 1% PCT INFL. -- 1990 STARTUP.

TABLE D3

REVENUE REQUIREMENTS SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

YR	CALENDAR YEAR	RETURN ON COMMON EQUITY	PREFERRED STOCK DIVIDENDS	INTEREST ON DEBT	INCOME TAXES	OTHER TAXES AND INSURANCE	RECOVERY OF CAPITAL: BOOK DEPRECIATION	RECOVERY OF CAPITAL: OTHER	FUEL/RAW MATERIAL COST	OPERATING AND MAINTENANCE COSTS	TOTAL REVENUE REQUIRED	REVENUE FROM BY-PRODUCTS	REVENUE FROM PRINCIPAL PRODUCT: TOTAL	MILLS PER KWH	MILLS PER KWH EXPRESSED IN MID-1980 DOLLARS
1	1990.	164972.	56341.	180438.	-13113.	68486.	66844.	0.	299397.	102591.	945757.	0.	945757.	108.41	39.85
2	1991.	161296.	54463.	174424.	76.	68486.	66844.	0.	332630.	112630.	990843.	0.	990843.	113.58	37.96
3	1992.	157667.	52555.	168409.	13265.	68486.	66844.	0.	369552.	122894.	1040642.	0.	1040642.	119.29	36.24
4	1993.	153925.	50787.	162395.	25454.	68486.	66844.	0.	418572.	136283.	1095665.	0.	1095666.	125.69	34.69
5	1994.	150292.	48629.	156380.	39644.	68486.	66844.	0.	456146.	149911.	1156492.	0.	1156482.	132.57	33.28
6	1995.	146560.	46551.	150365.	52533.	68486.	66844.	0.	506778.	164902.	1223719.	0.	1223719.	140.28	32.02
7	1996.	142877.	45073.	144351.	65622.	68486.	66844.	0.	563930.	181393.	1298076.	0.	1298076.	146.60	30.88
8	1997.	139195.	43195.	138336.	79211.	68486.	66844.	0.	625526.	199532.	1380325.	0.	1380325.	158.23	29.85
9	1998.	135512.	41317.	132321.	92466.	68486.	66844.	0.	694960.	219465.	1471326.	0.	1471326.	166.66	28.92
10	1999.	131630.	39439.	126307.	105509.	68486.	66844.	0.	772166.	241433.	1572029.	0.	1572029.	180.25	28.09
11	2000.	128147.	37561.	120292.	115776.	68486.	66844.	0.	857804.	265577.	1683439.	0.	1683489.	192.98	27.35
12	2001.	124465.	35683.	114278.	131967.	68486.	66844.	0.	953020.	292134.	1806877.	0.	1805377.	207.12	26.69
13	2002.	120783.	33805.	108263.	145157.	68486.	66844.	0.	1058605.	321348.	1943490.	0.	1943490.	222.78	26.09
14	2003.	117100.	31927.	102246.	150346.	68486.	66844.	0.	1176332.	353463.	2094766.	0.	2094766.	240.13	25.57
15	2004.	113418.	30049.	96234.	171535.	68486.	66844.	0.	1306965.	368831.	2262301.	0.	2262301.	259.33	25.13
16	2005.	109735.	28170.	90219.	184724.	68486.	66844.	0.	1451972.	427714.	2447855.	0.	2447365.	280.66	24.69
17	2006.	106053.	26292.	84205.	197913.	68486.	66844.	0.	1613141.	473436.	2653419.	0.	2653419.	304.16	24.33
18	2007.	102370.	24414.	78190.	192519.	68486.	66844.	0.	1792193.	517534.	2862557.	0.	2862557.	328.14	23.86
19	2008.	98698.	22536.	72175.	187125.	68486.	66844.	0.	1991133.	569287.	3096275.	0.	3096275.	354.93	23.47
20	2009.	95005.	20658.	66161.	181731.	68486.	66844.	0.	2212149.	626216.	3357251.	0.	3357251.	384.84	23.13
21	2010.	91323.	18780.	60146.	176337.	68486.	66844.	0.	2457698.	688828.	3648452.	0.	3648452.	418.23	22.85
22	2011.	87640.	16902.	54132.	170943.	68486.	66844.	0.	2730592.	757722.	3973171.	0.	3973171.	455.45	22.62
23	2012.	83958.	15024.	48117.	165549.	68486.	66844.	0.	3033588.	833494.	4335060.	0.	4335060.	496.93	22.44
24	2013.	80275.	13146.	42102.	160155.	68486.	66844.	0.	3373316.	916843.	4738168.	0.	4738168.	543.14	22.30
25	2014.	76593.	11268.	36088.	154761.	68486.	66844.	0.	3744421.	1008527.	5186989.	0.	5186989.	594.59	22.19
26	2015.	72910.	9390.	30073.	149367.	68486.	66844.	0.	4160052.	1109380.	5686503.	0.	5686503.	651.85	22.12
27	2016.	69228.	7512.	24058.	143973.	68486.	66844.	0.	4621619.	1220318.	6242239.	0.	6242238.	715.55	22.37
28	2017.	65545.	5634.	18044.	138579.	68486.	66844.	0.	5134840.	1342350.	6860322.	0.	6860322.	796.41	22.05
29	2018.	61863.	3756.	12029.	133185.	68486.	66844.	0.	5754617.	1476585.	7547555.	0.	7547555.	865.18	22.05
30	2019.	58180.	1878.	6015.	127791.	68486.	66844.	0.	6338840.	1624244.	8311479.	0.	8311476.	952.75	22.08
											LEVELIZED			170.26 *	29.11 **

LEVELIZED FIXED CHARGE RATE IN CURRENT DOLLARS = .185675
* LEVELIZED USING RETURN ON EQUITY OF 16.2[illegible] PCT./YEAR
** LEVELIZED USING RETURN ON EQUITY OF 5.455 PCT./YEAR

REVENUE REQD FOR METHANOL COPRODUCED WITH ELECTRICITY -- CASE B2 -- 7/21/81
BASE CASE. -- 35 PCT CF. -- 10 PCT INFL. -- 1990 STARTUP.

TABLE D5

CAPITAL OUTLAY SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

DESIGN/ CONSTRUCTION PERIOD (YEAR)	CALENDAR YEAR	PLANT FACILITIES INVESTMENT: IN MID-1981 DOLLARS	PLANT FACILITIES INVESTMENT: AMOUNT OF ESCALATION	PLANT FACILITIES INVESTMENT: ESCALATED INVESTMENT	ALLOWANCE FOR FUNDS DURING CONSTRUCTION: EQUITY	ALLOWANCE FOR FUNDS DURING CONSTRUCTION: INTEREST	OTHER OUTLAYS*	TOTAL OUTLAY	GRANTS IN AID OF CONSTRUCTION	INVESTMENT TAX CREDITS	OTHER INCOME TAX OFFSETS	NET OUTLAY FOR PLANT
1.	1986.	209835.	161901.	371736.	--	--	2307.	374043.	0.	0.	0.	374043.
2.	1987.	349725.	331790.	681515.	--	--	0.	681515.	0.	0.	0.	681515.
3.	1988.	489615.	559918.	1049533.	--	--	0.	1049533.	0.	0.	0.	1049533.
4.	1989.	349725.	474908.	824633.	--	--	184509.	1009143.	0.	0.	0.	1009143.
	TOTALS	1398900.	1525517.	2927417.	0.	668514.	186817.	3114234.	0.	287443.**	0.	3114234.

GROSS DEPRECIABLE INVESTMENT	=	3712185.
NET NON-DEPRECIABLE PLANT OUTLAY	=	70563.
TOTAL NON-DEPRECIABLE INVESTMENT	=	70563.
TOTAL INVESTMENT	=	3782748.

* PREPAID ROYALTIES, LAND, ORGANIZATION AND STARTUP EXPENSES, AND WORKING CAPITAL

** TO BE NORMALIZED OVER PERIOD OF COMMERCIAL OPERATION

GROSS DEPRECIABLE INVESTMENT = ESCALATED PLANT FACILITIES INVESTMENT LESS GRANTS-IN-AID OF CONSTRUCTION PLUS ALLOWANCE FOR FUNDS DURING CONSTRUCTION PLUS PRE-PAID ROYALTIES PLUS ORGANIZATION AND STARTUP EXPENSES

PLANT FINANCING:

COMMON EQUITY	= (.350)X(3782748.)	=	1323962.
PREFERRED STOCK	= (.150)X(3782748.)	=	567412.
DEBT	= (.500)X(3782748.)	=	1891374.
			3782748.

REVENUE REQD FOR METHANOL COPRODUCED WITH ELECTRICITY -- CASE 82 -- 7/29/81
BASE CASE. -- 95 PCT CF. -- 10 PCT INFL. -- 1990 STARTUP.

TABLE D6

CAPITAL RECOVERY SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

PERIOD OF COMMERCIAL OPERATION (YEAR)	CALENDAR YEAR	DEBT BALANCE (BEGINNING OF YR.)	DEBT PRINCIPAL PAYMENT	PREFERRED STOCK BALANCE (BEGINNING OF YEAR)	RECOVERY OF PREFERRED	COMMON EQUITY OUTSTANDING (BEGINNING OF YEAR)	ANNUAL RECOVERY OF COMMON EQUITY: THROUGH BOOK DEPRECIATION	ANNUAL RECOVERY OF COMMON EQUITY: OTHER
1.	1990.	1891374.	63046.	567412.	18914.	1323962.	29406.	0.
2.	1991.	1828328.	63046.	548498.	18914.	1294555.	29406.	0.
3.	1992.	1765282.	63046.	529585.	18914.	1265150.	29406.	0.
4.	1993.	1702237.	63046.	510671.	18914.	1235744.	29406.	0.
5.	1994.	1639191.	63046.	491757.	18914.	1206338.	29406.	0.
6.	1995.	1576145.	63046.	472843.	18914.	1176932.	29406.	0.
7.	1996.	1513099.	63046.	453929.	18914.	1147526.	29406.	0.
8.	1997.	1450053.	63046.	435016.	18914.	1118120.	29406.	0.
9.	1998.	1387008.	63046.	416102.	18914.	1088714.	29406.	0.
10.	1999.	1323962.	63046.	397189.	18914.	1059308.	29406.	0.
11.	2000.	1260916.	63046.	378275.	18914.	1029902.	29406.	0.
12.	2001.	1197870.	63046.	359361.	18914.	1000496.	29406.	0.
13.	2002.	1134824.	63046.	340447.	18914.	971090.	29406.	0.
14.	2003.	1071779.	63046.	321534.	18914.	941684.	29406.	0.
15.	2004.	1008733.	63046.	302620.	18914.	912279.	29406.	0.
16.	2005.	945687.	63046.	283706.	18914.	882372.	29406.	0.
17.	2006.	882641.	63046.	264792.	18914.	853465.	29406.	0.
18.	2007.	819595.	63046.	245879.	18914.	824060.	29406.	0.
19.	2008.	756550.	63046.	226965.	18914.	794554.	29406.	0.
20.	2009.	693504.	63046.	208051.	18914.	765245.	29406.	0.
21.	2010.	630458.	63046.	189137.	18914.	735342.	29406.	0.
22.	2011.	567412.	63046.	170224.	18914.	706436.	29406.	0.
23.	2012.	504366.	63046.	151310.	18914.	677130.	29406.	0.
24.	2013.	441321.	63046.	132396.	18914.	647524.	29406.	0.
25.	2014.	378275.	63046.	113482.	18914.	618218.	29406.	0.
26.	2015.	315229.	63046.	94569.	18914.	588312.	29406.	0.
27.	2016.	252183.	63046.	75655.	18914.	559406.	29406.	0.
28.	2017.	189137.	63046.	56741.	18914.	530000.	29406.	0.
29.	2018.	126092.	63046.	37827.	18914.	500594.	29406.	0.
30.	2019.	63046.	63046.	18914.	18914.	471188.	29406.	0.
31.	2020.	0.	0.	0.	0.	441782.	0.	0.

REVENUE REQD FOR METHANOL COPRODUCED WITH ELECTRICITY -- CASE B2 -- 7/20/81
BASE CASE. -- 90 PCT CF. -- 10 PCT INFL. -- 1990 STARTUP.

TABLE D7

REVENUE REQUIREMENTS SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

YR	CALENDAR YEAR	RETURN ON COMMON EQUITY	PREFERRED STOCK DIVIDENDS	INTEREST ON DEBT	INCOME TAXES	OTHER TAXES AND INSURANCE	RECOVERY OF CAPITAL: BOOK DEPRECIATION	RECOVERY OF CAPITAL: OTHER	FUEL/RAW MATERIAL COST	OPERATING AND MAINTENANCE COSTS	TOTAL REVENUE REQUIRED	REVENUE FROM ELECTRIC POWER	REVENUE FROM COPRODUCT METHANOL: TOTAL	$ PER MMBTU	$ PER MMBTU-MEOH EXPRESSED IN MID-1980 DOLLARS
1	1990.	211834.	72345.	231693.	-16461.	87823.	111366.	0.	428818.	145346.	1252763.	945757.	307006.	15.18	5.58
2	1991.	207129.	69934.	223970.	466.	87823.	111366.	0.	454196.	159880.	1314763.	990843.	323920.	16.02	5.35
3	1992.	202424.	67522.	216247.	17393.	87823.	111366.	0.	504612.	175869.	1383254.	1040642.	342612.	16.95	5.15
4	1993.	197719.	65111.	238524.	34320.	87823.	111366.	0.	560624.	193455.	1458941.	1095665.	363275.	17.97	4.96
5	1994.	193014.	62699.	200801.	51247.	87823.	111366.	0.	622853.	212800.	1542603.	1156482.	385121.	19.10	4.80
6	1995.	188309.	60288.	193078.	68174.	87823.	111366.	0.	691996.	234080.	1635107.	1223719.	411388.	20.35	4.64
7	1996.	183604.	57876.	185355.	85101.	87823.	111366.	0.	768801.	257489.	1737413.	1298076.	439337.	21.73	4.51
8	1997.	178899.	55465.	177632.	102028.	87823.	111366.	0.	854139.	283237.	1850536.	1380325.	470261.	23.26	4.39
9	1998.	174194.	53053.	169908.	118955.	87823.	111366.	0.	948947.	311561.	1975807.	1471326.	504481.	24.95	4.28
10	1999.	169489.	50642.	162185.	135882.	87823.	111366.	0.	1054280.	342717.	2114333.	1572029.	542354.	26.83	4.18
11	2000.	164784.	48230.	154462.	152809.	87823.	111366.	0.	1171305.	376989.	2267767.	1683489.	584278.	28.90	4.10
12	2001.	160079.	45819.	146739.	169735.	87823.	111366.	0.	1301320.	414688.	2437569.	1806877.	630692.	31.19	4.02
13	2002.	155374.	43407.	139016.	186662.	87823.	111366.	0.	1445767.	456157.	2625571.	1943490.	682081.	33.74	3.95
14	2003.	150669.	40996.	131293.	203599.	87823.	111366.	0.	1606247.	501772.	2833755.	2094766.	738989.	36.55	3.89
15	2004.	145964.	38584.	123570.	220516.	87823.	111366.	0.	1784540.	551949.	3064313.	2262301.	802012.	39.67	3.84
16	2005.	141259.	36173.	115847.	237443.	87823.	111366.	0.	1982624.	607144.	3319679.	2447865.	871814.	43.12	3.79
17	2006.	136555.	33761.	108124.	254370.	87823.	111366.	0.	2202696.	667859.	3602552.	2653419.	949133.	46.95	3.76
18	2007.	131850.	31350.	100400.	247467.	87823.	111366.	0.	2447195.	734645.	3892094.	2862557.	1029537.	50.92	3.70
19	2008.	127145.	28938.	92677.	240563.	87823.	111366.	0.	2718834.	808109.	4215454.	3096276.	1119178.	55.36	3.66
20	2009.	122440.	26527.	84954.	233660.	87823.	111366.	0.	3025624.	888920.	4576313.	3357251.	1219062.	60.30	3.62
21	2010.	117735.	24115.	77231.	226757.	87823.	111366.	0.	3355913.	977612.	4978751.	3648452.	1330299.	65.80	3.60
22	2011.	113030.	21704.	69508.	219853.	87823.	111366.	0.	3728420.	1075593.	5427295.	3973171.	1454125.	71.92	3.57
23	2012.	108325.	19292.	61785.	212950.	87823.	111366.	0.	4142275.	1183153.	5926957.	4335060.	1591907.	78.74	3.56
24	2013.	103620.	16881.	54062.	206047.	87823.	111366.	0.	4602057.	1301468.	6483332.	4738168.	1745164.	86.32	3.54
25	2014.	98915.	14469.	46339.	199143.	87823.	111366.	0.	5112695.	1431615.	7102565.	5186989.	1915576.	94.75	3.54
26	2015.	94210.	12058.	38616.	192240.	87823.	111366.	0.	5680428.	1574776.	7791515.	5686503.	2105012.	104.12	3.53
27	2016.	89505.	9646.	30892.	185337.	87823.	111366.	0.	6310955.	1732254.	8557777.	6242238.	2315539.	114.53	3.53
28	2017.	84800.	7235.	23169.	178433.	87823.	111366.	0.	7011471.	1905479.	9409775.	6860322.	2549454.	126.10	3.54
29	2018.	80095.	4823.	15446.	171530.	87823.	111366.	0.	7789745.	2096027.	10356854.	7547555.	2809299.	138.95	3.54
30	2019.	75390.	2412.	7723.	164627.	87823.	111366.	0.	8654405.	2305630.	11409375.	8311476.	3097898.	153.23	3.55
											LEVELIZED			* 25.24	** 4.32

LEVELIZED FIXED CHARGE RATE IN CURRENT DOLLARS = .185731

* LEVELIZED USING RETURN ON EQUITY OF 16.000 PCT./YEAR

** LEVELIZED USING RETURN ON EQUITY OF 5.455 PCT./YEAR

REVENUE REQD FOR METHANOL COPRODUCED WITH ELECTRICITY -- CASE B2 -- 7/20/81
BASE CASE. -- 90 PCT CF. -- 10 PCT INFL. -- 1990 STARTUP.

TABLE D8

PROJECT CASH FLOW SCHEDULE
FOR AN
INVESTOR-OWNED UTILITY

(THOUSAND DOLLARS)

YR	CALENDAR YEAR	TOTAL REVENUE AT: PRICES NOT LEVELIZED	TOTAL REVENUE AT: LEVELIZED PRICE *	TAXES ON INCOME WITH REVENUE AT: PRICES NOT LEVELIZED	TAXES ON INCOME WITH REVENUE AT: LEVELIZED PRICE *	OTHER TAXES AND INSURANCE	PREFERRED STOCK COST +	DEBT PRINCIPAL AND INTEREST	FUEL/RAW MATERIAL COST	OPERATING AND MAINTENANCE COSTS	COMMON EQUITY PORTION OF RECURRING INVESTMENT	PLANT SALVAGE VALUE, WORKING CAPITAL, AND LAND	CASH FLOW TO COMMON EQUITY WITH REVENUE AT: PRICES NOT LEVELIZED	CASH FLOW TO COMMON EQUITY WITH REVENUE AT: LEVELIZED PRICE *
1	1990.	1252763.	1456149.	-16461.	83657.	87823.	91259.	294739.	408610.	145346.	0.	0.	241240.	344479.
2	1991.	1314763.	1511235.	455.	92235.	87823.	88847.	287016.	454196.	159390.	0.	0.	236535.	331189.
3	1992.	1383254.	1551034.	17393.	100008.	87823.	86436.	279293.	504612.	175568.	0.	0.	231830.	316995.
4	1993.	1458941.	1615358.	34322.	106751.	87823.	84024.	271570.	560624.	193455.	0.	0.	227125.	301802.
5	1994.	1542603.	1666874.	51247.	112438.	87823.	81613.	263847.	622853.	212300.	0.	0.	222420.	285503.
6	1995.	1635107.	1734111.	68174.	116923.	87823.	79201.	256124.	691990.	234080.	0.	0.	217715.	267970.
7	1996.	1737413.	1805463.	85101.	120098.	87823.	76790.	248400.	768851.	257489.	0.	0.	213010.	249078.
8	1997.	1850566.	1890717.	102028.	121788.	87823.	74378.	240677.	854138.	283237.	0.	0.	208305.	228675.
9	1998.	1975857.	1961718.	118955.	121855.	87823.	71967.	232954.	948947.	311561.	0.	0.	203600.	206601.
10	1999.	2114383.	2082421.	135862.	120143.	87823.	69555.	225231.	1054285.	342717.	0.	0.	198895.	182671.
11	2000.	2267767.	2193881.	152509.	116427.	87823.	67144.	217508.	1171325.	376789.	0.	0.	194190.	156685.
12	2001.	2437569.	2317269.	169735.	110590.	87823.	64732.	209785.	1301320.	414568.	0.	0.	189485.	128421.
13	2002.	2625571.	2453862.	186662.	102123.	87823.	62321.	202062.	1445767.	456157.	0.	0.	184780.	97631.
14	2003.	2833755.	2605158.	203589.	91028.	87823.	59909.	194339.	1606247.	501772.	0.	0.	180075.	64340.
15	2004.	3064313.	2772693.	220516.	76923.	87823.	57498.	186616.	1784540.	551949.	0.	0.	175370.	27344.
16	2005.	3319679.	2959257.	237443.	59479.	87823.	55086.	178892.	1982624.	607144.	0.	0.	170665.	-12792.
17	2006.	3602552.	3163811.	254370.	38334.	87823.	52675.	171169.	2202696.	667359.	0.	0.	165961.	-56744.
18	2007.	3892094.	3372549.	247467.	-8166.	87823.	50263.	163446.	2447195.	734545.	0.	0.	161256.	-102263.
19	2008.	4215454.	3596668.	240563.	-59253.	87823.	47852.	155723.	2718834.	808109.	0.	0.	156551.	-152469.
20	2009.	4576313.	3867643.	233660.	-115259.	87823.	45440.	148000.	3025624.	886920.	0.	0.	151846.	-207875.
21	2010.	4978751.	4159844.	226757.	-176966.	87823.	43029.	140277.	3355913.	977512.	0.	0.	147141.	-269044.
22	2011.	5427296.	4483563.	219853.	-244941.	87823.	40617.	132554.	3728426.	1075593.	0.	0.	142436.	-336603.
23	2012.	5926967.	4345452.	212950.	-319558.	87823.	38206.	124831.	4142275.	1183153.	0.	0.	137731.	-411246.
24	2013.	6483332.	5243560.	206047.	-451555.	87823.	35794.	117108.	4602067.	1301468.	0.	0.	133026.	-493744.
25	2014.	7102565.	5597381.	199143.	-492759.	87823.	33383.	109384.	5112896.	1431515.	0.	0.	128321.	-584951.
26	2015.	7791515.	6195695.	192240.	-592951.	87823.	30971.	101661.	5680428.	1574776.	0.	0.	123616.	-685813.
27	2016.	8557777.	6752630.	185337.	-703518.	87823.	28560.	93938.	6310955.	1732254.	0.	0.	118911.	-797382.
28	2017.	9409776.	7370714.	178433.	-825631.	87823.	26149.	86215.	7011471.	1905479.	0.	0.	114206.	-920822.
29	2018.	10356854.	8057947.	171530.	-961452.	87823.	23737.	78492.	7789745.	2096027.	0.	0.	109501.	-1057424.
30	2019.	11409376.	8321870.	164627.	-1109451.	87823.	21325.	70769.	8654406.	2305530.	0.	441782.	546578.	-766840.

PRESENT VALUE AT BEGINNING OF 1990 OF CASH FLOWS TO COMMON EQUITY DISCOUNTED AT 16.00 PCT./YEAR:
*
WITH REVENUE AT LEVELIZED PRICE = $ 1323961737.
WITH REVENUE AT PRICES NOT LEVELIZED = $ 1323961737.

COMMON EQUITY OUTSTANDING AT BEGINNING OF 1990 = $ 1323961737.

* ONLY PRINCIPAL PRODUCT PRICE IS LEVELIZED, USING RETURN ON EQUITY OF 16.00 PCT./YEAR

\+ RECOVERY AND DIVIDENDS

DEDICATED COAL TO METHANOL FACILITY -- FLUOR DESIGN -- 7/20/1981
NON-UTILITY OWNERSHIP

TABLE E1

CAPITAL OUTLAY SCHEDULE FOR A NON-UTILITY COMPANY

(THOUSAND DOLLARS)

DESIGN/ CONSTRUCTION PERIOD (YEAR)	CALENDAR YEAR	PLANT FACILITIES INVESTMENT: IN MID-1980 DOLLARS	AMOUNT OF ESCALATION	ESCALATED INVESTMENT	ALLOWANCE FOR FUNDS DURING CONSTRUCTION: EQUITY	INTEREST	OTHER OUTLAYS*	TOTAL OUTLAY	GRANTS IN AID OF CONSTRUCTION	INVESTMENT TAX CREDITS	OTHER INCOME TAX OFFSETS	NET OUTLAY FOR PLANT
1.	1985.	131250.	80129.	211379.	--	0.	3071.	214451.	0.	25842.	1457.	192151.
2.	1986.	262500.	202535.	465035.	--	0.	0.	465035.	0.	45852.	3206.	415977.
3.	1987.	393750.	373557.	767307.	--	0.	0.	767307.	0.	75657.	5290.	686361.
4.	1988.	262500.	300192.	562692.	--	0.	0.	562692.	0.	55481.	3879.	503332.
5.	1989.	262500.	356461.	618961.	--	0.	191201.	810162.	0.	61030.	49933.	699149.
	TOTALS	1312500.	1312875.	2625375.	1228418.	0.	194272.	2819647.	0.	258862.	63815.	2496970.

GROSS DEPRECIABLE INVESTMENT	=	2654849.
NET NON-DEPRECIABLE PLANT OUTLAY	=	-107879.**
EQUITY PORTION OF AFDC	=	1228418.
TOTAL NON-DEPRECIABLE INVESTMENT	=	1120539.
TOTAL INVESTMENT	=	3725388.

* PREPAID ROYALTIES, LAND, ORGANIZATION AND STARTUP EXPENSES, AND WORKING CAPITAL

GROSS DEPRECIABLE INVESTMENT = ESCALATED PLANT FACILITIES INVESTMENT LESS GRANTS-IN-AID OF CONSTRUCTION LESS EXPENSABLE PORTION OF ESCALATED PLANT FACILITIES INVESTMENT PLUS PREPAID ROYALTIES

PLANT FINANCING:

COMMON EQUITY	3725388.
	3725388.

**CONSISTS OF:

LAND	=	3071.
WORKING CAPITAL	=	92127.
AFDC INTEREST	=	0.
EXPENSABLE PORTION OF ESCALATED PLANT FACILITIES INVESTMENT	=	36755.
ORGANIZATION AND START-UP EXPENSES	=	92644.
INVESTMENT TAX CREDITS	=	- 258862.
OTHER INCOME TAX OFFSETS	=	- 63815.
TOTAL	=	-127879.

TABLE E2

CAPITAL RECOVERY SCHEDULE FOR A NON-UTILITY COMPANY

(THOUSAND DOLLARS)

PERIOD OF COMMERCIAL OPERATION (YEAR)	CALENDAR YEAR	DEBT BALANCE (BEGINNING OF YR.)	DEBT PRINCIPAL PAYMENT *	PREFERRED STOCK BALANCE (BEGINNING OF YEAR)	RECOVERY OF PREFERRED *	COMMON EQUITY OUTSTANDING (BEGINNING OF YEAR)	ANNUAL RECOVERY OF COMMON EQUITY * THROUGH BOOK DEPRECIATION	ANNUAL RECOVERY OF COMMON EQUITY * OTHER **
1.	1990.	0.	0.	0.	0.	3725389.	117218.	51767.
2.	1991.	0.	0.	0.	0.	3556402.	117218.	51767.
3.	1992.	0.	0.	0.	0.	3387917.	117218.	51767.
4.	1993.	0.	0.	0.	0.	3218432.	117218.	51767.
5.	1994.	0.	0.	0.	0.	3049947.	117218.	51767.
6.	1995.	0.	0.	0.	0.	2880461.	117218.	51767.
7.	1996.	0.	0.	0.	0.	2711975.	117218.	51767.
8.	1997.	0.	0.	0.	0.	2542491.	117218.	51767.
9.	1998.	0.	0.	0.	0.	2373505.	117218.	51767.
10.	1999.	0.	0.	0.	0.	2204521.	117218.	51767.
11.	2000.	0.	0.	0.	0.	2035535.	117218.	51767.
12.	2001.	0.	0.	0.	0.	1866552.	117218.	51767.
13.	2002.	0.	0.	0.	0.	1697565.	117218.	51767.
14.	2003.	0.	0.	0.	0.	1528580.	117218.	51767.
15.	2004.	0.	0.	0.	0.	1359594.	117218.	51767.
16.	2005.	0.	0.	0.	0.	1190609.	117218.	51767.
17.	2006.	0.	0.	0.	0.	1021524.	117218.	51767.
18.	2007.	0.	0.	0.	0.	852539.	117218.	51767.
19.	2008.	0.	0.	0.	0.	683554.	117218.	51767.
20.	2009.	0.	0.	0.	0.	514569.	117218.	51767.
21.	2010.	0.	0.	0.	0.	345583.	0.	0.

* RECOVERED THROUGH BOOK DEPRECIATION WHEN SUFFICIENT BOOK DEPRECIATION IS AVAILABLE. OTHERWISE, RECOVERY IS THROUGH OTHER CHARGES.

** EQUITY PORTION OF NON-DEPRECIABLE INVESTMENT LESS WORKING CAPITAL LESS LAND

DEDICATED COAL TO METHANOL FACILITY -- FLUOR DESIGN -- 7/20/1981
NON-UTILITY OWNERSHIP

TABLE E3

YEAR-BY-YEAR
REVENUE REQUIREMENTS SCHEDULE
FOR A
NON-UTILITY COMPANY
(SEE NOTE)
(THOUSAND DOLLARS)

							RECOVERY OF CAPITAL						REVENUE FROM PRINCIPAL PRODUCT		
YR	CALENDAR YEAR	RETURN ON COMMON EQUITY	PREFERRED STOCK DIVIDENDS	INTEREST ON DEBT	INCOME TAXES	OTHER TAXES AND INSURANCE	BOOK DEPRECIATION	OTHER *	FUEL/RAW MATERIAL COST	OPERATING AND MAINTENANCE COSTS	TOTAL REVENUE REQUIRED	REVENUE FROM BY-PRODUCTS	TOTAL	$ PER MMBTU	$ PER MMBTU EXPRESSED IN MID-1980 DOLLARS
1	1990.	745678.	0.	0.	561611.	78761.	117218.	51767.	447251.	202627.	2204513.	0.	2204513.	31.73	11.66
2	1991.	711290.	0.	0.	554017.	78761.	117219.	51767.	496895.	222889.	2232829.	0.	2232829.	32.14	10.74
3	1992.	677493.	0.	0.	546223.	78761.	117219.	51767.	552051.	245178.	2268682.	0.	2268682.	32.65	9.92
4	1993.	643696.	0.	0.	538428.	78761.	117218.	51767.	613329.	269696.	2312886.	0.	2312886.	33.29	9.19
5	1994.	609899.	0.	0.	530634.	78761.	117218.	51767.	681409.	296666.	2366344.	0.	2366344.	34.06	8.55
6	1995.	576092.	0.	0.	522840.	78761.	117218.	51767.	757045.	326332.	2430056.	0.	2430056.	34.99	7.98
7	1996.	542295.	0.	0.	515046.	78761.	117218.	51767.	841077.	358966.	2505139.	0.	2505139.	36.06	7.48
8	1997.	508498.	0.	0.	507252.	78761.	117218.	51767.	934435.	394862.	2592795.	0.	2592795.	37.32	7.04
9	1998.	474701.	0.	0.	499457.	78761.	117218.	51767.	1038159.	434349.	2694412.	0.	2694412.	38.78	6.65
10	1999.	440904.	0.	0.	491663.	78761.	117219.	51767.	1153394.	477783.	2811491.	0.	2811491.	40.47	6.31
11	2000.	407107.	0.	0.	483869.	78761.	117218.	51767.	1281421.	525562.	2945735.	0.	2945735.	42.40	6.01
12	2001.	373310.	0.	0.	476075.	78761.	117219.	51767.	1423659.	578118.	3099908.	0.	3099908.	44.60	5.75
13	2002.	339513.	0.	0.	468280.	78761.	117219.	51767.	1581695.	635930.	3273154.	0.	3273154.	47.11	5.52
14	2003.	305716.	0.	0.	461486.	78761.	117218.	51767.	1757252.	699523.	3470723.	0.	3470723.	49.95	5.32
15	2004.	271919.	0.	0.	427791.	78761.	117218.	51767.	1952317.	769475.	3669148.	0.	3669148.	52.81	5.11
16	2005.	238122.	0.	0.	394916.	78761.	117218.	51767.	2169013.	846422.	3896220.	0.	3896220.	56.08	4.93
17	2006.	204325.	0.	0.	362131.	78761.	117219.	51767.	2409773.	931065.	4155041.	0.	4155041.	59.80	4.78
18	2007.	170528.	0.	0.	329346.	78761.	117218.	51767.	2677258.	1024171.	4449050.	0.	4449050.	64.04	4.66
19	2008.	136731.	0.	0.	296561.	78761.	117218.	51767.	2974434.	1126589.	4782061.	0.	4782061.	68.83	4.55
20	2009.	102934.	0.	0.	263776.	78761.	117219.	51767.	3304596.	1239247.	5158300.	0.	5159390.	74.24	4.46
												LEVELIZED		** 36.26	*** 7.07

LEVELIZED FIXED CHARGE RATE IN CURRENT DOLLARS = .366404 **
NOTE: PRODUCTS ARE NOT SOLD AT YEAR-BY-YEAR REVENUE REQUIREMENTS. THEY ARE SOLD AT MARKET PRICES. HOWEVER, THESE REVENUES ARE USED TO DEVELOP THE STARTING PRICES SHOWN BELOW. (SEE USER'S MANUAL)

* NON-DEPRECIABLE INVESTMENT LESS WORKING CAPITAL LESS LAND
** LEVELIZED USING RETURN ON EQUITY OF 20.000 PCT./YEAR
*** LEVELIZED USING RETURN ON EQUITY OF 3.091 PCT./YEAR

TABLE E3 (CONTINUED)
REVENUE REQUIREMENTS SCHEDULE
FOR A
NON-UTILITY COMPANY

STARTING PRICES OF PRIMARY PRODUCT AT THE BEGINNING OF 1990, THE FIRST YEAR OF COMMERCIAL OPERATION **

AT GENERAL INFLATION RATE OF 10.00 PCT./YEAR	=$ 19.47 PER MMBTU
AT ESCALATION RATE OF THE PRICE OF COMPETITIVE ALTERNATIVE OF 0.00 PCT./YEAR ***	=$ 36.26 PER MMBTU

INFLATION-INDEPENDENT PRICES OF PRIMARY PRODUCT IN MID - 1980, THE BASE YEAR FOR COST DATA INPUT **

AT GENERAL INFLATION RATE OF 10.00 PCT./YEAR	=$ 7.87 PER MMBTU
AT ESCALATION RATE OF THE PRICE OF COMPETITIVE ALTERNATIVE OF 0.00 PCT./YEAR ***	=$ 36.26 PER MMBTU

** THE PRICE OF THE PRINCIPAL PRODUCT WHICH IF ALLOWED TO INCREASE AT
- THE SPECIFIED RATE OF GENERAL INFLATION, OR
- THE SPECIFIED RATE OF ESCALATION OF THE PRICE OF THE COMPETITIVE ALTERNATIVE
WOULD PROVIDE THE SAME DCF RATE OF RETURN AS EITHER THE CALCULATED YEAR-BY-YEAR PRICES OR THE CALCULATED LEVELIZED PRICES

***INCLUDES 10.00 PCT./YEAR GENERAL INFLATION RATE AND A REAL DECREASE OF 9.09 PCT./YEAR OF THE PRICE OF THE COMPETITIVE ALTERNATIVE

DEDICATED COAL TO METHANOL FACILITY -- FLUOR DESIGN -- 7/20/1981
NON-UTILITY OWNERSHIP

TABLE E4

CASH FLOW SCHEDULE FOR A NON-UTILITY COMPANY
WITH PRINCIPAL PRODUCT SOLD AT ESCALATED REQUIRED STARTING PRICE

(THOUSAND DOLLARS)

YEAR	CALEN-DAR YEAR	REQUIRED PRICE, $ PER MMBTU	REVENUE FROM PRINCIPAL PRODUCT	REVENUE FROM BY-PRODUCTS	TOTAL REVENUE	TAXES ON INCOME	OTHER CASH DISBURSE-MENTS *	COMMON EQUITY PORTION OF RE-CURRING INVEST-MENT	PLANT SALVAGE VALUE, WORKING CAPITAL, AND LAND	CASH FLOW TO COMMON EQUITY
1	1990.	21.41	1487768.	0.	1487768.	298886.	728639.	0.	0.	550243.
2	1991.	23.55	1636545.	0.	1636545.	266467.	796547.	0.	0.	577592.
3	1992.	25.91	1800298.	0.	1800298.	315542.	875991.	0.	0.	608667.
4	1993.	28.50	1980221.	0.	1980221.	374623.	961786.	0.	0.	643810.
5	1994.	31.35	2178242.	0.	2178242.	438012.	1056836.	0.	0.	663394.
6	1995.	34.49	2396066.	0.	2396066.	506103.	1162138.	0.	0.	727824.
7	1996.	37.94	2635672.	0.	2635672.	579325.	1278804.	0.	0.	777544.
8	1997.	41.73	2899239.	0.	2899239.	558145.	1408959.	0.	0.	833035.
9	1998.	45.90	3189163.	0.	3189163.	743973.	1551268.	0.	0.	894822.
10	1999.	50.49	3508080.	0.	3508080.	834663.	1709939.	0.	0.	963478.
11	2000.	55.54	3858898.	0.	3858898.	933520.	1885744.	0.	0.	1039624.
12	2001.	61.10	4244777.	0.	4244777.	1140305.	2083538.	0.	0.	1123938.
13	2002.	67.21	4669254.	0.	4669254.	1155729.	2296376.	0.	0.	1217158.
14	2003.	73.93	5136180.	0.	5136180.	1286557.	2535536.	0.	0.	1320987.
15	2004.	81.32	5649798.	0.	5649798.	1402973.	2800543.	0.	0.	1446282.
16	2005.	89.45	6214777.	0.	6214777.	1536574.	3094197.	0.	0.	1584007.
17	2006.	98.39	6836255.	0.	6836255.	1682361.	3419595.	0.	0.	1734294.
18	2007.	108.23	7519881.	0.	7519881.	1941423.	3780191.	0.	0.	1898267.
19	2008.	119.06	8271869.	0.	8271869.	2014943.	4179783.	0.	0.	2077142.
20	2009.	130.96	9099655.	0.	9099655.	2204205.	4522604.	0.	345683.	2617930.

PRESENT VALUE AT BEGINNING OF 1990 OF CASH FLOWS TO COMMON EQUITY DISCOUNTED AT 23.30 PCT./YEAR = $ 3725387504.

COMMON EQUITY OUTSTANDING AT BEGINNING OF 1990 = $ 3725387504.

* OTHER TAXES AND INSURANCE, PREFERRED STOCK COST, DEBT PRINCIPAL AND INTEREST, FUEL/RAW MATERIAL COST, AND OPERATING AND MAINTENANCE COSTS

*** WARNING ***
SUM OF FUEL ESC. PERIODS IS GREATER THAN THE LIFE OF THE ENTIRE PROJECT

FLESCR PERIODS = 50.0 -0.0 -0.0
PROJ. BK. LIFE = 20.0
YR. OF COMM OP = 1990.0
YR. OF INIT DC = 1985.0

Appendix F

AREA AND UNIT NUMBERING

Each plant consists of a number of facilities or systems called units. The units are grouped into areas having similar purposes. The areas and units are numbered according to the following consistent convention for identification. The table below shows the area and unit numbering system.

Area	Area Description	Unit	Unit Description
10	Feed Systems	10	Coal Handling, Grinding, and Slurry Preparation
		11	Oxidant Feed
20	On-site Systems	20	Gasification and Ash Handling
		21	Gas Cooling and Particulate Removal
		22A	Acid Gas Removal - Fuel Gas
		22B	Acid Gas Removal - Synthesis Gas
		23	Sulfur Recovery
		24	Tail Gas Treating
		25	Zinc Oxide Treatment and Methanol Plant
30	Utility Systems	30	Steam, Condensate and Boiler Feedwater System
		32*	Cooling Water
		33*	Plant and Instrument Air System
		34*	Potable and Utility Water
		35*	Fuel Systems
		36*	Nitrogen System
40	Off-site Facilities	40*	Effluent Water Treating
		41*	Flare System
		42*	Fire Water System
		43*	Buildings
		44*	Railroad Loading and Unloading
		45*	Electrical Distribution
50	Combined-Cycle System	50	Gas Turbine Power Generation
		51	Heat Recovery and Steam Turbine Power Generation

*Costs of these systems are included in the General Facilities section for each of the six estimates of Total Plant Investment

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