Economics of Texaco Gasification–Combined Cycle Systems

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Project 239
Final Report
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Prepared by
Fluor Engineers and Constructors, Inc.
Irvine, California

ELECTRIC POWER RESEARCH INSTITUTE

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Economic Studies of Coal Gasification Combined Cycle Systems for Electric Power Generation

Prepared by

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ABSTRACT

This report presents the results of an economic screening study for air blown Texaco coal gasification coupled with combined cycle power generation. The objective of this study was to identify whether an air blown Texaco gasifier had economic incentives greater than oxygen blown Texaco gasification.

This process arrangement extends the work covered in the Combined Cycle Report (EPRI AF-642), which included the Lurgi dry ash gasifier, the British Gas Corporation Slagger, and three entrained processes offered by Combustion Engineering, Foster Wheeler, and Texaco. All these processes were integrated with combined cycle plants based on advanced gas turbine technology (2,400°F Combustion Outlet) estimated by Westinghouse to be available in the 1981-1985 time period.

The evaluations were based on complete "grass-roots" facilities sized to conform to the present electric utility practice of building units of approximately 1,000 MW capacity.

The conclusion reached in this supplement report is that within the accuracy of the study using the Texaco process, air blown gasification is economically equivalent to oxygen blown gasification.

It is concluded that development emphasis should be placed on power generation, rotating machinery, heat transfer equipment, and further gasification pilot plant experiments to maximize the overall thermal efficiency of air blown Texaco gasification. The air blown case presented here is more speculative than the oxygen case. This is because the Texaco process has not been as well demonstrated for air blown operation as for oxygen blown, and because of some added uncertainty in the design assumptions for some of the high temperature heat transfer equipment. Both processes have the potential for commercialization in the mid to late 1980's.

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EPRI PERSPECTIVE

PROJECT DESCRIPTION

This final report, Economic Study of Air Blown Texaco Gasification Combined Cycle System For Electric Power Generation, is one of a series of economic evaluations under Research Project 239-2 comparing the costs of electric power from gasification-combined cycle power systems based on a variety of different coal gasification technologies. The results of previous economic evaluations of gasification-combined cycle power systems, including the Lurgi dry ash gasifier, the British Gas Corporation Slagger, the two stage entrained Foster Wheeler device, Combustion Engineering's atmospheric pressure entrained gasifier and an oxygen blown Texaco unit have been reported in EPRI Report AF-642, Economic Studies of Coal Gasification Combined Cycle Systems For Electric Power Generation. The present study extends the comparative economic results presented in EPRI AF-642 to include capital requirements and operating costs for a "grass-roots" air blown Texaco coal gasification-combined cycle power plant.

PROJECT OBJECTIVES

The purpose of this study was to evaluate the capital requirements and operating costs associated with electricity production from an air blown Texaco coal gasification-combined cycle power plant to determine whether or not air blowing of the gasifier offered any appreciable economic advantages over the comparable oxygen blown system. To be consistent with previous studies, the design was based on a plant capacity of approximately 1,000 MW, Illinois #6 coal, a Chicago Illinois site, a 2,400 ^{o}F advanced gas turbine, and mid 1976 dollars with no escalation.

CONCLUSIONS AND RECOMMENDATIONS

Within the accuracy of these screening type estimates it can be concluded that there is no significant economic difference between air blown or oxygen blown Texaco based combined cycle power plants. This result is somewhat surprising due to the extremely high cost of oxygen production. One possible explanation for the higher than expected cost for the air blown system could be the conservative design approach adopted by Texaco based on the fact that very little pilot plant data exist for air blown gasification. This study also highlights the necessity for further development emphasis on power generation cycles, high temperature rotating machinery and high temperature heat transfer equipment to maximize the overall thermal efficiency of Texaco based coal gasification-combined cycle power systems.

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SUMMARY

This study was performed as an extension of previous work[1] to identify whether significant economic incentives exist for air blown (new Case EATC) versus oxygen blown (EXTC), water slurry fed Texaco coal gasification processes coupled with gas turbine combined cycle power plants to generate electricity. This case study was performed using advanced gas turbine designs based on a 2400°F combustor outlet temperature. The availability of these turbines has been discussed elsewhere.[2]

The air blown Texaco single stage entrained case is designated EATC. The design is based on feeding the coal in a water slurry. Where comparisons are reported with the previous Case EXTC[3] (oxygen blown Texaco coal gasifier/combined cycle plant) the water slurry case is the basis.

The plant size was selected to match current utility practice of building plants in the 1000-1500 megawatt (MW) capacity range. Both plants feed a constant coal rate equivalent to 10,000 tons/day of Illinois No. 6, and produce in the range of 1100 to 1200 megawatts of power (Table S-1).

The evaluation was done by using data from EPRI and process developers to prepare process designs and cost estimates for integrated grass roots plants. Economic calculations for cost of services were then made. This information is summarized in Table S-2.

The major conclusions of this study are as follows:

> The Texaco coal gasification combined cycle systems have the potential for being available for commercialization in the mid to late 1980's. Both air and oxygen blown systems appear to be more efficient than the current coal fired boiler technology.[4]

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

TABLE S-1

SUMMARY OF OPERATING RESULTS

	TEXACO EXTC	TEXACO EATC
Gasification and Gas Cleaning System		
Coal Feed Rate, Lb/Hr m.f.	798,333	798,333
Oxygen or Air/Coal Ratio Lb/Lb m.f.(1)	0.858	1.081
Oxidant Temperature, °F	300	1,000
Steam/Coal Ratio Lb/Lb m.f.(4)	0	0.0086
Slurry Water/Coal Ratio Lb/Lb m.f. (5)	0.503	0.522
Gasification Section Average Pressure, psig	600	600
Crude Gas Temperature, °F	2,300-2,600	2,300-2,600
Crude Gas HHV (Dry Basis), Btu/SCF (2)	281.1	102.5
Temperature of Fuel Gas to Gas Turbine, °F	781	821
Power System		
Gas Turbine Inlet Temperature, °F	2,400	2,400
Pressure Ratio	17:1	17:1
Gas Turbine Exhaust Temperature, °F	1,140	1,135
Steam Conditions, psig/°F/°F	1450/900/1000	1450/900/1000
Condensing Pressure, Inches Hg abs.	2.5	2.5
Stack Temperature, °F	272	272
Gas Turbine Power, MW (3)	745	548
Steam Turbine Power, MW (3)	448	634
Power Consumed, MW	36	43.5
Net System Power, MW	1,157	1,138
Overall System		
Process and Deaerater Makeup Water, gpm/1000 MW	362	289
Cooling Tower Makeup Water, gpm/1000 MW	7,588	9,163
Cooling Water Circulation Rate, gpm/MW	347	415
Cooling Tower Heat Rejection, % of Coal HHV (6)	38.7	47.2
Air Cooler Heat Rejection, % of Coal HHV	5.2	0.8
Net Heat Rate, Btu/kWh	8,813	8,958
Overall System Efficiency (Coal→Power), % of Coal HHV	38.7	38.1

NOTES:
(1) Dry basis, 100% O_2 for oxygen blown case
(2) Excluding the HHV of H_2S, COS and NH_3
(3) At generator terminals
(4) Includes moisture in oxidant air
(5) Small changes in this ratio do not significantly alter the results presented here.
(6) See Table CS-1 for details.

TABLE S-2

SUMMARY OF ECONOMIC RESULTS

PRODUCTION AT DESIGN CAPACITY (1)	TEXACO EXTC	TEXACO EATC
Net Power, MW	1,157	1,138
Overall Plant Heat Rate, Btu/kWh	8,813	8,958
TOTAL CAPITAL (2)		
Total Capital @ $1/MM Btu Coal, $/kW	816	826
Total Capital @ $2/MM Btu Coal, $/kW	831	841
AVERAGE COSTS OF SERVICES (2)		
Annual Cost @ $1/MM Btu Coal, $1000/yr	262,088	264,078
Per Unit @ $1/MM Btu Coal, mills/kWh	37.21	37.84
Annual Cost @ $2/MM Btu Coal, $1000/yr	327,280	329,271
Per Unit @ $2/MM Btu Coal, mills/kWh	46.47	47.18

NOTES: (1) At 100% operating factor
(2) Mid-1976 dollars and 70% operating factor

- There appears no reason to favor oxygen blown over air blown Texaco gasification for combined cycle power production. In fact, as experimental plant data on air blown coal gasification is extended and exchanger equipment developed to superheat process steam to higher conditions with gasifier effluent, the reverse may be true.

- Plant investment and overall cost of service are essentially identical within the accuracy of the methods used in this study. However, the air blown process has the greater potential for cost reductions resulting from further development work on the gasifiers and associated heat exchangers.

- The cost of the combined cycle equipment is less than half of the investment for both cases. Money spent to reduce the cost of this equipment will benefit both cases.

The technical criteria used in preparing the plant designs are given in the criteria section of this report. Briefly, these criteria are:

- Use data provided by process developers.
- Produce no net products except electricity and sulfur.
- Meet environmental restrictions for an Illinois plant location (1.2 lb SO_2/MM Btu of coal fired).
- Provide all facilities required to permit stand-alone operation of a grass roots plant.

The economic criteria used for capital costs and costs of services estimates are also detailed in the Criteria section of this report. They are summarized as follows:

- Mid-1976 dollars with no escalation.
- Thirty-six month construction period.
- Eight percent construction loan interest, compounded over the plant construction schedule.
- Coal cost of $1.00/MM Btu and $2.00/MM Btu.
- Seventy percent operating load factor.
- Twenty-five year plant life.
- Fifty:fifty debt:equity ratio.

- Eight percent annual bond interest.
- Twelve percent annual return on equity after taxes.

Total capital requirements for each system were determined by adding capital related charges such as preproduction costs, paid-up royalties, initial chemical and catalyst costs, construction loan interest and working capital to the estimated plant investments.

Plant investments include a contingency which is divided into two parts. First is a 15 percent project contingency which is intended to cover estimating uncertainty, and additional equipment that could result from a detailed design of a definitive project at an actual site. The second is a process contingency which is applied to unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. Historically, as a new technology develops from the conceptual stage to commercial reality, a variety of technical problems which were not considered during the early stages of the development emerge. Solution of these problems generally results in an increase in the cost of the technology due to the need for more expensive materials of construction, more complex equipment specifications and sometimes the need for additional processing equipment. A total plant process contingency is arrived at by applying a separate contingency to individual process units based on their state of development and accumulating the results.

The Texaco gasification process has been commercially used with petroleum residues for many years in synthesis gas applications. The process has been demonstrated on a pilot scale with coal by Texaco Development Corp. A demonstration scale plant using coal and oxygen has been built for synthesis gas production in West Germany for RuhrChemie. This unit is scheduled for a first quarter 1978 startup. Another demonstration scale oxygen blown unit, for ammonia synthesis gas, has recently been announced for TVA. Thus, while the Texaco process has been commercially demonstrated on petroleum, it has not yet been demonstrated on a large scale with coal, nor has it yet been demonstrated in combination with a combined cycle power plant. No air blown demonstration plants have been announced.

High temperature heat exchanges coupled to the gasifier effluent line have been developed by two West German firms, Steinmuller and Siegener. Several of these units have seen extended commercial service. Gasification in general would profit greatly by further developments in this type of equipment. No superheater exchangers have yet been built, so far as Fluor can determine, and the design used in this study is entirely conceptual at this point. Should added development work indicate that the design or operation of such superheating equipment is not practical, it would be likely to adversely impact the overall thermal efficiency of the air blown plant, or the cost of services, or both.

Based on present favorable pilot data, considering the simplicity of the gasifier and its feed system, it is estimated that extension to both the above new areas should be relatively simple.

In recognition of the need to determine the operational behavior of a total integrated system, EPRI is sponsoring a study (RP-913) to develop a dynamic simulation model of a Texaco/combined cycle plant.

INTRODUCTION AND SCOPE

The study reported here represents a continuation and extension of earlier economic studies[1] done for EPRI by Fluor. The object of the new work, Case EATC, was to investigate the economics associated with producing electricity from coal in air blown Texaco gasification-combined cycle power plants versus the previously investigated oxygen blown Texaco gasifier, Case EXTC. These plants are based on the Texaco gasification process integrated with advanced gas turbine (2400°F combustor outlet) combined cycle power plants.

Design for the air blown Texaco gasification units, and for the Selexol® acid gas removal unit, were based on information provided by appropriate licensors. The power systems were calculated by Fluor based on similar data[2] from Westinghouse. Plant costs were estimated by Fluor. Economic evaluation criteria were supplied by EPRI.

A block flow diagram and flow sheets are provided for individual process units within the plant where necessary to depict what is included that is specific to this case.

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

CRITERIA

TECHNICAL CRITERIA

Plant designs were based on criteria established by the Electric Power Research Institute (EPRI). These criteria included coal data, site location, gasifier material and heat balances, gasifier equipment requirements and general plant requirements.

For this supplement, data on gasification system operation was provided by Texaco Development Corporation.

Power system performance was estimated by Fluor based on information developed during the preparation of the Combined Cycle Report[1].

The coal analysis is given in Table C-1. Coal was assumed delivered to the site washed and sized.

The site for the plant is the Chicago area; Table C-2 shows pertinent conditions for the site. Raw water makeup in the plant is assumed to be Chicago city water. The Chicago Department of Public Works provided an analysis of finished water from the South District filtration plant, Table C-3. This data was extracted from EPRI report AF-244[2].

In all cases, net plant products were restricted to electricity, sulfur and ammonia. No hydrocarbon by-products were allowed. Plant sulfur emissions were restricted to 1 lb SO_2/MM Btu (HHV) of coal fired.

Fuel, steam and electric power are assumed to be available to the plant at the necessary conditions for start-up and emergency situations. Because the plant is a grass roots installation, it will be self-supporting. In addition to the process and utilities described in this report, the following facilities are provided and included in the cost estimate for each case:

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

2. "Economics of Current and Advanced Gasification Processes for Fuel Gas Production," EPRI AF-244, July, 1976.

- Cooling tower
- Plant and instrument air
- Potable and utility water
- Fuel Gas and Nitrogen Systems
- Firewater
- Flares
- Effluent water treating
- Electrical substation and distribution
- Buildings
- Maintenance
- Laboratory
- Rail
- Road

Generally, process equipment is commercially available equipment. Advanced equipment designs are incorporated where:

- the equipment is expected to be commercially available in the near future;
- the equipment is viewed as a logical, economic extension of the present state of the art.

This is particularly true of the gas turbines used here which are based on a 2400°F combustor outlet temperature.

Redundant equipment or systems are provided where failure would jeopardize a substantial fraction of plant capacity. Major high cost equipment is not spared where experience indicates minimal probability of failure or where multiple trains are provided which limit the impact of a failure should it occur. In addition, redundancy is not provided where storage permits bypass of equipment for a sufficient period of time to accomplish reasonable maintenance and repair. The sparing provided is noted in the plant description section for each case, and on the flow diagrams. The degree of redundancy is compatible with a 90 percent onstream factor in the early years of plant life. The plant design depicted here is intended to represent what is possible when the technology is fully established, and not to necessarily reflect the approach to be taken on a "first of a kind" plant.

TABLE C-1

COAL ANALYSIS

Type	Illinois No. 6
PROXIMATE ANALYSIS (Wt %)	
Moisture	4.2
Ash	9.6
Fixed Carbon	52.0
Volatile Matter	34.2
	100.0
ULTIMATE ANALYSIS - DAF COAL (Wt %)	
Carbon	77.26
Hydrogen	5.92
Oxygen	11.14
Nitrogen	1.39
Sulfur	4.29
Other	-
	100.00
HEATING VALUE - AS RECEIVED	
Higher Heating Value (HHV) (Btu/lb)	12,235
Net Heating Value (LHV) (Btu/lb)	11,709

TABLE C-2

SITE CONDITIONS

LOCATION	Chicago, Illinois
ELEVATION	600 ft
DESIGN AMBIENT PRESSURE	14.4 psia
DESIGN AMBIENT TEMPERATURES	
Summer Dry Bulb	88°F
Summer Wet Bulb	75°F
Winter Dry Bulb	0°F

TABLE C-3

WATER ANALYSIS

Silica (SiO_2)	1.8 ppm
Iron (Fe)	0.09
Manganese (Mn)	0
Calcium (Ca)	39
Magnesium (Mg)	10
Sodium (Na)	3.3
Potassium (K)	0.7
Carbonate (CO_3)	0
Bicarbonate (HCO_3)	132
Sulfate (SO_4)	23
Chloride (Cl)	7.2
Fluoride (F)	0.1
Nitrate (NO_3)	--
Dissolved Solids	168
Hardness as $CaCO_3$	
Total	138
Noncarbonate	30
Color	1 unit
pH	7.9
Turbidity	0
Specific Conductance @ 25°C	275 micromhos

ECONOMIC CRITERIA

A consistent set of criteria for estimating capital requirements and cost of services was supplied by EPRI. Criteria for gasification-combined cycle power plants are summarized in Tables C-4 and C-5. These criteria are the same as those used in the main report.[1]

Operating labor requirements were determined after the plant design was completed and the associated costs computed in accordance with rates shown in Table C-5. Similarly, initial and annual catalyst and chemical requirements and utilities were estimated after designs were completed and costed at expected unit costs.

Plant investment estimates contain a contingency. The contingency has been divided into two parts. First is a project contingency which is intended to cover additional equipment that would result from a more detailed design of a definitive project at an actual site. The second is a process contingency which is applied to unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. Historically, as a new technology develops from the conceptual stage to commercial reality, a variety of technical problems which were not considered during the early stages of the development emerge. Solution of these problems generally results in an increase in the cost of the technology due to the need for more expensive materials of construction, more complex equipment specifications and sometimes the need for additional processing equipment. A total plant process contingency is arrived at by applying a separate contingency to individual process units based on their state of development and accumulating the results. The process contingency allowances, shown as a percentage of the installed plant costs before any project or other process contingencies have been added, are listed in Table C-6.

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

TABLE C-4

CAPITAL INVESTMENT BASIS FOR GASIFICATION - COMBINED CYCLE PLANTS

ITEM	BASIS
Total Plant Investment	- Mid-1976 dollars with no escalation. - Chicago, Illinois location. - Clear and level site.
Total Plant Investment Definition	- The total plant investment is defined as the sum of: (a) Process (or onsite) plant investment costs. (b) General facilities (or offsites) investment costs. (c) Contingencies. These items are discussed below:
Process Plant Investment	- Total constructed cost of all onsite processing units including all direct and indirect construction costs. All sales taxes (5% of total materials) are included.
General Facilities	- The capital cost of the offsite facilities is to be explicitly accounted for. Offsite facilities include roads, buildings, railroad loading and unloading systems, electrical distribution and substations, cooling water systems, inerting systems, effluent water treatment facilities, etc. All sales taxes (5% of total materials) are included.
Project Contingency	- This contingency factor is intended to cover additional equipment that would

ITEM	BASIS
	result from a more detailed design of a definitive project at an actual site. An allowance of 15% of the sum of the Process Plant Investment and the General Facilities cost is used.
Process Contingency	- This contingency factor is to be applied to unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. Process contingency allowances are shown in Table C-8.
Total Capital Requirement	- The total capital requirement includes all capital necessary to complete the entire project. These items include: (a) Total plant investment. (b) Royalties. (c) Preproduction costs. (d) Construction loan interest. (e) Initial chemical and catalyst charge. (f) Working capital.
Paid-up Royalties	- 0.5% of total plant investment.
Preproduction Costs	- One month variable operating costs excluding coal. Variable costs are catalysts and chemicals, utilities, and maintenance materials. - Two month's fixed costs excluding income taxes. Fixed costs are operating and maintenance labor, administrative and support labor, general and administrative expense, and property taxes and insurance.

ITEM	BASIS
	- 5% of total plant investment (this charge allows for possible changes in process equipment, and charges associated with depreciation, bond interest, and return on equity during the preproduction period).
	- 25% of one month's coal at full load.
Construction Loan Interest	- 0.1249x Total Plant Investment (based on compounded 8%/year interest over the plant construction expenditure schedule).
Construction Expenditures	see schedule below

Year	Percent of Total Plant Investment
1	25
2	50
3	25

Expenditures in a given year are assumed uniform over that year.

ITEM	BASIS
Working Capital	- 1.5 months of total operating costs plus 3.5% of total plant investment (this charge allows for accounts receivable).
	- One month's supply of chemicals and catalysts at full plant capacity.
	- One month's supply of coal at full plant capacity.
Land	- Since land costs are site-specific and variable, they have not been included for this study.

TABLE C-5

COST OF SERVICES BASIS FOR GASIFICATION - COMBINED CYCLE PLANTS

ITEM	BASIS
Operating Load Factor	- 70%
Cost of Coal Delivered	- $1.00/MMBtu and $2.00/MMBtu
Chicago City Water	- 40 cents/1,000 gallons
Ash Disposal	- $1.00/ton
By-Product Ammonia Credit	- $100/ton
By-Product Sulfur Credit	- None
Maintenance	- Annual maintenance costs are normally estimated as a percentage of the total installed plant cost of the facilities. The percentage varies widely depending on the nature of the processing conditions and the type of design. Maintenance costs shown below were used.

Process Unit	Maintenance % of Total Plant Investment/Yr
Coal Handling	3.0
Oxidant Feed	2.0
Gasification & Ash Handling	4.5
Gas Cooling	3.0
Acid Gas Removal & Sulfur Recovery	2.0
Fuel Gas Compression	3.0
Process Condensate Treating	3.0
Steam, Condensate & BFW	1.5

ITEM	BASIS
	<table><tr><th>Process Unit</th><th>Maintenance % of Total Plant Investment/Yr</th></tr><tr><td>Support Facilities</td><td>1.5</td></tr><tr><td>Combined Cycle</td><td>1.5</td></tr></table>
Maintenance Labor/Materials Ratio	- 40/60
Operating Labor	- $11 per manhour (this labor rate corresponds to a direct labor charge of $8/hour plus a 35% payroll burden).
Administrative & Support Labor	- 30% of operating and maintenance labor.
General & Administrative Expense	- 60% of operating and maintenance labor.
Property Taxes & Insurance	- 2.5%/yr. of plant investment.
Cost of Capital	- The capital charges (income taxes, interest on debt, return on equity, and depreciation) are computed on a levelized basis with a 10% discount rate. The discount rate is based on the average cost of money. Using this basis, the capital charges will be 15.6% per year of the Total Capital Requirement. The investment factors that form the basis for the 15.6%/yr. capital charge are shown below:

Depreciation	Straight Line
Tax Life	25 years
Plant Life	25 years
Debt/Equity Ratio	50/50
Bond Interest	8% annually
Bond Life	25 years

ITEM	BASIS	
	Return on Equity after Taxes	12% annually
	Income Tax Rate	52%
	Escalation Rate	Not included
	Investment Tax Credit	Not included

The capital charge is based on the Total Capital Requirement with working capital treated the same as depreciable capital.

TABLE C-6

PROCESS CONTINGENCIES

CASE	PERCENT EXTC	EATC
Coal Handling	0	0
Oxidant Feed	0	0
Gasification	15	15
Ash Handling	5	5
Gas Cooling	0-15(1)	0-15(1)
Acid Gas Removal	0	0
Sulfur Recovery (Claus)	0	0
Tail Gas Treating	15	15
Process Condensate Treatment, Steam, Condensate and BFW	0	0
Support Facilities	0	0
Combined Cycle	5	5

(1) 15% applied to the high temperature waste heat boilers with or without superheater, 0% to remaining low temperature gas cooling equipment.

DISCUSSION OF RESULTS

The evaluation presented here should be considered a screening type evaluation. Fluor did not attempt a comprehensive analysis of the Texaco technology, but instead based our designs on the gasifier performance information supplied. Within the budget for this study, it was not feasible to fully optimize the design for each of the units. Outside the gasification and combined cycle system, the design was based on currently available equipment, sometimes with some extensions to large sizes. Occasionally this approach points out the need for the development of a new kind of equipment. With this in mind, the reader should guard against assuming that the comparisons given here are complete or final. Under other circumstances, and at other times, it is possible that the conclusions could change.

In performing such evaluations, especially for relatively new or unfamiliar technology, a tendency exists for plant cost estimates to be somewhat optimistic. This is always a hazard where there is not a full and complete mechanical definition of each item in the plant. In an attempt to offset this tendency, we have applied a "process contingency" as well a project contingency to the plant cost estimates. This is discussed in greater detail in the Economic Criteria. The process contingency is unrelated to estimating accuracy, but instead is intended to reflect the degree to which any specific technology is developed. The accuracy of the plant investment estimates is judged to be $\pm$25%.

In choosing between air blown or oxygen blown gasification for electricity generation, the utility plant owner must consider several factors in addition to economics. Based strictly on the results this work and that reported in EPRI AF-642 (January 1978), the oxygen blown case is slightly more economical than the air blown case. In reality, within the accuracy of these studies, the two types of plants must be considered equivalent on an efficiency or economic basis. If they are in fact equivalent, then an air blown plant could well turn out to be preferable on the basis of eliminating the added operational complexities involved with the oxygen plant and its compression system. From a design standpoint, the air blown gasification system has untapped potential for improvement. As machinery and process development proceeds, such improvement seems inevitable.

The cases presented here are summarized in Tables S-1, S-2 (Summary section), and in Tables D-1, D-2 and D-3 (this section).

For engineering studies of this type, much importance is placed on the thermal efficiency of major processing units such as the coal gasifier. In a highly integrated gasification-combined cycle power plant, however, the only quantity of importance is the overall system efficiency in converting coal to electricity. This efficiency depends not only on the individual component efficiencies, but also how these components are linked together. The following table depicts the overall system efficiency (coal to electricity) for the two Texaco cases:

	Oxygen Blown Texaco System EXTC	Air Blown Texaco System EATC
Overall System Efficiency (Coal to Power)	38.7%	38.1%

The air blown Texaco process requires added development before it is ready for commercialization. The Texaco (oxygen blown) process has been commercially used in the United States for gasification of petroleum residue, and extensive pilot work has been done on coal. A 150 ton/day oxygen blown Texaco coal gasification plant has been built in West Germany and is now in preparation for start-up.

The entrained flow Texaco processes, air or oxygen blown, are not radically different economically from the Lurgi or the slagger.[1] They do, however, have the advantage of simplicity. Because of higher operating temperatures, the Texaco gasification process does not generate hydrocarbon by-products. This simplifies plant design and operations, and may make these plants more adaptable to future environmental regulations. This would be particularly true if stricter criteria were enforced regarding exposure to aromatics, phenolics and/or hydrocarbons. These compounds are destroyed in high temperature, single stage entrained processes.

A brief discussion of salient points related to air blown and oxygen blown Texaco cases follows.

1. "Economic Studies of Coal Gasification Combined Cycle System for Electric Power Generation," EPRI AF-642, January 1978, Cases MACW and MXSC.

TABLE D-1

OVERALL PERFORMANCE SUMMARY

CASE	TEXACO EXTC	TEXACO EATC
OPERATING PARAMETERS		
Coal Type	Illinois #6	
Oxidant	Oxygen	Air
Overall Efficiency, %	38.7	38.1
Total Capital Requirements, \$/kW[1]	816	825
Cost of Services Mills/kWh[1]	37.2	37.8
% of Total Power from Gas Turbine	62.4	46.4
Lbs Oxygen/Lb m.f. Coal	0.86	1.081
Lbs Air/Lb m.f. Coal[2]	-	4.64
Total Lbs Steam/Lb m.f. Coal[3]	0.46	0.48
% Coal Carbon Converted to CH_4	0.16	0.38
HEAT LOSS FLOWS, MM BTU/HR		
Total Stack Gas	1817	1395
Power Surface Condenser	2687	4171
Compressor Surface Condenser	1067	-
Compressor Intercoolers	568	70
Gasifier Effluent Cooling	26	112
Ash Heat (Sensible and HHV)	81	50
Process Condensate Treating	-	-
Acid Gas Removal	78	244

1. Based on \$1.00/MM Btu Coal.
2. Dry Basis
3. Excludes moisture in coal and oxidant streams, but includes moisture in char and coal slurries.

TABLE D-2

SUMMARY OF CAPITAL INVESTMENT ESTIMATES - $/KW

PLANT INVESTMENT	TEXACO EXTC	TEXACO EATC
Coal Handling	19.07	19.77
Oxidant Feed	101.48	20.84
Gasification and Ash Handling	20.97	52.85
Gas Cooling	57.91	104.69
Acid Gas Removal and Sulfur Recovery	24.71	37.76
Process Condensate Treating	--	--
Fuel Gas Compression	--	--
Steam, Condensate and BFW	0.71	0.77
Support Facilities	47.72	50.34
Combined Cycle	262.94	244.40
Contingency	102.14	112.21
TOTAL PLANT INVESTMENT	637.65	643.63
ILLINOIS SALES TAX	14.40	14.30
CAPITAL CHARGES		
Preproduction Costs	40.06	40.82
Paid-up Royalty	3.19	3.22
Initial Catalyst and Chemical Charges	0.44	1.33
Construction Loan Interest	79.64	80.39
TOTAL CAPITAL CHARGES	123.33	125.56
WORKING CAPITAL	41.15	41.99
TOTAL CAPITAL REQUIREMENTS	816.53	825.48

NOTES: Mid-1976 dollars
Coal Cost - $1.00/MM Btu

EXTC

This case was intermediate in overall cost of services when compared to the other cases in the main report.[1] A thermal efficiency of approximately 38.7 percent was achieved. This gasification process, despite the presence of some water in the gasifier as slurry water, uses mainly oxygen to achieve gasification. The process is analogous to partial oxidation and as such, it seems a relatively minor extension of Texaco's commercial experience in the partial oxidation of petroleum residue, so that this entrained bed process would be expected to be commercially available at an early date.

The gasifier operates in the 600 psig range. This pressure was selected to reduce to five the number of gasifiers required. This case had the highest requirement for oxygen of any of the oxygen blown cases.[2]

Although this case produced a gasifier effluent that was partially combusted, the effluent was also much hotter (over 2200°F) than in any of the other cases.[3] This sensible heat was almost as useful in the overall plant heat balance as heating value in the product. Thus, with the effective use of both sensible and latent heat, this case had a good overall efficiency. This case had a somewhat higher overall cost of services than the other plants with similar high thermal efficiencies largely because of the capital costs of the large oxidant feed systems and the high temperature heat exchangers in the gasifier effluent cooling system.

EATC

This case was slightly higher in overall cost of service than EXTC, and therefore intermediate in overall cost of service to the other cases in the main report.[4]

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1. "Economic Studies of Coal Gasification Combined Cycle System for Electric Power Generation," EPRI AF-642, January, 1978, Table S-2.

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

3. Ibid.

4. "Economic Studies of Coal Gasification Combined Cycle System for Electric Power Generation," EPRI AF-642, January, 1978, Table S-2.

A thermal efficiency of approximately 38.1 percent was achieved. The air blown case produces a larger fraction of total power from the steam turbine than the oxygen blown case due to the hot nitrogen in the gasifier effluent. All the cycles studied were calculated at high pressure steam conditions of 1450 psig/ 900°F/1000°F. Since the air blown case generates the most steam, this case could potentially benefit more from higher steam conditions, e.g., 2400 psig/1000°F/ 1000°F, which would increase the steam cycle efficiency. If the high temperature process steam generators with superheating to 1000°F can be developed, air blown gasification could have a higher efficiency than reported here.

Air blown gasification requires more oxygen per pound of coal than oxygen blown. In fact, for this case, essentially all the oxygen required was supplied by air since almost no feedwater decomposed. As a result of higher oxygen consumption, the air blown gasifier effluent has less heating value, more sensible heat and a higher concentration of carbon dioxide than EXTC gasifier effluent. The lower chemical heat reduced the gas turbine count from seven in EXTC to six for this case, whereas the extra sensible heat increased the steam generated in the process units, thereby increasing the steam generated power. The higher carbon dioxide content reduced the concentration of hydrogen sulfide in the sulfur removal unit effluent, but did not reduce it enough to alter the conceptual design of the sulfur recovery units.

The gasifier conditions were similar for both cases, operating in the 600 psig range with the effluents in the 2300 to 2600°F range. As stated in the process discussion, the required number of air blown gasifiers for this case is not stated in this report, not only because Texaco considers the information proprietary but also because the design is subject to refinements. Further pilot plant air blowing studies are expected to reduce gasifier costs and the estimated gasification heat losses for Case EATC.

The plant investments were similar in both cases, slightly less for oxygen blowing. The capital saved by not having to install an air separation plant was consumed in additional expenditures in gasification, gas cooling and sulfur removal due to the higher flow rates of nitrogen bearing gas. The cost of the gas cooling system for Case EATC is up because of two factors: the gasifier count has increased, and a new high temperature gas to gas heat exchanger has been introduced. This piece of equipment produces superheated steam by heat exchange with gasifier effluent. The design for this equipment is conceptual at this point. It was judged, for purposes of this study, to be a reasonable extension of heat transfer technology for the mid to late 1980's. It allows efficient use to be made of the large amounts of gasifier effluent heat. The reader is cautioned, however, that Fluor knows of no such actual device in service at this time, and that substantial engineering problems can be foreseen in developing such equipment.

Finally, it is of interest to examine the percentage contribution of each subsection of each plant to the installed plant cost. Such information is shown in Table D-3. These data indicate that the costs of gasification and ash handling contribute between 4 percent and 10 percent to the total installed plant cost. This suggests that the economic incentive for developing new gasifiers is not great (i.e., reducing the cost of the gasification section of these types of systems is not terribly significant). Developing a gasifier, however, that will reduce the need for downstream processing equipment could have major economic significance.

The other important information to be gleaned from Table D-3 is the fact that the combined cycle portion of each plant contributed approximately 50 percent to the total installed plant cost. This suggests very strongly that development work aimed at simplifying the power generating part of the system in order to reduce its cost could have significant impact on the overall system cost. Care must be exercised in any attempts to reduce costs by simplifying the power system as this could lead to a degradation of the thermal efficiency of the system which would probably result in increasing the cost of the system. EPRI is currently funding a number of screening studies -- RP986-2, United Technologies; RP986-3, General Electric; and RP990-3, Westinghouse) -- to investigate techniques for simplifying the power equipment without degrading the overall power output from the plant.

TABLE D-3

PERCENTAGE CONTRIBUTION OF PLANT SUBSECTIONS TO INSTALLED PLANT COSTS

PERCENT

PLANT INVESTMENT	TEXACO EXTC	TEXACO EATC
Coal Handling	3.56	3.72
Oxidant Feed	18.95	3.92
Gasification and Ash Handling	3.92	9.94
Gas Cooling	10.81	19.71
Acid Gas Removal and Sulfur Recovery	4.61	7.10
Waste Water Treating	--	--
Steam, Condensate and BFW	0.14	0.15
Support Facilities	8.91	9.47
Combined Cycle	49.10	45.99
INSTALLED PLANT COST	100.00	100.00

NOTE: Installed Plant Costs do not include contingencies
Mid-1976 Dollars

PLANT DESCRIPTION - CASE EATC

GENERAL

A grass roots plant for electric power generation based on single stage, entrained bed, air blown gasifiers of the Texaco type, integrated with combined cycle generating equipment, is shown schematically on the block flow diagram EATC-1-1. This plant consumes 10,000 ST/day of Illinois No. 6 coal. Coal is fed to the gasifiers as a water slurry containing approximately 65.7 wt. % dry coal. This case is identified as Case EATC.

The main plant consists of oxidant feed, gasification, gas cooling, acid gas removal units and combined cycle power systems. The oxidant feed unit is in three parallel operating trains. The gas cooling and acid gas removal units are in four operating parallel trains. There are six parallel gas turbine, heat recovery steam generator sets and three steam turbines (includes two small turbines which generate one percent of total power).

In addition to the main processing trains, the plant includes necessary offsite, utility and environmental facilities. Coal receiving, storage and conveying is done in a single train to minimize space and operating labor requirements. Hydrogen sulfide removed from gasified coal is processed through sulfur recovery facilities which produce elemental sulfur. Other operating facilities in the plant are raw water treating, steam generation, cooling water, and effluent water treating. Process condensate generated in Case EATC is recycled back to the gasification unit as in Case EXTC. Support facilities to sustain an independent plant operation are provided as well. Table EATC-1 summarizes major equipment sections in the plant. This table shows the number of operating and spare sections.

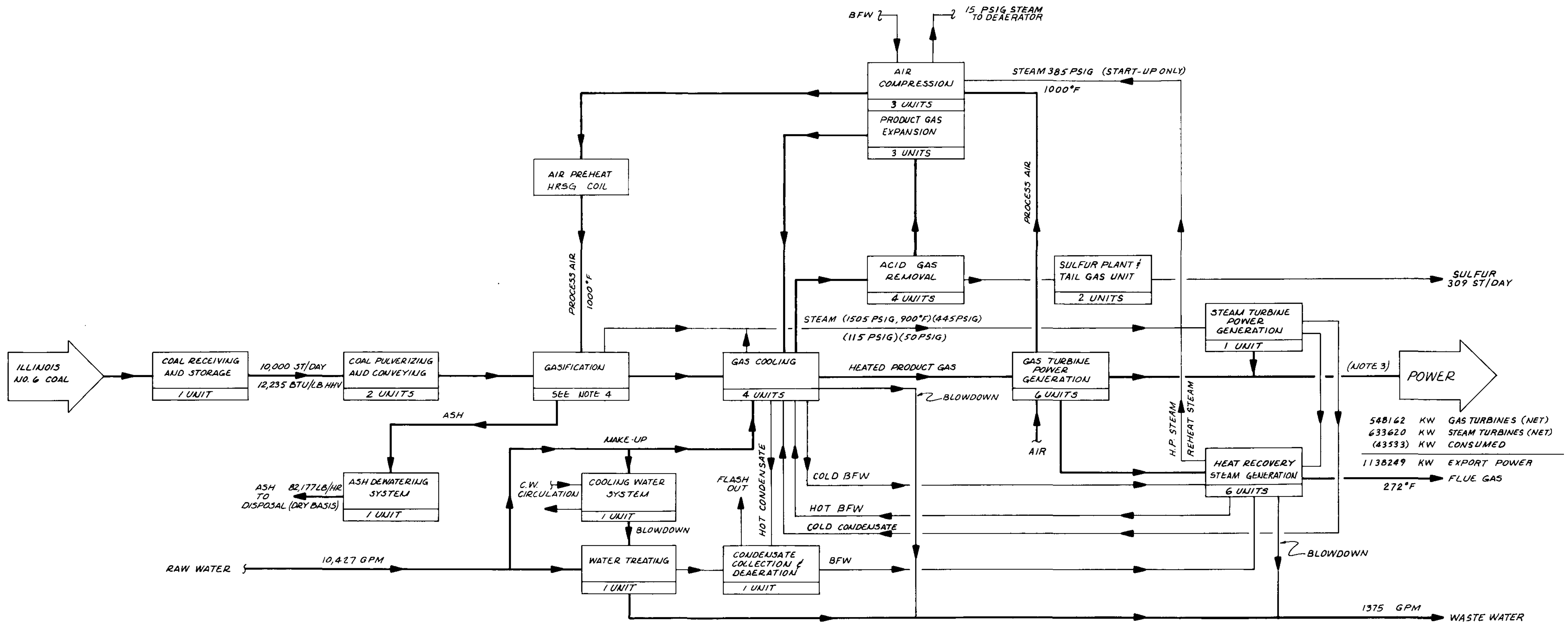
TABLE EATC-1

MAJOR EQUIPMENT SECTIONS: CASES EATC AND EXTC

Unit No.	Unit Name	Case EXTC Operating	Case EXTC Spare	Case EATC Operating	Case EATC Spare
10	Coal Handling	1	0	1	0
11	Oxidant Feed	5	0	3	0
20	Coal Grinding	2	0	2	0
20	Slurry Preparation	1	0	1	0
20	Gasification	5	1	*	0
20	Ash Handling	1	0	1	0
20	Particulate Scrubbing	5	1	*	0
21	Gas Cooling	3	0	4	0
22	Acid Gas Removal	3	0	4	0
23	Sulfur Recovery	2	1	2	1
24	Tail Gas Treating	2	1	2	1
30	Steam, BFW and Condensate System				
	. Condensate Collection and Deaeration	1	0	1	0
	. Water Treating	1	0	1	0
32	Cooling Water System	1	0	1	0
40	Effluent Water Treating	1	0	1	0
50	Gas Turbine/Generator	7	0	6	0
51	Heat Recovery Steam Generator	7	0	6	0
51	Steam Turbine Generator	1	0	3**	0

*The number of gasifiers required is confidential Texaco information and subject to refinement.

**Includes two separate low pressure steam turbine-generators, which contribute 1% of total power.



NOTES:

1. ONLY NUMBER OF OPERATING UNITS ARE SHOWN.
2. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.
3. EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
4. THE NUMBER OF GASIFIER TRAINS IS CONFIDENTIAL TEXACO INFORMATION.

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FLUOR

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PROCESS FLOW DIAGRAM
OVERALL BLOCK FLOW DIAGRAM
TEXACO PROCESS - AIR BLOWN
ELECTRIC POWER RESEARCH INSTITUTE · PALO ALTO, CA.

SCALE: NONE
DRAWING NUMBER: EATC-1-1
REVISION: 3

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COAL PREPARATION

Process Flow Diagram EATC-10-1 depicts the process arrangement of coal preparation equipment in this section for Case EATC. There is one train of coal unloading, stacking, reclaiming and conveying equipment. The equipment in the grinding circuit is arranged in two parallel trains, each capable of processing 300 TPH. The coal slurry preparation is on a one 100%-train basis.

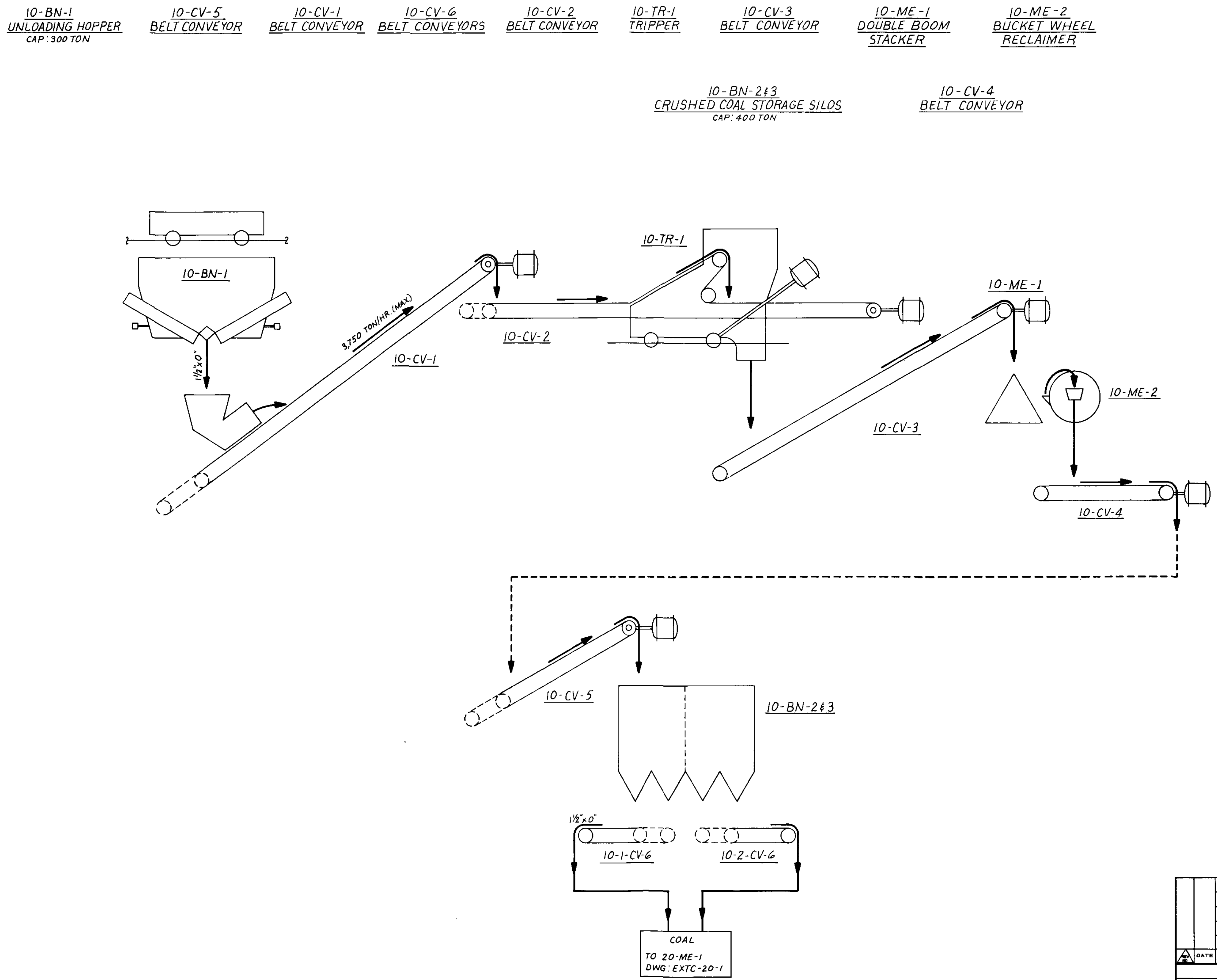
Washed, 1-1/2" x 0 coal is received at the plant site by unit train. The coal is unloaded from 100-ton bottom dump cars into an unloading hopper, withdrawn from the hopper by two vibrating feeders and transported by belt conveyors to a tripper. The tripper distributes coal to a traveling belt stacking system. The stacker travels on tracks and forms storage piles on either side. The unloading and stacking system is designed to handle a three day supply in eight hours.

Coal is reclaimed from storage piles by a bridge type bucket wheel reclaimer rated at 500 tons per hour. This machine is a rail mounted bridge which supports a rotating bucket wheel and belt conveyor. The wheel moves across the face of the pile, making a vertical cut across the many layers of coal. At the end of each cut, the reclaimer moves ahead a predetermined distance and the wheel makes another cut in the opposite direction. The excavated coal is carried by a series of conveyors to the crushed coal storage silos.

Coal is withdrawn from the coal storage silo by two vibrating feeders and transported by a belt conveyor to coal grinding and slurry preparation unit. The coal grinding and slurry preparation area is proprietary to the Texaco process.

Equipment Notes

All the equipment is commercially available.



NOTE

ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A ONE 100% TRAIN BASIS.

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FLUOR

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PROCESS FLOW DIAGRAM
COAL PREPARATION
TEXACO PROCESS-AIR BLOWN

EPRI — PALO ALTO, CALIFORNIA

SCALE: NONE | DRAWING NUMBER: EATC-10-1 | REVISION: 2

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GASIFICATION AND ASH HANDLING

Process Flow Diagram EATC-20-1 shows the gasification step for Case EATC. The required number of gasifiers for this case is not stated in this report because Texaco considers the information proprietary and subject to refinement (see Process Discussion). The ash handling system consists of one 100% capacity train. The boxes on EATC-20-1 represent proprietary sections of the Texaco Coal Gasification Process. Each of these sections contains many units of equipment.

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

Coal slurry and air 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 600 psig and temperatures in the range of 2300°F to 2600°F. Part of the coal burns with air to produce a hot flue gas. This reaction provides heat for the endothermic steam/carbon and carbon/CO_2 reactions. The coal's hydrogen and carbon 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. The ash melts to form slag. The gasification temperature must be sufficiently above the ash flow point to ensure free flowing molten slag. At the high temperatures prevailing in the gasifier, some of the ammonia in the recycled water is destroyed.

Most of the ash in the form of slag falls into a water quench at the bottom of the gasifier. The resultant ash slurry leaves the gasifier and enters the slag dewatering unit. A slag/ash cake from dewatering unit 20-ME-2 is disposed to landfill. Overflow from the slag dewatering unit is recycled to the coal grinding system.

Raw hot gas from the gasifier is cooled in a gas cooling unit, 20-1-ME-3, to a temperature well below the ash softening point. This gas cooling system is of proprietary design and allows for the removal of ash entrained in the crude gas for the protection of downstream heat exchange equipment. High pressure (HP) steam at 1520 psia, saturated intermediate pressure (IP) steam at 445 psig, saturated medium pressure (MP) steam at 115 and low pressure (LP) 50 psig, are produced in Unit 20-1-ME-3. Hot boiler feedwater near HP steam saturation temperature (598°F) and feedwater at 347°F are supplied from steam generation (HRSG) units located in Unit 51. Boiler feedwater is also supplied from deaerator 51-DA-1 located in Unit 51 for LP steam production.

The raw gas leaves 20-1-ME-3 and flows to the gas scrubbing unit, 20-1-ME-4. Ammonia absorber bottoms from the gas cooling area (Flow Diagram: EATC-21-1) and hot process condensate are used for gas scrubbing. Water from 20-1-ME-4 is recycled to 20-ME-1. The clean gases from 20-1-ME-4 flow to the gas cooling section. In subsequent sections of this report dealing with economics, the reader's attention is called to the fact that the cost of equipment included in 20-1-ME-3 and 4 is included in the gas cooling system costs.

Equipment Notes

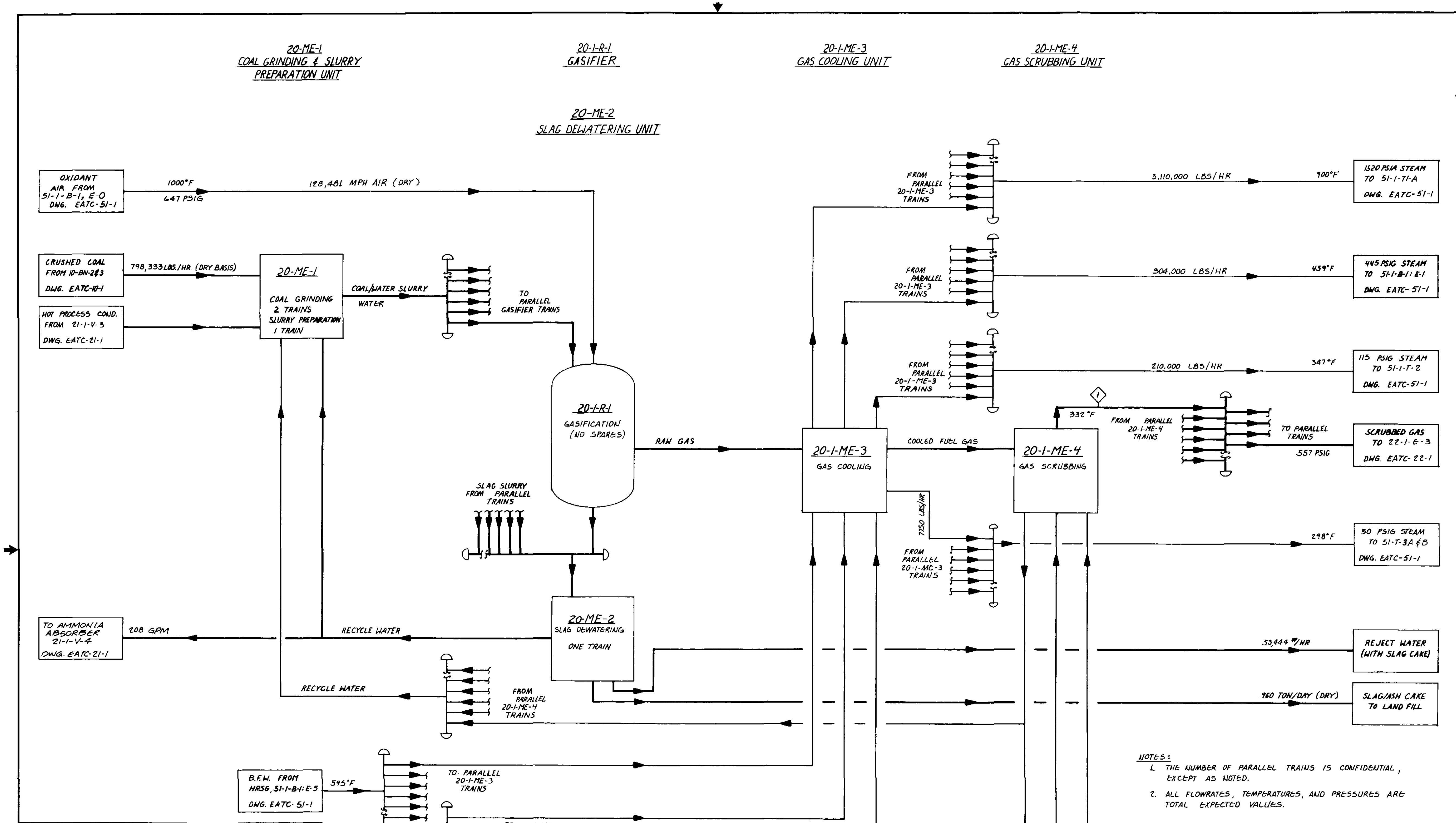
The Texaco gasifier is commercially proven for the gasification of liquid hydrocarbons. Coal gasification is still in the pilot plant stage.

The Texaco coal gasification research facility at Montebello, California, is presently testing coals and chars in a 350 psia 15 ton/day gasifier. A 150 ton/day oxygen blown Texaco coal gasifier is scheduled to start up in Germany early in 1978. It is important to note, however, that the bulk of the pilot plant data refer to operation with oxygen blowing. Very little air blown gasification data exists for the Texaco unit.

The slag dewatering unit is commercially proven.

The gas scrubbing unit equipment is commercially available.

Two key features of this design are a pair of high temperature heat exchangers coupled to each gasifier effluent line. The first is a waste heat boiler for generating 1500 psi saturated steam. This is not an off-the-shelf item; however, successful designs for such units have been developed by Steinmuller and by Siegener, both firms of West Germany. Some of these units have seen extended service. The second exchanger is a superheater for the 1500 psi steam. This unit, so far as Fluor can determine, is entirely conceptual at this point. While we believe the design used for cost estimating is realistic, no examples of such a unit currently exist.



MATERIAL BALANCE (TOTAL)

COMPONENT	PROCESS STREAM NO. ⟨1⟩ NET SCRUBBER EFFLUENT MPH	MOL %	
CH_4	176.3	0.08	
H_2	19678.6	9.30	
CO	32952.0	15.58	
CO_2	13001.6	6.15	
H_2S	883.2	0.42	
COS	77.8	0.04	
N_2	101162.0	47.83	
AR	1213.4	.57	
NH_3	178.1	.09	
H_2O	42176.4	19.94	
TOTAL MPH	211500.0		
LB/HR	5174348.		
MOL. WT.	24.46		

NOTES:

1. THE NUMBER OF PARALLEL TRAINS IS CONFIDENTIAL, EXCEPT AS NOTED.
2. ALL FLOWRATES, TEMPERATURES, AND PRESSURES ARE TOTAL EXPECTED VALUES.

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FLUOR

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PROCESS FLOW DIAGRAM
COAL GASIFICATION/ASH HANDLING
TEXACO PROCESS - AIR BLOWN

EPRI PALO ALTO, CALIFORNIA

SCALE NONE | DRAWING NUMBER EATC-20-1 | REVISION 3

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OXIDANT FEED

Process Flow Diagram EATC-11-1 shows the oxidant feed system for Case EATC. There are three parallel operating trains. Each train has one booster air compressor and associated heat exchangers. No spare train is provided in this section.

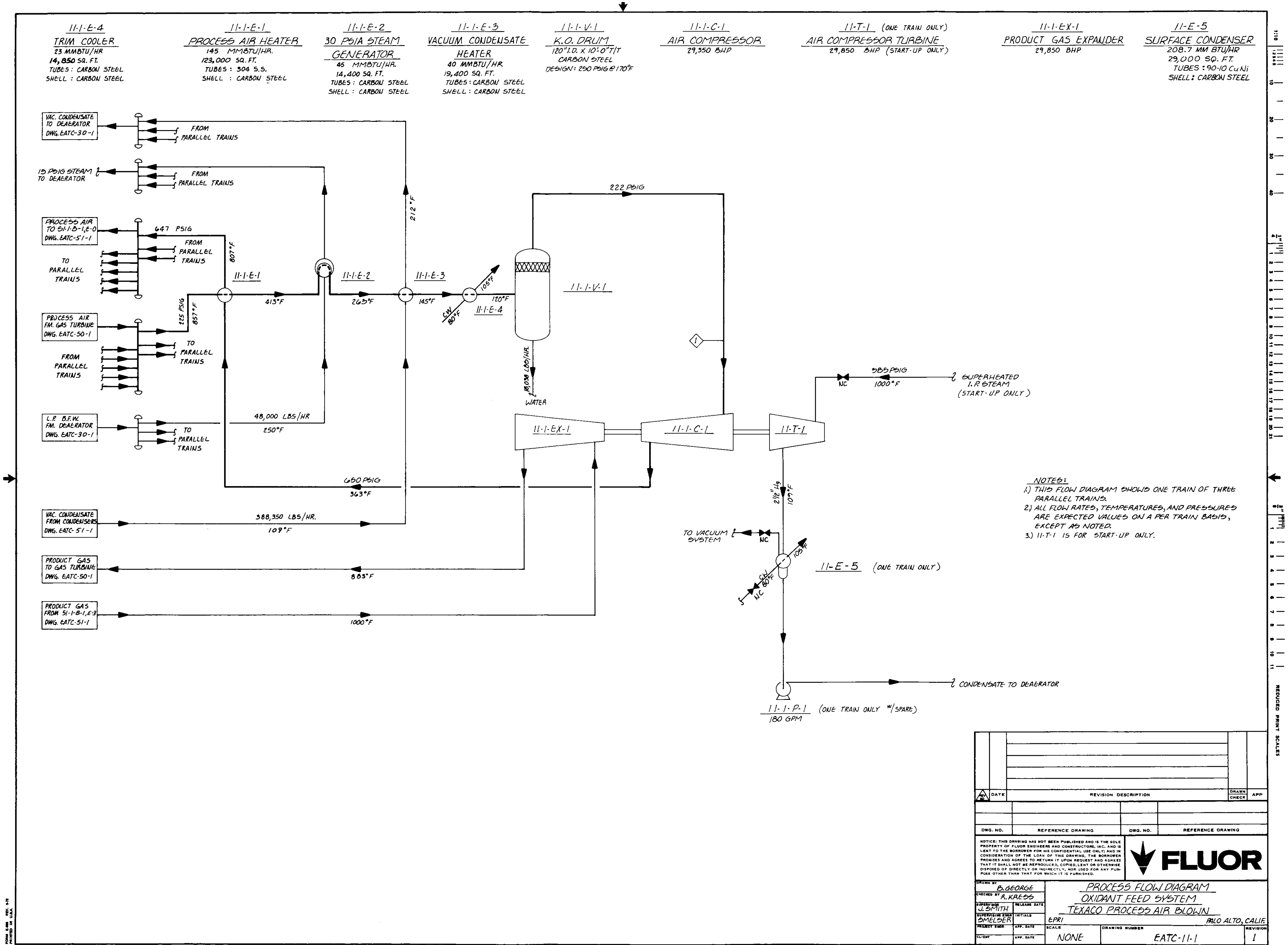
Process air in Case EATC is extracted from the gas turbines at 857° and 225 psig. The air is cooled to 120° and compressed to 650 psig in the booster compressor, 11-1-C-1. Heat in the extracted air is recovered by heating air leaving the booster compressor exchanger (11-1-E-1). The remaining cooling generates LP steam (11-1-E-2), and heats condensate (11-1-E-3). Some extraction air heat is rejected to cooling water (11-1-E-4).

Finally, the process air leaving exchanger 11-1-E-1 is heated to 1000°F in the HRSG.

The 29,550 hp required by each air booster compressor is supplied by a fuel gas expander, 11-1-EX-1. One train is provided with a steam turbine driver (11-T-1) for start-up. The steam turbine driver is a condensing type machine operating at inlet conditions of 385 psig, 1000°F, with exhaust pressure at 2-1/2" Hg abs. This steam turbine provides added capacity for operation during turndown or upset conditions. Each of the three operating fuel gas expanders are driven by fuel gas which has been preheated to 1000°F in the heat recovery steam generators (HRSGs) located in Unit 51. The fuel gas is expanded from 497 psig, 1000°F to approximately 280 psig, 822°F and flows to the gas turbines.

Equipment Notes

The compressors are commercially available. The turbines with 1000°F inlet temperatures represent an extension of the present state of the art for turbines. However, no problem is expected in obtaining these turbines in the next few years.



11-1-E-4
TRIM COOLER
23 MMBTU/HR
14,850 SQ. FT.
TUBES: CARBON STEEL
SHELL: CARBON STEEL
11-1-E-1
PROCESS AIR HEATER
145 MMBTU/HR.
123,000 SQ. FT.
TUBES: 304 S.S.
SHELL: CARBON STEEL
11-1-E-2
30 PSIA STEAM GENERATOR
45 MMBTU/HR.
14,400 SQ. FT.
TUBES: CARBON STEEL
SHELL: CARBON STEEL
11-1-E-3
VACUUM CONDENSATE HEATER
40 MMBTU/HR
19,400 SQ. FT.
TUBES: CARBON STEEL
SHELL: CARBON STEEL
11-1-V-1
K.O. DRUM
120" I.D. X 10'-0" T/T
CARBON STEEL
DESIGN: 250 PSIG @ 170°F
11-1-C-1
AIR COMPRESSOR
29,550 BHP
11-T-1 (ONE TRAIN ONLY)
AIR COMPRESSOR TURBINE
29,850 BHP (START-UP ONLY)
11-1-EX-1
PRODUCT GAS EXPANDER
29,850 BHP
11-E-5
SURFACE CONDENSER
208.7 MM BTU/HR
29,000 SQ. FT.
TUBES: 90-10 CuNi
SHELL: CARBON STEEL
VAC. CONDENSATE TO DEAERATOR DWG. EATC-30-1
FROM PARALLEL TRAINS
15 PSIG STEAM TO DEAERATOR
FROM PARALLEL TRAINS
PROCESS AIR TO 51-1-B-1,E-0 DWG. EATC-51-1
647 PSIG
FROM PARALLEL TRAINS
TO PARALLEL TRAINS
807°F
PROCESS AIR FM. GAS TURBINE DWG. EATC-50-1
225 PSIG
857°F
TO PARALLEL TRAINS
FROM PARALLEL TRAINS
L.P. B.F.W. FM. DEAERATOR DWG. EATC-30-1
TO PARALLEL TRAINS
11-1-E-1
413°F
11-1-E-2
263°F
11-1-E-3
145°F
212°F
CW 80°F
105°F
120°F
11-1-E-4
222 PSIG
11-1-V-1
1
18,038 LBS/HR.
WATER
48,000 LBS/HR
250°F
11-1-EX-1
11-1-C-1
11-T-1
585 PSIG
1000°F
NC
SUPERHEATED I.P. STEAM (START-UP ONLY)
650 PSIG
363°F
VAC. CONDENSATE FROM CONDENSERS DWG. EATC-51-1
388,350 LBS/HR.
109°F
2½" Hg
109°F
TO VACUUM SYSTEM
NC
11-E-5 (ONE TRAIN ONLY)
CW 80°F
NC
105°F
PRODUCT GAS TO GAS TURBINE DWG. EATC-50-1
883°F
PRODUCT GAS FROM 51-1-B-1,E-3 DWG. EATC-51-1
1000°F
CONDENSATE TO DEAERATOR
11-1-P-1 (ONE TRAIN ONLY W/SPARE)
180 GPM
NOTES:
1.) THIS FLOW DIAGRAM SHOWS ONE TRAIN OF THREE PARALLEL TRAINS.
2.) ALL FLOW RATES, TEMPERATURES, AND PRESSURES ARE EXPECTED VALUES ON A PER TRAIN BASIS, EXCEPT AS NOTED.
3.) 11-T-1 IS FOR START-UP ONLY.
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REVISION DESCRIPTION
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CHECKED BY R. KRESS
SUPERVISOR J. SMITH
SMELSER
PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
TEXACO PROCESS AIR BLOWN
EPRI
PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER EATC-11-1
REVISION 1
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GAS COOLING

Case EATC

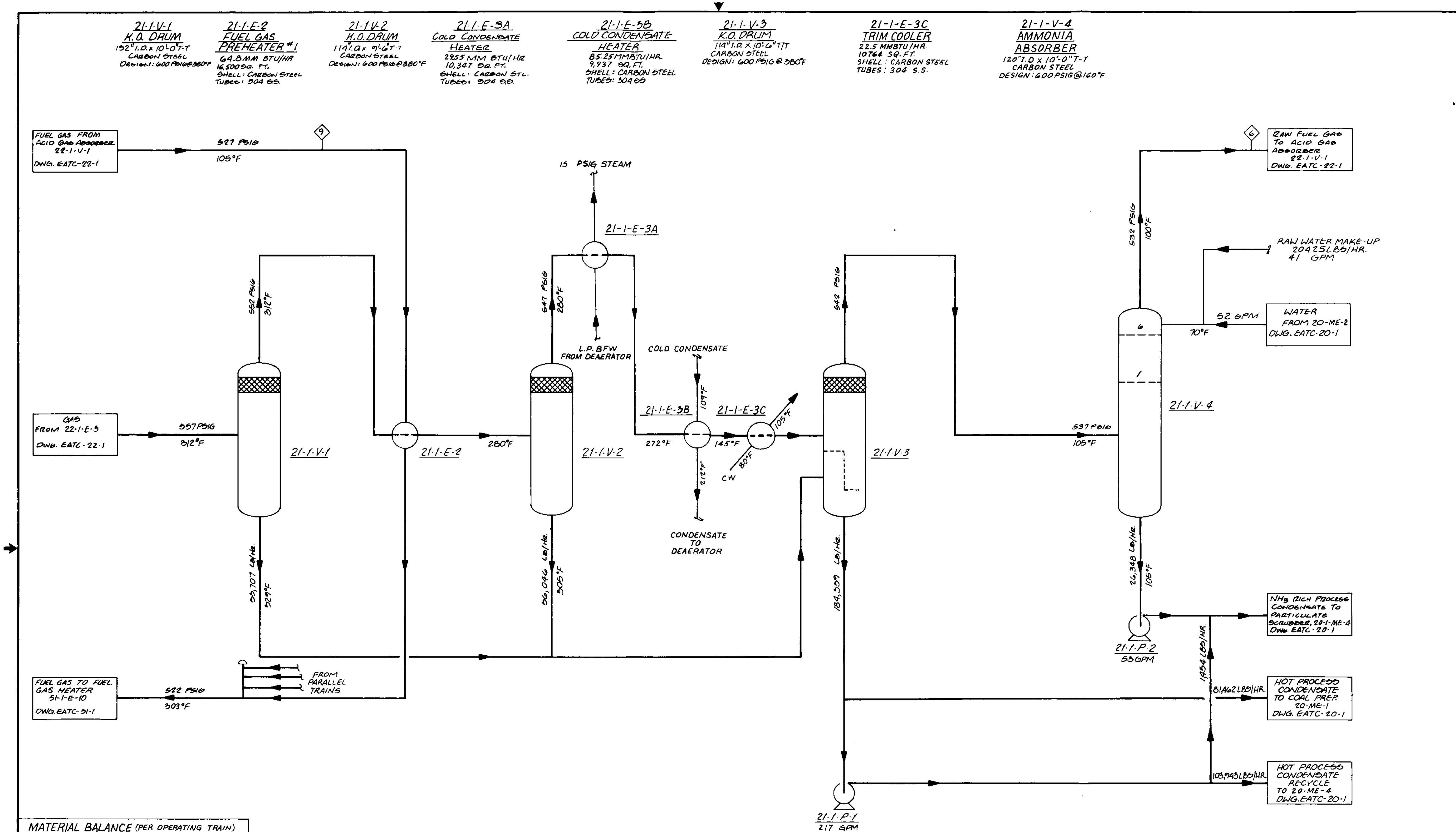
Process Flow Diagram EATC-21-1 shows one of the four parallel trains in the gas cooling section. No spare train is provided.

Clean gasifier effluent from the particulate scrubbing section, after passing through the Selexol reboiler, is cooled to approximately 105°F in a series of exchangers, 21-1-E-2 and 21-1-E-3 A, B and C. The effluent from the Selexol reboiler, after separation of condensate in the knockout drum 21-1-V-1, is then cooled by exchanging heat against fuel gas from the acid gas removal section. The condensate produced in cooling is separated in 21-1-V-2. Further gas cooling is obtained in exchanger 21-1-E-3 by making 30 psia steam and heating vacuum condensate; then the gas is trim cooled with water. 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. Hot condensate from 21-1-V-3 is then pumped to the gasification unit (Flow Diagram: EATC-20-1).

The overhead gases from knockout drum 21-1-V-3 flow to an ammonia absorber, 21-1-V-4. Ammonia is removed by contacting the gas countercurrently with the water on the trays. The ammonia-free overhead gases from the absorber then flow to the acid gas removal unit for H_2S removal. The ammonia-rich process condensate from the bottom of the absorber is pumped to the particulate scrubber unit, 20-1-ME-4.

Equipment Notes

All equipment is commercially available.



MATERIAL BALANCE (PER OPERATING TRAIN)						
	PROCESS STREAM NUMBER					
	5 PARTICULATE SCRUBBER OVHD		6 AMMONIA SCRUBBER OVHD		9 FUEL GAS	
COMPONENT	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	44.1	0.08	44.1	0.10	43.4	0.10
H_2	4919.7	9.30	4919.7	11.61	4904.2	11.86
CO	8238.0	15.58	8238.0	19.45	8181.7	19.79
CO_2	3250.4	6.15	3250.4	7.67	2684.0	6.49
H_2S	220.8	0.42	220.8	0.52	8.2	0.02
COS	19.5	0.04	19.5	0.05	11.9	0.03
N_2	25290.5	47.83	25290.5	59.71	25176.5	60.88
AR	303.4	0.57	303.4	0.72	302.0	0.73
NH_3	44.5	0.08	0.4	0	0.1	0
H_2O VAPOR	10544.1	19.94	71.2	0.17	41.5	0.10
TOTAL VAPOR MPH	52,875.0		42,358.0		41,535.5	
H_2O LIQ	0					
TOTAL LB/HR	1,293,587.0		1,115,017.0		1,076,987	
MOL WT VAPOR	24.46		26.32		26.04	

FORM E-388 REV. 9-72 PRINTED IN U.S.A.

11-75 DIETERICH-POST 2112X-6 .003 FILM

NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FOUR PARALLEL TRAINS.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY B. GEORGE
CHECKED BY R. KRESS
SUPERVISOR J. SMITH
RELEASE DATE
SUPERVISING ENGR SNELSER
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAMS
GAS COOLING
TEXACO PROCESS-AIR BLOWN
EPRI PALO ALTO, CALIFORNIA

SCALE NONE | DRAWING NUMBER EATC-21-1 | REVISION 3

ACID GAS REMOVAL

Case EATC

Process Flow Diagram EATC-22-1 depicts one of the four parallel acid gas removal trains. 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). Hydrogen sulfide in the crude gas is absorbed in Selexol® solvent in order to reduce sulfur in the treated gas to 1.0 pound sulfur dioxide (SO_2) equivalent per million Btu (HHV) coal charged to the plant.

The cooled ammonia-free gas flows through an acid gas absorber, 22-1-V-1, where it contacts Selexol® solvent countercurrently in a packed tower. The treated gas from the top of the absorber flows through a knockout drum, 22-1-V-3, and, after heat exchange with gasifier effluent in the upstream unit, to heat recovery steam generators (HRSG) located in Unit 51 for further heating.

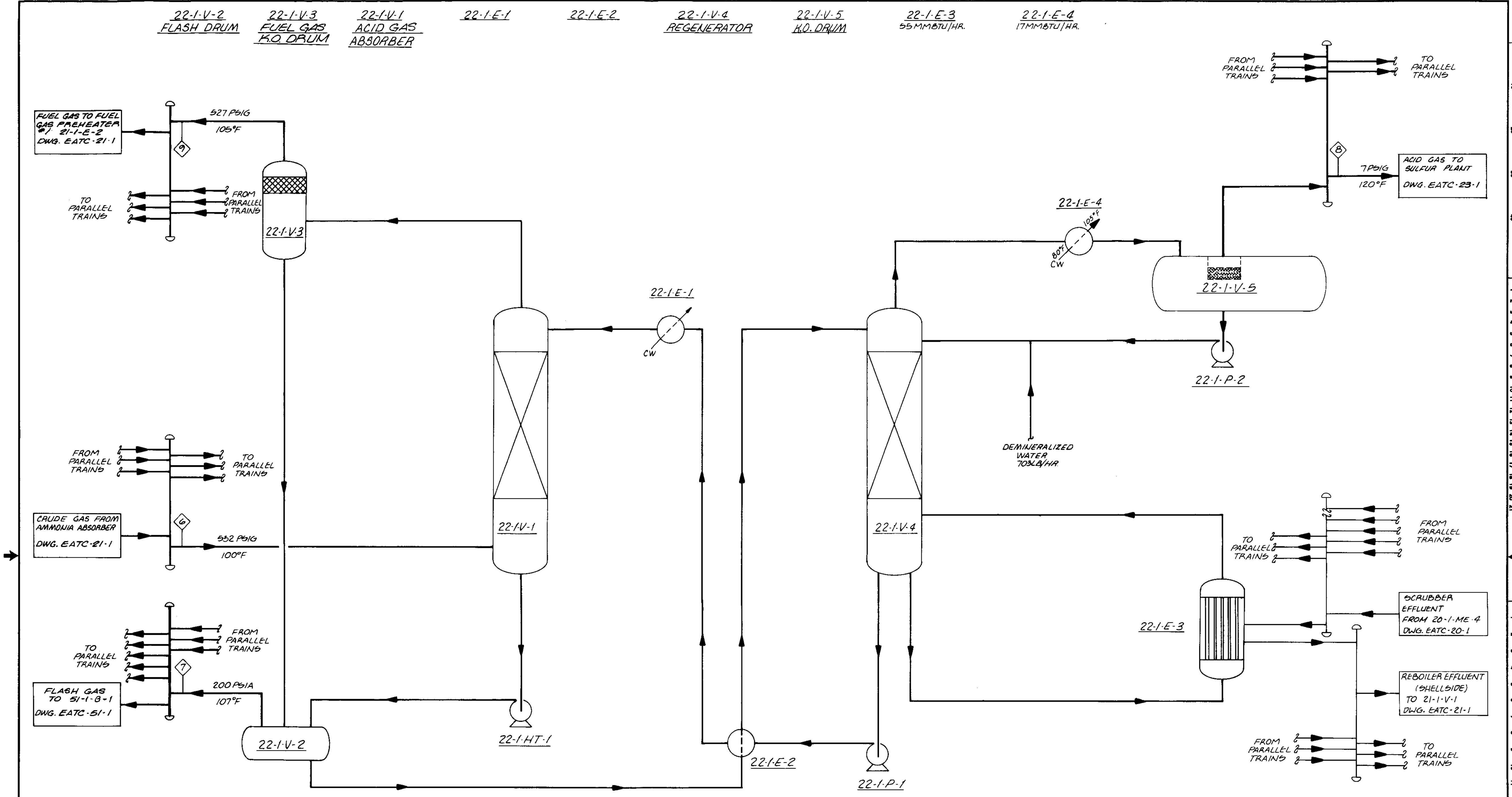
The rich solvent from the bottom of the absorber is let down through a hydraulic turbine, 22-1-HT-1, which supplies a portion of the power required by the lean solution pump, 22-1-P-1. It then flows to flash drum 22-1-V-2 where most of the dissolved hydrocarbon gases in the solvent flash off. Approximately 96% of the dissolved H_2S and most of the dissolved COS are retained in the solvent because of their selective absorption in the Selexol® solvent.

The rich solvent solution from the flash drum exchanges heat with hot regenerated solution in 22-1-E-2 and 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. Reboil heat is supplied by process gas from the particulate scrubbing section 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. Heat is first exchanged with rich solution in 22-1-E-2 in order to reduce reboiler duty. The lean solution is cooled down to operating temperature with cooling water in exchanger 22-1-E-1.

Acid gas from the regenerator overhead is cooled to 120°F in cooler 22-1-E-4. The condensate produced in cooling is separated in the knockout drum, 22-1-V-5, and then pumped back to the regenerator through 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 23.8% H_2S on a volume basis and flows to the sulfur recovery unit for further processing.

Equipment Notes

The majority of equipment in this section is all carbon steel. The equipment has been used in very similar service for a number of years.



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER ⟨6⟩ CRUDE GAS MPH	⟨6⟩ CRUDE GAS MOL%	⟨7⟩ FLASH GAS MPH	⟨7⟩ FLASH GAS MOL%	⟨8⟩ ACID GAS MPH	⟨8⟩ ACID GAS MOL%	⟨9⟩ FUEL GAS MPH	⟨9⟩ FUEL GAS MOL%
CH_4	44.1	0.10	0.3	.17	0.4	0.05	43.4	0.10
H_2	4,919.7	11.61	11.9	7.47	3.6	0.41	4,904.2	11.86
CO	8,238.0	19.45	34.1	21.43	22.2	2.51	8,181.7	19.79
CO_2	3,250.4	7.67	29.4	18.45	537.0	60.75	2,684.0	6.49
H_2S	220.8	0.52	1.9	1.20	210.7	23.83	8.2	0.02
COS	19.5	0.05	.3	0.13	7.3	.82	11.9	0.03
N_2	25,290.5	59.71	80.0	50.24	34.0	3.84	25,176.5	60.88
A_R	303.4	0.72	1.0	0.60	0.4	0.05	302.0	0.73
NH_3	0.4	0.00	0.0	0.00	.3	0.03	0.1	0.00
H_2O	71.2	0.17	.5	0.31	68.2	7.71	41.5	0.10
TOTAL MPH	42,358.0		159.5		884.1		41,353.5	
TOTAL LB/HR	1,115,017		4,650		34,095		1,076,987	
MOL WT	26.32		29.16		38.56		26.04	

FORM E-308 REV. 9-72 PRINTED IN U.S.A.

11-78 DIETERICH-POST 2111X-8 .003 FILM

NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FOUR PARALLEL TRAINS.

2. ALL FLOW RATES, TEMPERATURES, PRESSURES, AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

3. DESIGN INCLUDES: SOLVENT STORAGE TANK 22-TK-1 7800 BBL
SUMP PUMP 22-P-3A ¢ B 35 GPM
SOLVENT TRANSFER PUMP 22-P-4 135 GPM

REV NO	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP
1	12/16/77	ADD NOTE 3	BG / RCS	
02	10/25/77	MASS BALANCE AND DUTIES	BG / RK	
01	9/29/77	REBOILER CHANGED	BG / RCS	

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

FLUOR

DRAWN BY B. GEORGE
CHECKED BY R. SIEBERT
SUPERVISOR J. SMITH
SUPERVISING ENGR SNELSER
RELEASE DATE
INITIALS
PROJECT ENGR. APP. DATE
CLIENT APP. DATE

PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
TEXACO PROCESS-AIR BLOWN

EPRI PALO ALTO, CALIFORNIA

SCALE: NONE
DRAWING NUMBER: EATC-22-1
REVISION: 3

REDUCED PRINT SCALES

AAEB3-CTA-SS-1

SULFUR RECOVERY AND TAIL GAS TREATING

The processes used in these units for Case EATC (Slurry Feed) are similar to those in Case EXTC. Typical Process Flow Diagrams EATC-23-1 and EATC-24-1 are included for reference. Descriptions of these units have been published previously.[1]

There are two 50% parallel operating sulfur recovery trains each followed by a tail gas treating unit. Sulfur recovery per train is 154.5 short tons per day. There is a third (spare) train because of the important environmental requirements these units fulfill.

Equipment Notes

The Claus sulfur process is an established commercial process and consequently the equipment requirements are well known. Tail gas treating units are a more recent development; however, the equipment has been operated in many commercial plants successfully.

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1. "Economic Studies of Coal Gasification Combined Cycle System for Electric Power Generation," EPRI AF-642, Jan. 1978, Page 74.

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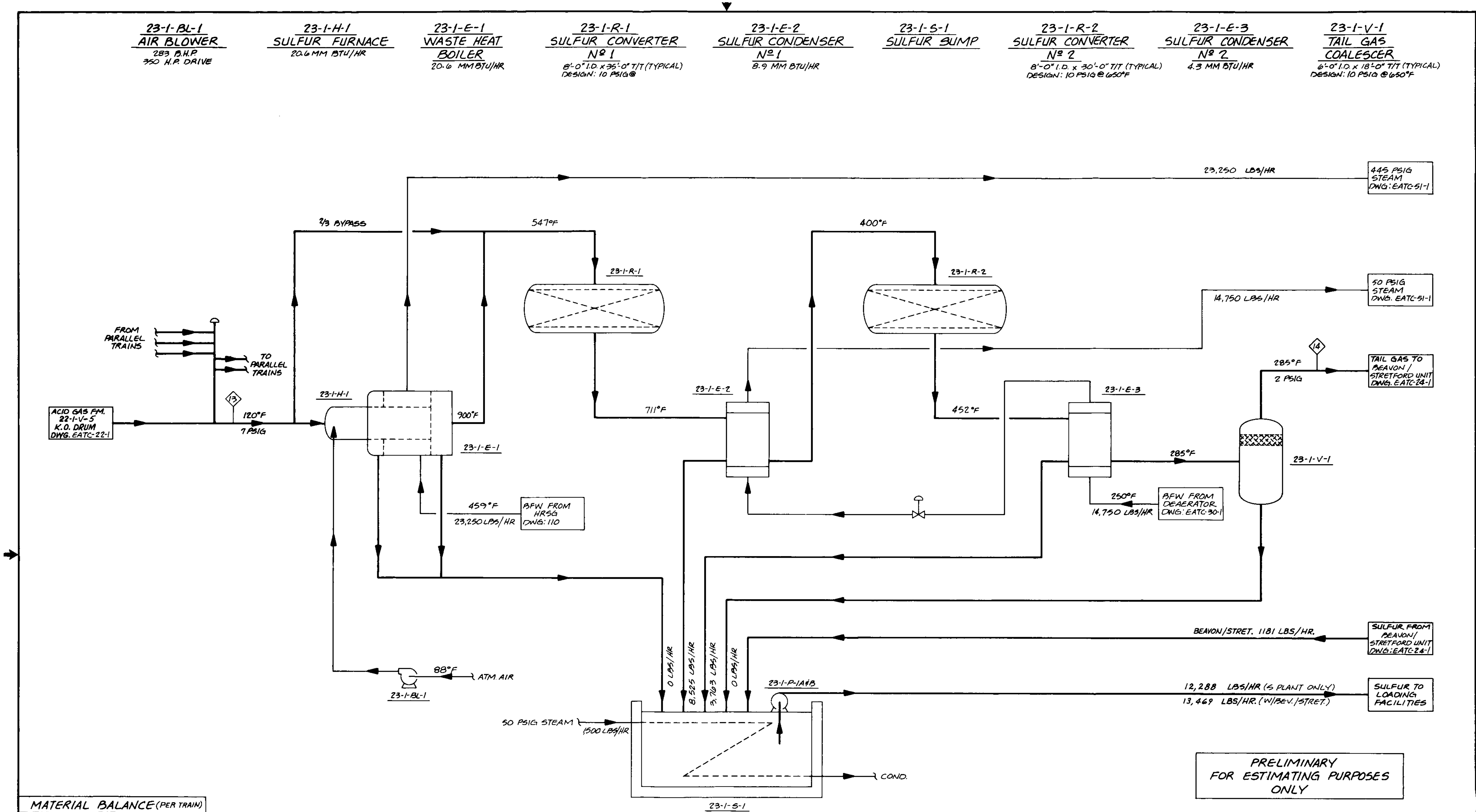


MATERIAL BALANCE (PER TRAIN)

STREAM №.	13		14	
	ACID GAS FM. H₂ SELEXOL UNIT		TAIL GAS TO T.G. TREATING UNIT	
COMPONENT	MPH	MOL%	MPH	MOL%
H_2	7.2	0.41	4.86	0.19
CO	44.3	2.53	29.56	1.16
CO_2	1074.0	61.31	1101.8	43.23
H_2S	405.1	23.12	31.0	1.22
COS	14.6	0.84	2.5	0.10
N_2	69.3	3.96	837.9	32.87
NH_3	0	0	0	0
CH_4	0.9	.05	0	0
H_2O VAPOR	136.4	7.79	540.3	21.19
SO_2	0	0	.86	.03
S_2	0.	0	0	0
S_6	0	0	0.05	0
S_8	0	0	0.23	.01
TOTAL VAPOR MPH	1751.8		2549.1	
H_2O LIQ.	0		0	
TOTAL LBS/HR	67623.8		83870.9	
MOL. WT. VAPOR	38.60		32.90	

FORM E-388 REV. 9-78 PRINTED IN U.S.A.

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, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE TYPICAL VALVES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
4. CONFIGURATION AND SIZE OF EQUIPMENT CORRESPOND TO A DESIGN SULFUR RECOVERY RATE OF 147.5 ST/D (91.8% RECOVERY)

REV.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

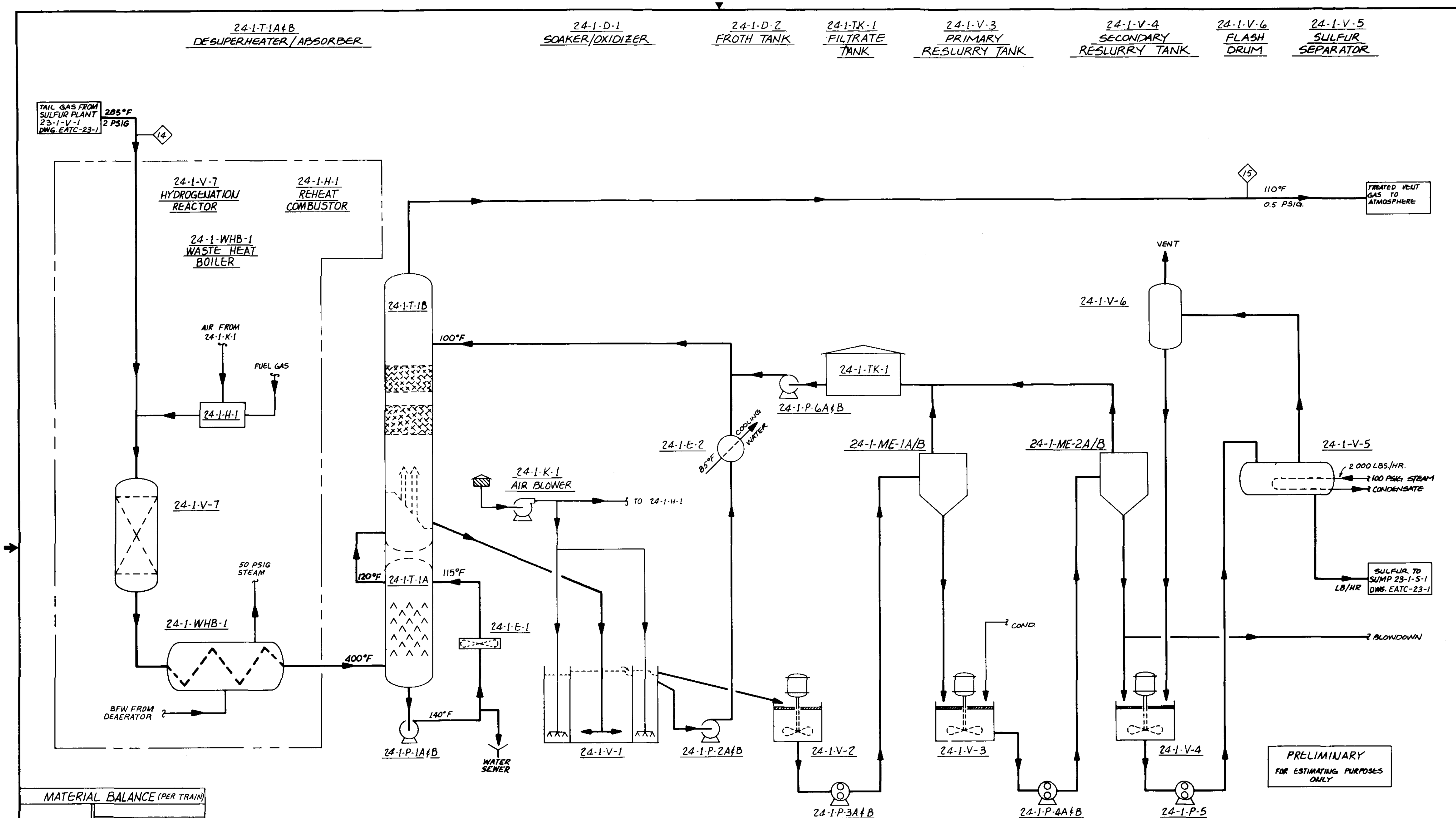
DWG. NO.	REFERENCE DRAWING	DWG. NO.	REFERENCE DRAWING

FLUOR

DRAWN BY: B. GUEVARA
CHECKED BY: J. SMITH
SUPERVISOR: J. SMITH
SUPERVISING ENGR: S. SMELSER
RELEASE DATE / INITIALS / PROJECT ENGR. / APP. DATE / CLIENT / APP. DATE

PROCESS FLOW DIAGRAM
SULFUR PLANT (TYPICAL)
TEXACO PROCESS-AIR BLOWN
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: EATC-23-1
REVISION: 2



MATERIAL BALANCE (PER TRAIN)

STREAM № / COMPONENT	14 TAIL GAS TO T.G. TREATING UNIT MPH	MOL %	15 ** VENT GAS FM. T.G. TREATING UNIT MPH	MOL %
H_2	4.86	0.19	4.86	0.23
CO	29.56	1.16	29.56	1.40
CO_2	1101.8	43.23	1104.3	52.27
H_2S	31.0	1.22	0.39*	100PPM
COS	2.5	0.10	0	0
N_2	837.9	32.87	837.9	39.67
NH_3	0	0	0	0
CH_4	0	0	0	0
H_2O VAPOR	540.3	.86	270.4	6.40
SO_2	0.86	0.03	–	0
S_2	0	21.19	–	0
S_6	.05	0	–	0
S_8	.23	.01	–	0
TOTAL VAPOR MPH	2549.1		2112.4	
H_2O LIQ.	0			
TOTAL LBS/HR	83870.9		77784.7	
MOL. WT. VAPOR	32.90		36.82	

* STATED AS H_2S BUT CONSISTS OF TOTAL SULFUR.

** EXCLUDING INCINERATION AND REDUCING GASES.

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, TEMPERATURES, PRESSURES AND EQUIPMENT SIZES ARE TYPICAL VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
3. CONFIGURATION AND SIZE OF EQUIPMENT CORRESPOND TO DESIGN BASIS OF 14.17 ST/D PER TRAIN SULFUR RECOVERY.

DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY B. GUEVARA
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
SUPERVISING ENGR. S. SMELSER

PROCESS FLOW DIAGRAM
BEAVON/STRETFORD UNIT (TYPICAL)
TEXACO PROCESS-AIR BLOWN
ELECTRIC POWER RESEARCH INSTITUTE — PALO ALTO, CALIFORNIA

SCALE: NONE — DRAWING NUMBER: EATC-24-1 — REVISION: 1

REDUCED PRINT SCALES

PROCESS CONDENSATE TREATING

Case EATC

Most of the sour process condensate generated in this case is used for the preparation of the coal slurry feed to the gasifiers. Some of the ammonia present as salts in the water decomposes to nitrogen and hydrogen at the temperature existing in the combustor zone of the gasifiers.

A small stream of ammonia (as ammonium salts) contaminated effluent leaving the process units is obtained in the gasification area (Flow Diagram: EATC-20-1). This effluent is treated in the effluent water treating unit (Unit 40). A unit for the recovery of by-product ammonia is therefore not provided in Case EATC.

Equipment Notes

All equipment is commercially available.

STEAM, BOILER FEEDWATER AND CONDENSATE

Case EATC

Process Flow Diagram EATC-30-1 schematically represents steam, boiler feedwater and condensate systems for this case.

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

High Pressure	HP - 1450 psig, 900°F
Intermediate Pressure	IP - 445 psig
Medium Pressure	MP - 115 psig
Low Pressure	LP - 50 psig
Low Pressure	LP - 15 psig

High pressure steam generation and superheating is carried out in the gas cooling units, 20-1-ME-3. Additional HP steam generation is obtained in heat recovery steam generators (HRSG) 51-1-B-1 of gas turbines 50-1-GT-1. There are six gas turbines and each gas turbine has its own HRSG. The HRSG is described in detail in Appendix A. The superheated HP steam from the process units combines with 900°F H.P. superheated steam from the HRSG superheaters (51-1-B-1:E-1). All the superheated HP steam is used to drive the single back pressure type turbine, 51-T-1. The HP end of Turbine 51-T-1 takes steam at 1450 psig, 900°F and exhausts at 445 psig.

Saturated intermediate pressure (IP) steam generation at 445 psig is obtained in the IP steam generators located in the sulfur plant, the gasification unit and the gas turbine coolers, 50-1-E-1. The saturated IP steam together with the exhaust steam from 51-T-1 is superheated to 1000°F in the HRSGs' reheaters (51-1-B-1:E-2). The superheated IP steam at 385 psig, 1000°F is then used in the IP end of 51-T-1 and the start-up condensing turbine, 11-T-1. The low pressure end of 51-T-1 exhausts steam to the 115 psig steam header.

A small quantity of the 115 psig steam is supplied to the sulfur plant (Unit 23). The balance of the 115 psig exhaust steam is used in the LP turbine, 51-T-2 and for BFW pump turbines. LP Turbine 51-T-2 and pump drivers are condensing turbines exhausting at 2-1/2" Hg abs.

The 50 psig steam header is supplied by generators in the gas cooling unit, 20-1-ME-3, and the steam generators in the sulfur plant. The 50 psig steam is mainly used in condensing turbine-generators 51-T-3 A and B, making additional electric power. A small amount of 50 psig steam is also used for steam tracing and in the sulfur pit.

The 15 psig steam header is supplied by steam generation in process exchangers, 21-1-E-3A, and in air cooler, 11-1-E-2. Primarily, this low pressure steam is used in the deaerators, 51-1-DA-1, with the balance entering the steam turbine-generators, 51-T-3 A and B, as admittance steam making electric power. The deaerator is a tray type unit operating at 15 psig. The deaerator provides for 10 minute storage.

Raw water is treated in a semiautomatic, resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for a 1505 psig boiler. Demineralized water is stored in Tank 30-TK-1. Demineralized water from the storage tank is transported to the deaerator through Pumps 30-P-4A&B. A small quantity of the makeup water is withdrawn from the discharge of these pumps and transported to Unit 22. The balance of the demineralized water flows to the deaerator, 51-DA-1. The condensate from the 115 psig and 50 psig steam users also flows to the deaerator.

The vacuum condensate from turbines 11-1-T-1, 51-T-2 and 51-T-3 A and B, is combined and flows to the deaerator after heat recovery from the crude gasifier effluents in 21-1-E-3B and also in the air cooler, 11-1-E-3.

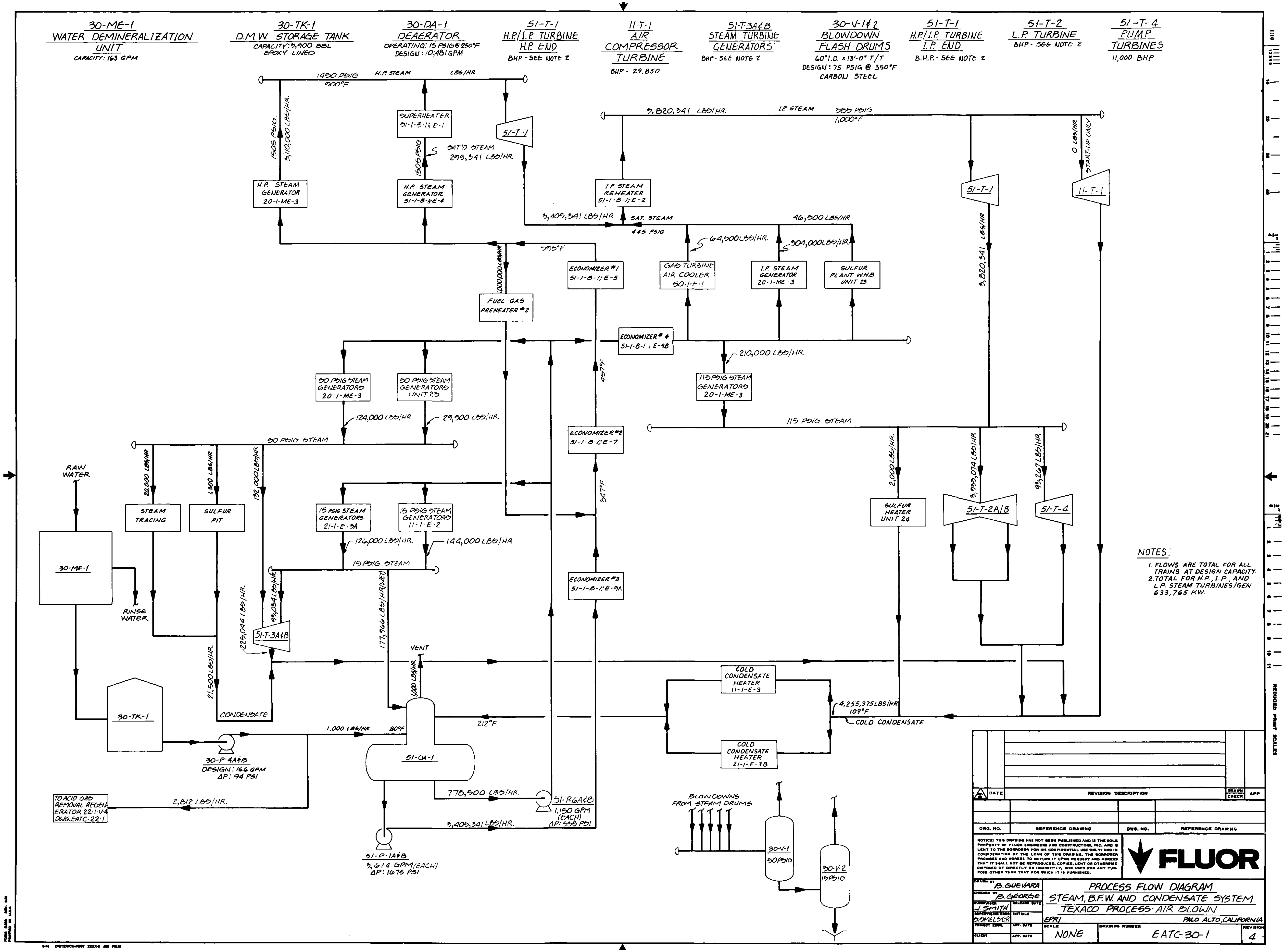
HP boiler feedwater (BFW) from the deaerator is pumped through high pressure boiler feedwater pumps (51-P-1A&B) to the HRSGs and process HP steam generators in 20-1-ME-1.

HP BFW to the HRSGs is first heated near the HP steam saturation temperature (598°F) in economizers 51-1-B-1:E-9A, E-7 and E-5. Part of the BFW is withdrawn downstream of 51-1-B-1:E-5 and supplied to the process HP steam generators in the gasification unit, and a portion of the hot high pressure BFW is used to preheat fuel gas in 51-1-E-10. The balance of the high pressure BFW flows to the HP steam generators in the HRSGs where saturated high pressure steam is generated.

LP BFW is heated to 347°F in an HRSG economizer, 51-1-B-1; E-9B, after the 15 psig process users in Units 11 and 20 have been supplied. The hot LP BFW feeds steam generators in the sulfur plant, at the gas turbine and in process steam generators in Units 20 and 21.

<u>Equipment Notes</u>

The combined cycle equipment is discussed in the next section and Appendix A.


30-ME-1
WATER DEMINERALIZATION UNIT
CAPACITY: 163 GPM
30-TK-1
D.M.W. STORAGE TANK
CAPACITY: 5,900 BBL
EPOXY LINED
30-DA-1
DEAERATOR
OPERATING: 15 PSIG @ 250°F
DESIGN: 10,481 GPM
51-T-1
H.P./I.P. TURBINE
H.P. END
BHP - SEE NOTE 2
11-T-1
AIR COMPRESSOR TURBINE
BHP - 29,850
51-T-3A&B
STEAM TURBINE GENERATORS
BHP - SEE NOTE 2
30-V-1&2
BLOWDOWN FLASH DRUMS
60" I.D. x 13'-0" T/T
DESIGN: 75 PSIG @ 350°F
CARBON STEEL
51-T-1
H.P./I.P. TURBINE
I.P. END
B.H.P. - SEE NOTE 2
51-T-2
L.P. TURBINE
BHP - SEE NOTE 2
51-T-4
PUMP TURBINES
11,000 BHP
1450 PSIG
900°F
H.P. STEAM
LBS/HR
1505 PSIG
3,110,000 LBS/HR.
SUPERHEATER
51-1-B-1; E-1
SAT'D STEAM
295,341 LBS/HR.
1505 PSIG
H.P. STEAM GENERATOR
20-1-ME-3
H.P. STEAM GENERATOR
51-1-B-&E-4
3,820,341 LBS/HR.
I.P. STEAM
385 PSIG
1,000°F
I.P. STEAM REHEATER
51-1-B-1; E-2
3,405,341 LBS/HR
SAT. STEAM
445 PSIG
46,500 LBS/HR
64,500 LBS/HR.
304,000 LBS/HR.
0 LBS/HR
START-UP ONLY
11-T-1
595°F
1,000,000 LBS/HR
ECONOMIZER #1
51-1-B-1; E-5
GAS TURBINE AIR COOLER
50-1-E-1
I.P. STEAM GENERATOR
20-1-ME-3
SULFUR PLANT W.H.B.
UNIT 23
3,820,341 LBS/HR
FUEL GAS PREHEATER #2
ECONOMIZER #4
51-1-B-1; E-9B
210,000 LBS/HR.
457°F
50 PSIG STEAM GENERATORS
20-1-ME-3
50 PSIG STEAM GENERATORS
UNIT 23
115 PSIG STEAM GENERATORS
20-1-ME-3
124,000 LBS/HR
29,500 LBS/HR.
115 PSIG STEAM
50 PSIG STEAM
ECONOMIZER #2
51-1-B-1; E-7
RAW WATER
20,000 LBS/HR
1,500 LBS/HR
132,000 LBS/HR
2,000 LBS/HR.
3,499,074 LBS/HR
93,267 LBS/HR
347°F
STEAM TRACING
SULFUR PIT
15 PSIG STEAM GENERATORS
21-1-E-3A
15 PSIG STEAM GENERATORS
11-1-E-2
SULFUR HEATER
UNIT 24
51-T-2A/B
51-T-4
126,000 LBS/HR.
144,000 LBS/HR
15 PSIG STEAM
30-ME-1
ECONOMIZER #3
51-1-B-1; E-9A
RINSE WATER
93,034 LBS/HR
229,044 LBS/HR.
177,966 LBS/HR (WET)
51-T-3A&B
VENT
21,500 LBS/HR.
1,000 LBS/HR
CONDENSATE
30-TK-1
COLD CONDENSATE HEATER
11-1-E-3
4,255,375 LBS/HR
109°F
COLD CONDENSATE
212°F
1,000 LBS/HR
80°F
30-P-4A&B
DESIGN: 166 GPM
ΔP: 94 PSI
51-DA-1
COLD CONDENSATE HEATER
21-1-E-3B
TO ACID GAS REMOVAL REGENERATOR 22-1-V-4
DWG. EATC-22-1
2,812 LBS/HR.
778,500 LBS/HR.
51-P-6A&B
1,150 GPM (EACH)
ΔP: 555 PSI
3,405,341 LBS/HR.
BLOWDOWNS FROM STEAM DRUMS
30-V-1
50 PSIG
30-V-2
15 PSIG
51-P-1A&B
3,614 GPM (EACH)
ΔP: 1675 PSI
NOTES:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. TOTAL FOR H.P., I.P., AND L.P. STEAM TURBINES/GEN. 633,765 KW.
DATE
REVISION DESCRIPTION
DWG. NO.
REFERENCE DRAWING
FLUOR
DRAWN BY B. GUEVARA
CHECKED BY B. GEORGE
J. SMITH
S. OMELSER
PROCESS FLOW DIAGRAM
STEAM, B.F.W. AND CONDENSATE SYSTEM
TEXACO PROCESS - AIR BLOWN
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EATC-30-1
REVISION 4

COMBINED CYCLE SYSTEM

Case EATC

Process Flow Diagrams EATC-50-1 and EATC-51-1 depict the combined cycle system for Case EATC. These diagrams show the total power block flows.

It is important to point out here that the combined cycle for Case EATC was designed by Fluor and not by Westinghouse as in Case EXTC. However, the Fluor calculations were based on six sets of cycle data provided by Westinghouse for the main report.[1]

There are six parallel trains of gas turbines 50-1-GT-1, generators 50-1-G-1 and heat recovery steam generators (HRSG) 51-1-B-1, and one 100 percent steam turbine (51-T1&2) and generator unit (51-G-1). Refer to Appendix A for a detailed description of the combined cycle system. Detailed performance information of the power block components, i.e., gas turbines, HRSG's and the steam turbine is provided in Appendix A.

The combined cycle system for Case EATC has the following distinct features:

- Equipment for fuel gas preheating is provided in Case EATC. Fuel gas produced in the process plant at 522 psig and 303°F is first heated to 550°F by heat exchange against hot feed water at 598°F pumped from the outlet of economizer one 51-1-B-1:E-5 in fuel gas heater, 51-1-E-10. The cooled feed water flows back to economizer two, 51-1-E-7. The fuel gas is further heated to 1000°F in a coil 51-1-B-1:E-3 provided in the reheater section of the HRSG. The hot fuel gas from the HRSG is subsequently expanded from 495 psig to 280 psig in expanders 11-1-EX-1 to supply the air booster compressors' power. Air compressor power for start-up is provided by one condensing steam turbine which takes steam from the hot reheat line.

- The boiler feed water pump drives use 115 psig steam.

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

The process cooling loads, where possible, are integrated into the condensate and makeup systems. Approximately 441 MM Btu/hr of low level process heat is recovered by heating cold condensate.

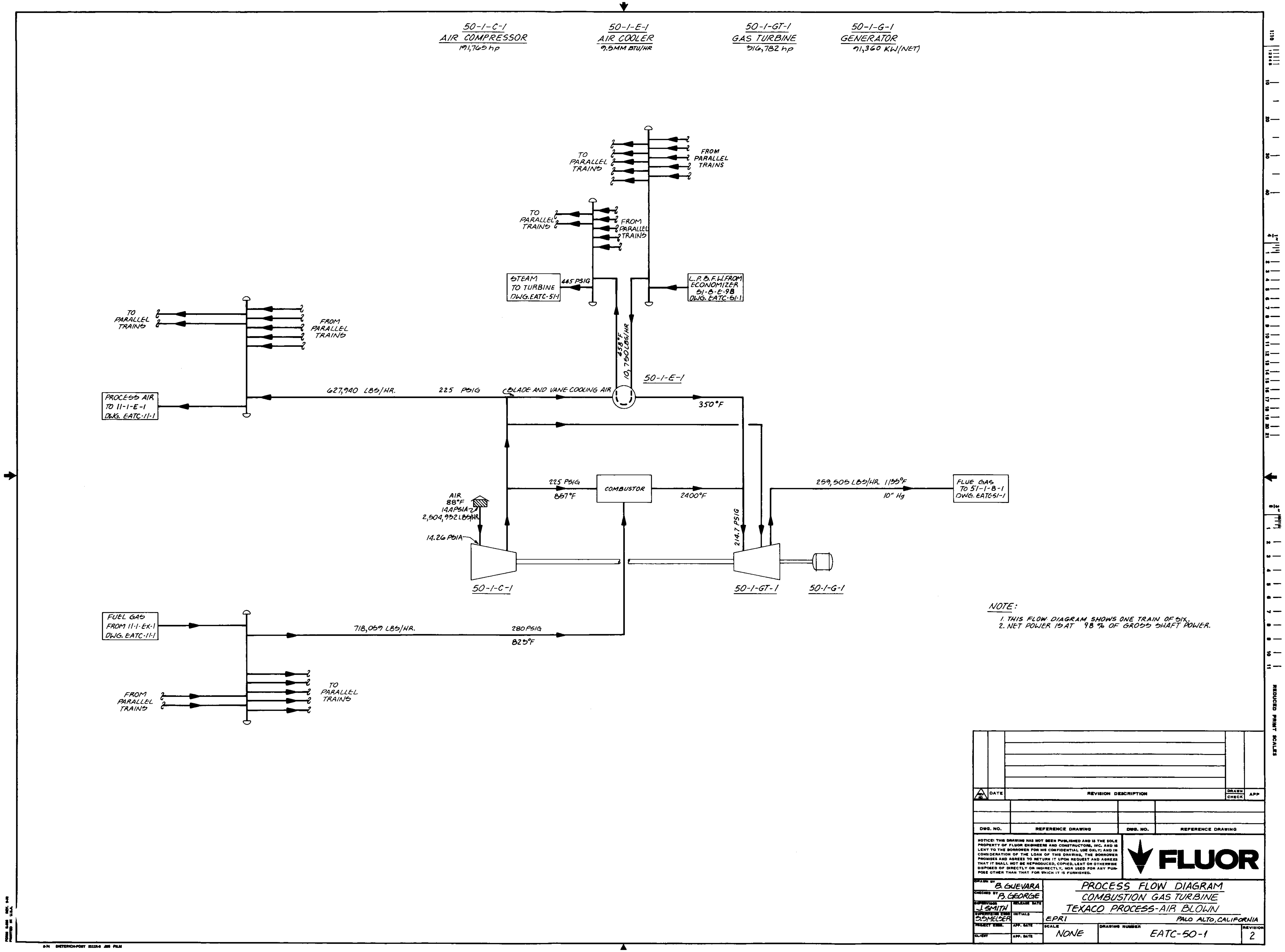
Equipment Notes

The proposed scheme of heating fuel gas in the HRSG (51-1-B-1:E-3) is a standard practice in ammonia plants and refinery units such as hydrocrackers, hydrotreaters and crude units. Gaseous or liquid hydrocarbon streams are commonly heated in coils placed in fired heaters which have heat recovery sections similar to HRSGs.

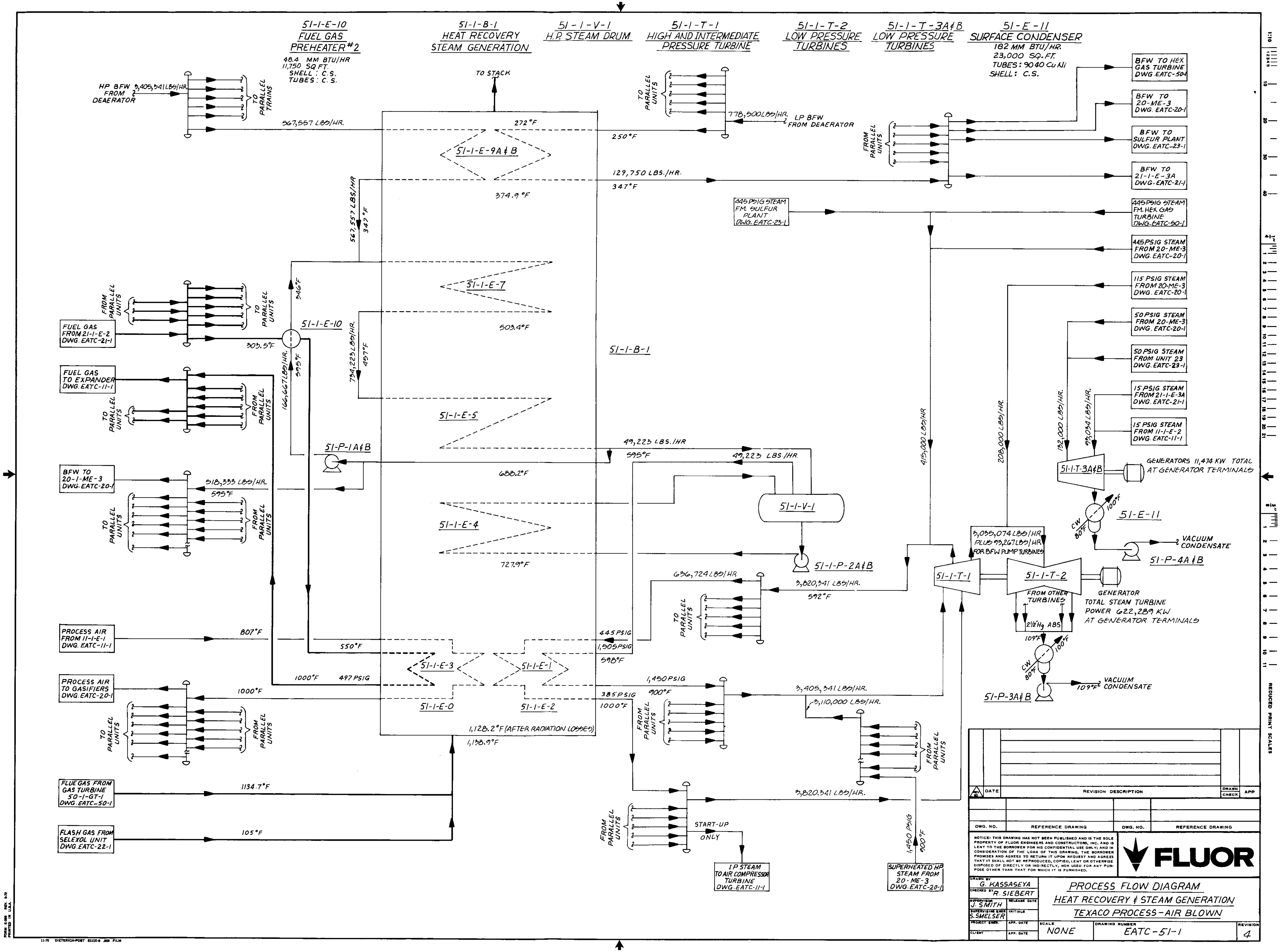
As the HRSGs recover heat from gas turbines' exhaust gases, the coils in HRSGs are not exposed to a direct radiant source, which happens in the fired heaters mentioned above. The HRSG's coils are therefore exposed to less severe conditions because of limited chances of localized hot spots occurring in the coils.

In case of fuel gas heater coil rupture in an HRSG, there will be fire in the HRSG's box. However, appropriate controls will be provided for the emergency shutdown of the affected HRSG and associated equipment and for the injection of the snuffing steam to the box to extinguish the fire. Since multiple HRSG trains are provided, it will still be possible to operate the plant at reduced load. This scheme is therefore safe and commercially proven.

Refer to Appendix A for other comments on the equipment state of art.



50-1-C-1
AIR COMPRESSOR
191,765 hp
50-1-E-1
AIR COOLER
9.5MM BTU/HR
50-1-GT-1
GAS TURBINE
316,782 hp
50-1-G-1
GENERATOR
71,360 KW(NET)
TO PARALLEL TRAINS
FROM PARALLEL TRAINS
STEAM TO TURBINE DWG. EATC-51-1
445 PSIG
L.P. B.F.W. FROM ECONOMIZER 51-8-E-98 DWG. EATC-51-1
458°F
10,750 LBS/HR
50-1-E-1
PROCESS AIR TO 11-1-E-1 DWG. EATC-11-1
627,940 LBS/HR.
225 PSIG
BLADE AND VANE COOLING AIR
350°F
225 PSIG
867°F
COMBUSTOR
2400°F
259,505 LBS/HR. 1135°F
10" Hg
FLUE GAS TO 51-1-8-1 DWG. EATC-51-1
AIR
88°F
14.4 PSIA
2,504,952 LBS/HR
14.26 PSIA
214.7 PSIG
50-1-C-1
50-1-GT-1
50-1-G-1
FUEL GAS FROM 11-1-EX-1 DWG. EATC-11-1
718,059 LBS/HR.
280 PSIG
825°F
NOTE:
1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF SIX.
2. NET POWER IS AT 98 % OF GROSS SHAFT POWER.
REDUCED PRINT SCALES
DATE
REVISION DESCRIPTION
DRAWN
CHECK
APP
DWG. NO.
REFERENCE DRAWING
DWG. NO.
REFERENCE DRAWING
FLUOR
DRAWN BY B. GUEVARA
CHECKED BY B. GEORGE
J. SMITH
S. SMELSER
PROCESS FLOW DIAGRAM
COMBUSTION GAS TURBINE
TEXACO PROCESS-AIR BLOWN
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EATC-50-1
REVISION 2



51-I-E-10
FUEL GAS
PREHEATER #2
48.4 MM BTU/HR
11,750 SQ.FT.
SHELL: C.S.
TUBES: C.S.
51-I-B-I
HEAT RECOVERY
STEAM GENERATION
51-I-V-I
H.P. STEAM DRUM
51-I-T-I
HIGH AND INTERMEDIATE
PRESSURE TURBINE
51-I-T-2
LOW PRESSURE
TURBINES
51-I-T-3A & B
LOW PRESSURE
TURBINES
51-E-II
SURFACE CONDENSER
182 MM BTU/HR.
23,000 SQ.FT.
TUBES: 90-10 Cu Ni
SHELL: C.S.
HP BFW 3,405,341 LBS/HR
FROM
DEAERATOR
TO PARALLEL TRAINS
567,557 LBS/HR.
TO STACK
272°F
250°F
51-I-E-9A & B
129,750 LBS./HR.
347°F
374.9°F
TO PARALLEL UNITS
778,500 LBS/HR.
LP BFW
FROM DEAERATOR
FROM PARALLEL UNITS
BFW TO HEX GAS TURBINE DWG. EATC-50-I
BFW TO 20-ME-3 DWG. EATC-20-I
BFW TO SULFUR PLANT DWG. EATC-23-I
BFW TO 21-I-E-3A DWG. EATC-21-I
445 PSIG STEAM FM. SULFUR PLANT DWG. EATC-23-I
445 PSIG STEAM FM. HEX GAS TURBINE DWG. EATC-50-I
445 PSIG STEAM FROM 20-ME-3 DWG. EATC-20-I
115 PSIG STEAM FROM 20-ME-3 DWG. EATC-20-I
50 PSIG STEAM FROM 20-ME-3 DWG. EATC-20-I
50 PSIG STEAM FROM UNIT 23 DWG. EATC-23-I
15 PSIG STEAM FROM 21-I-E-3A DWG. EATC-21-I
15 PSIG STEAM FROM 11-I-E-2 DWG. EATC-11-I
567,557 LBS/HR
347°F
346°F
FROM PARALLEL UNITS
TO PARALLEL UNITS
FUEL GAS FROM 21-I-E-2 DWG. EATC-21-I
305.5°F
51-I-E-10
51-I-E-7
503.4°F
754,223 LBS/HR.
457°F
166,667 LBS/HR.
595°F
51-I-B-I
FUEL GAS TO EXPANDER DWG. EATC-11-I
TO PARALLEL UNITS
FROM PARALLEL UNITS
51-I-E-5
51-P-IA & B
49,223 LBS./HR
595°F
49,223 LBS./HR.
688.2°F
415,000 LBS/HR
208,000 LBS/HR.
132,000 LBS/HR.
93,034 LBS/HR.
GENERATORS 11,474 KW TOTAL
AT GENERATOR TERMINALS
51-I-T-3A & B
BFW TO 20-I-ME-3 DWG. EATC-20-I
518,333 LBS/HR.
595°F
TO PARALLEL UNITS
FROM PARALLEL UNITS
51-I-V-I
51-I-E-4
CW 80°F
100°F
51-E-II
VACUUM CONDENSATE
51-P-4A & B
3,033,074 LBS/HR
PLUS 93,267 LBS/HR
FOR BFW PUMP TURBINES
727.9°F
51-I-P-2A & B
636,724 LBS/HR
3,820,341 LBS/HR.
592°F
51-I-T-I
51-I-T-2
GENERATOR
TOTAL STEAM TURBINE
POWER 622,289 KW
AT GENERATOR TERMINALS
FROM OTHER TURBINES
2 1/2" Hg ABS
TO PARALLEL UNITS
PROCESS AIR FROM 11-I-E-I DWG. EATC-11-I
807°F
550°F
445 PSIG
1,505 PSIG
598°F
109°F
CW 80°F
100°F
51-I-E-3
51-I-E-I
1000°F
497 PSIG
1,450 PSIG
900°F
385 PSIG
1000°F
VACUUM CONDENSATE
109°F
51-P-3A & B
PROCESS AIR TO GASIFIERS DWG. EATC-20-I
1000°F
51-I-E-0
51-I-E-2
3,405,341 LBS/HR.
3,110,000 LBS/HR.
TO PARALLEL UNITS
FROM PARALLEL UNITS
FROM PARALLEL UNITS
1,128.2°F (AFTER RADIATION LOSSES)
1,138.9°F
FLUE GAS FROM GAS TURBINE 50-I-GT-I DWG. EATC-50-I
1134.7°F
FLASH GAS FROM SELEXOL UNIT DWG. EATC-22-I
105°F
3,820,341 LBS/HR.
FROM PARALLEL UNITS
START-UP ONLY
IP STEAM TO AIR COMPRESSOR TURBINE DWG. EATC-11-I
1,450 PSIG
900°F
SUPERHEATED HP STEAM FROM 20-ME-3 DWG. EATC-20-I
DATE
REVISION DESCRIPTION
DRAWN
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DWG. NO.
REFERENCE DRAWING
DWG. NO.
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CHECKED BY R. SIEBERT
SUPERVISOR J. SMITH
S. SMELSER
PROCESS FLOW DIAGRAM
HEAT RECOVERY & STEAM GENERATION
TEXACO PROCESS - AIR BLOWN
SCALE NONE
DRAWING NUMBER EATC-51-I
REVISION 4
REDUCED PRINT SCALES

PROCESS DISCUSSION

The table below summarizes pertinent heat and material balance results.

TABLE EATC-2

SUMMARY OF SYSTEM PERFORMANCE - CASE EXTC AND EATC

	Case EXTC	Case EATC
GASIFICATION AND GAS CLEANING SYSTEM		
Coal Feed Rate, lbs/hr (m.f.)	798,333	798,333
Oxygen or Air (1)/Coal Ratio, lbs/lb m.f.	0.858	1.081
Oxidant Temperature, °F	300	1,000
Steam/Coal Ratio, lbs/lb m.f. (4)	0	0.0086
Slurry Water/Coal Ratio, lbs/lb m.f. (5)	0.503	0.522
Gasification Section Average Pressure, psig	600	600
Crude Gas Temperature, °F	2300-2600	2300-2600
Crude Gas HHV (dry basis), Btu/SCF (2)	281.1	102.5
Temperature of Fuel Gas to Gas Turbine, °F	781	821
POWER SYSTEM		
Gas Turbine Inlet Temperature, °F	2,400	2,400
Pressure Ratio	17:1	17:1
Turbine Exhaust Temperature, °F	1,140	1,135
Steam Conditions, psig/°F/°F	1,450/900/1,000	1,450/900/1,000
Condenser Pressure, Inches Hg abs	2.5	2.5
Stack Temperature, °F	272	272
Gas Turbine Power (3), MW	745	548
Steam Turbine Power (3), MW	448	634
Power Consumed, MW	36	43.5
Net System Power, MW	1,157	1,138
OVERALL SYSTEM		
Process and Deaerator Makeup Water, gpm/1000 MW	362	289
Cooling Tower Makeup Water, gpm/1000 MW	7,588	9,163
Cooling Water Circulation Rate, gpm/MW	347	415
Cooling Tower Heat Rejection, % of Coal HHV (6)	38.7	47.2
Air Cooler Heat Rejection, % of Coal HHV	5.2	0.8
Net Heat Rate, Btu/kWh	8,813	8,958
Overall System Efficiency (Coal → Power), % of Coal HHV	38.7	38.1

(1) Dry Basis, 100% O_2
(2) Excluding the HHV of H_2S, COS and NH_3
(3) At Generator Terminals
(4) Includes moisture in oxidant air
(5) Small changes in this ratio do not significantly alter the results presented here.
(6) See Table CS-1 for details.

Gasification

The required number of gasifiers for this case is not stated in this report because Texaco considers the information proprietary and subject to refinement. Texaco's experience to date has been primarily in oxygen blown coal gasification, so the required number and size of gasifiers for Case EXTC had been established with a high level of confidence. For air blown gasification, the minimum residence time requirement for coal slurry and air in their pilot plant reactor is still under investigation. In addition, Texaco is studying the optimum reactor size, and indicates that this might increase. Therefore, the actual number of gasifiers required for Case EATC could not be established with a high degree of confidence. For purposes of this economic screening evaluation it was, however, necessary to select some basis. Therefore, Fluor used the oxygen blown gasifier information from Case EXTC as a standard cost and size basis for one gasifier train. In order to set a conservative basis for this study Fluor added a penalty to the oxygen blown gasification residence time used in Case EXTC, and established the number of standard gasifiers which would be required for this study proportional to the total gas flow. Texaco has confirmed from pilot plant data that the penalty factor used for residence time represents a conservative engineering basis. Texaco states that further pilot plant studies of air blown gasification are expected to reduce the gasifier costs for Case EATC. The conservative number of gasifiers used in this study means that the gasification heat losses at one percent of coal HHV are larger than would result from an optimum design, and larger than actually did result in Case EXTC.

Gasifier material balances for full capacity operation are given in Tables EATC-3 and EXTC-3 for the air and oxygen blown Texaco gasifier cases.

Most of the data presented in the above two tables were received from Texaco Development Corporation. For the particular coal used for this study, Texaco indicated that slurry concentrations in the range of 60% solids to possibly 70% solids could be achieved. For Case EATC the slurry concentration was 65.7% solids. It is important to keep in mind, however, the fact that slurrying characteristics of coals vary greatly and that it is not valid to extrapolate performance estimates presented in this report to other coals that will process different slurrying characteristics. It is important too, to note that slurry concentrations this high are somewhat beyond the range of current practice. Should it prove necessary, in operation, to run with a more dilute slurry, there could be some adverse impact on overall plant efficiency.

TABLE EATC-3

MATERIAL BALANCE - CASE EATC

FEEDS

	T (°F)	lb/hr	lb mol/hr
Coal	163		
Moisture		35,000	1,942.8
Ash		80,000	
MAF Coal			
Carbon		554,985	46,205.9
Hydrogen		42,525	21,094.6
Oxygen		80,022	2,500.8
Nitrogen		9,985	356.4
Sulfur		30,816	961.1
TOTAL COAL		833,333	
Oxidant (dry)	1000		
Oxygen		863,337	26,979.3
Argon		48,469	1,213.4
Nitrogen		2,826,661	100,894.6
TOTAL OXIDANT		3,738,467	129,087.3
Water (including air moisture)	163	388,994	23,534.3
TOTAL FEEDS		4,960,794	

EFFLUENTS

	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	2,300-2,600			
CH_4		2,828	176.3	0.09
H_2		39,672	19,678.6	10.21
CO		923,019	32,952.0	17.09
CO_2		572,211	13,001.6	6.74
H_2S		30,101	883.2	0.46
COS		4,674	77.8	0.04
N_2		2,834,153	101,162.0	52.48
Ar		48,469	1,213.4	0.63
H_2O		422,646	23,459.5	12.17
NH_3		3,034	178.1	0.09
TOTAL GASIFIER EFFLUENT		4,880,794	192,781.9	100.00
	2,300-2,600			
Ash				
Carbon		Nil		
Ash		80,000		
TOTAL ASH		80,000		
TOTAL EFFLUENTS		4,960,794		

TABLE EXTC-3

MATERIAL BALANCE - CASE EXTC

FEEDS	T (°F)	lb/hr	lb mol/hr	EFFLUENTS	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Coal	140			Gasifier Effluent	2,300-2,600			
Moisture		35,000	1,942.8	CH_4		1,158	72.2	0.08
Ash		80,000		H_2		52,364	25,974.2	28.84
MAF Coal				CO		1,071,001	38,236.4	42.45
Carbon		554,985	46,205.9	CO_2		345,232	7,844.4	8.71
Hydrogen		42,525	21,094.6	H_2S		30,907	906.9	1.01
Oxygen		80,022	2,500.8	COS		3,256	54.2	0.06
Nitrogen		9,985	356.4	N_2		16,725	597.1	0.66
Sulfur		30,816	961.1	Ar		4,326	108.3	0.12
TOTAL COAL		833,333		H_2O		290,137	16,106.4	17.88
				NH_3		3,034	178.1	0.19
Oxidant	300			TOTAL GASIFIER EFFLUENT		1,818,140	90,078.2	100.00
Oxygen		684,687	21,397.3		2,300-2,600			
Argon		4,326	108.3	Ash				
Nitrogen		9,241	329.9	Carbon		Nil		
TOTAL OXIDANT		698,254	21,835.5	Ash		80,000		
				TOTAL ASH		80,000		
Water	140	366,553	20,364.1					
TOTAL FEEDS		1,898,140		TOTAL EFFLUENTS		1,898,140		

Little information is available on the production rate of trace compounds in this type of gasifier. It is known, for example, that in pilot runs, some of the nitrogen in the feed coal is converted to ammonia. In this design, ammonia has been assumed to be rapidly complexed as ammonium carbonate in the various process condensates. These ammonia bearing waters are eventually recycled to the gasifier via the coal slurry. At gasification temperatures, the equilibrium for ammonia formation is very unfavorable and the gasifier is thus assumed capable of destroying excess ammonia. The presence of ammonia in the process condensates is thought to have a beneficial effect by acting as a corrosion inhibitor. Small amounts of this water are removed from the plant and treated in water treatment facilities to destroy ammonia. The figures given in the tables and flow sheets for ammonia should be regarded as tentative estimates only.

Gasifier temperatures are believed to be high enough to destroy all hydrocarbons except methane.

Although the gasifiers were operated at essentially the same temperature, air blowing required 26% more oxygen than oxygen blowing. This oxygen converts 28% of the coal carbon to carbon dioxide versus 17% for oxygen blowing. As a result the air blown case produces only 82% as much chemical heat as measured by moles per hour of CO and H_2, a total of 52,630.6 versus 64,210.6 moles per hour in the oxygen case. In addition, we note that essentially no feedwater decomposes in the air blown reactors, 0.32%, compared to almost 28% in the oxygen blown case. Therefore, the hydrogen production in the air blown case is less than the hydrogen in the coal; whereas for oxygen blowing, the hydrogen produced is more than 20% greater than the coal hydrogen.

The number of gasifiers used in this study is based on pilot plant air flows. Since the pilot plant compressor was not large enough to provide the high air flow needed to study low residence times, the assumed number of gasifiers is conservatively high.

Acid Gas Removal

One of the important design considerations in coal gasification is acid gas removal. Acid gas removal processes tend to absorb both hydrogen sulfide (H_2S) and carbon dioxide (CO_2). While in many applications removal of both is desired, for gas turbine power generation there are substantial disadvantages to removing

CO_2. Absorption of CO_2 increases solvent circulation rates, equipment sizes and process heat loads, and takes away "working fluid" from the gas turbine generator. Further, the design and sizes of the downstream sulfur recovery units are affected in directions that increase cost. The Selexol® process removes H_2S in preference to CO_2 and, therefore, accomplishes an important objective. The process is used in this case because it accomplishes this objective and it compares favorably with other similar processes economically.

The Selexol® process results in an H_2S concentration over 23 percent in the acid gas feed to the sulfur recovery unit. At H_2S concentrations in this range, a sulfur plant design commonly referred to as "split flow" may be employed that avoids use of fuel gas in the sulfur furnace. Fuel gas must be burned in the furnace to sustain a flame if H_2S concentration is under 15 percent. In the split flow design the flame can be sustained by burning acid gas only.

Acid gas removal for air blown gasification required one more train than the three trains used for oxygen blowing, and larger diameter towers because of the presence of the nitrogen diluent. Energy consumption for regeneration of selexol solvent was greater in the air blown case due to lower partial pressures of acid gas components in the process gas and the associated higher solvent circulation rate.

Gas Cooling

Sensible heat in the gasifier effluent at 2300 to 2600°F, is removed to heat boiler feed water and make steam. Since the air blown case molar flow rate of effluent is more than twice the oxygen blown case, a significant increase in gas cooling duty results. To effectively integrate process heat into the HRSG/Steam-turbine cycle, the air blown case includes process steam superheating in the gas cooling unit whereas, in the oxygen blown case, process steam superheating is accomplished in the HRSG's. Process heat recovery for case EATC is 1.6 times the heat recovered in the HRSG's. To accommodate this heat load, heat exchangers in the gas cooling unit generate steam at several pressure levels.

HRSG and Steam Turbine

The reduced gas turbine flue gas flow rate for EATC along with the greater boiler feed water requirements of the process steam generators mean that less sensible heat is available in the HRSG units to generate steam. Accordingly the three steam pressure level design for EXTC has been rearranged to provide only high pressure steam generation. The other main difference in HRSG configuration between cases is the addition of a heat recovery coil used to heat process air to 1000°F.

Total high pressure superheated steam flow to the steam turbines is essentially the same for both cases, although most of the superheating of this steam is accomplished in the process units for the air blown case, and in the HRSGs for the oxygen blown case. Superheated intermediate pressure steam flow and low pressure steam flow to the turbine are approximately fifty percent larger in the air blown case due to additional process sensible heat carried by nitrogen in the gasifier effluent. To accommodate the additional sensible heat at lower process temperatures in the air blown case, low pressure process steam generators were added which feed low pressure steam to two small 50% steam turbine generators. These generators provide one percent of the total electric power produced.

Process Energy Balances

Tables EATC-4 and EXTC-4 present overall process energy balances for air and oxygen blown cases at 100 percent capacity operation. The boundary for each balance encompasses the entire plant. Energy content of streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kWh. These energy balances close to less than one-half percent. The discrepancies result from approximations used for some process units and for calculating some heat loads.

Data from Tables EATC-4 and EXTC-4 are shown in MM Btu/hr and as percent of coal higher heating value in Table EATC-5.

The tables show that the air blown case results in slightly less of the coal energy charged to the plant converted to power than does the oxygen blown case. Coal charged at 10,000 ton/day is equivalent to 10,196 MM Btu/hr HHV. The air blown case produces 3,885 MM Btu/hr as electrical energy or 38.1 percent of the coal HHV. The oxygen blown case produces 3,948 MM Btu/hr power equivalent or 38.7 percent of the coal HHV.

TABLE EATC-4

ENERGY BALANCE - CASE EATC

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

	MM Btu/hr					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Gas Turbine Suction Air	--	102	258			360
Demineralized and Raw Water		2				2
Auxiliary Power Inputs					149	149
TOTAL	10,196	109	258	0	149	10,712
HEAT OUT						
Ash Slurry		50				50
Gasifier Heat Losses				112		112
Gas Cooling		91				91
Sulfur Product	105	1				106
Air Coolers		82				82
Oxidant Compressor Interstage Cooling		70				70
Gas Turbines					1,871	1,871
Sulfur Plant Effluent Gas		2	18			20
Steam Turbines					2,163	2,163
Power Block Losses				47	193	240
Turbo-Generator Condensers			4,171			4,171
HRSG Stack Gas		816	579			1,395
Steam Heat and Power Losses		22	24			46
Selexol Overhead Condenser			68			68
Selexol Solvent Cooler		176				176
Waste Water Effluent		28				28
TOTAL	105	1,124	4,860	159	4,375	10,689

$$\frac{\text{Input - Output}}{\text{Input}} = 0.22\%$$

TABLE EXTC-4

ENERGY BALANCE CASE EXTC

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

	MM Btu/hr					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Air Compressor Suction Air		22	53			75
Gas Turbine Combustion Air		117	282			399
Demineralized and Raw Water		9				9
Auxiliary Power Inputs					123	123
TOTAL	10,196	153	335	0	123	10,807
HEAT OUT						
Ash Slurry		81				81
Gasifier Heat Losses				26		26
Gas Cooling		19	6			25
Sulfur Product	105	1				106
Oxidant Compressor Interstage Cooling		535	33			568
Oxidant Compressor Surface Condensers			1,067			1,067
Gas Turbines					2,541	2,541
Sulfur Plant Effluent Gas		2	19			21
Steam Turbines					1,530	1,530
Power Block Losses (1)				43	172	215
Steam Turbine Condenser			2,687			2,687
HRSG Stack Gas		1,027	790			1,817
Steam Heat Losses			22			22
Selexol Overhead Condenser			24			24
Selexol Solvent Cooler		54				54
Air Separation Plant Waste Gas		18				18
Waste Water Effluent		19				19
TOTAL	105	1,756	4,648	69	4,243	10,821

$$\frac{\text{Output - Input}}{\text{Input}} = 0.13\%$$

(1) Includes mechanical and electrical losses.

If all power consumed by the process units is included, the net system efficiency is 38.7 percent for the oxygen blown case compared to 38.1 percent for the air blown case. The heat rate based on net power produced is 8,813 Btu/kWh for the oxygen blown case and 8,958 Btu/kWh for the air blown case.

These net heat rates indicate oxygen blowing to be as efficient as air blowing, within the accuracy of the calculations done for these cases. This result was surprising, because past work[1] based on different gasification technologies had indicated air blown gasification might be more efficient.

Comparisons drawn from the tables illustrate some of the differences between the air and oxygen blown cases. Oxygen consumption is more for the air blown case, because it is necessary to supply combustion heat to heat air nitrogen and reactor heat losses are higher due to the larger number of reactors required. Power generated in the gas turbine is lower for the air blown case, and steam turbine power recovery is higher.

Energy leaving the plant as stack gas is higher for the oxygen blown case: 1,817 MM Btu/hr versus 1,395 MM Btu/hr for the air blown case. This represents 17.8 percent of the coal feed HHV for the oxygen case and 13.7 percent for the air blown case, and reflects the higher fuel flow rate (particularly hydrogen flow rate) existing in the oxygen blown case. The combined total heat losses in HRSG units and surface condenser units are essentially constant between the two cases (5571 MM Btu/hr for the oxygen blown case versus 5566 MM Btu/hr for air). Thus the energy recovery between the two cases is essentially the same. Although gas turbine power is lower for the air blown case, more steam turbine power is generated partially due to higher process steam generation and partially due to significantly less process steam demand for compressor drivers, the net result being that both cases generate essentially the same amount of electricity.

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

TABLE EATC-5

ENERGY BALANCE AS PERCENT COAL HHV - CASES EATC AND EXTC

	Case EXTC		Case EATC	
	MM Btu/hr	Percent	MM Btu/hr	Percent
IN				
Coal HHV	10,196	100.0	10,196	100.0
OUT				
Net Power	3,948	38.72	3,885	38.10
Sulfur Product, HHV	105	1.03	105	1.03
Ammonia Product, HHV	0	0	0	0
Selexol Sensible and Latent	171	.77	244	2.39
Oxidant Interstage Cooling	568	5.57	70	.69
Ash Slurry Sensible	81	0.79	50	.49
HRSG Stack Gases	1817	17.82	1,395	13.68
Rejected at Condensers	3754	36.82	4,171	40.91
Other Sensible Losses	(94)	(.92)	117	1.15
Other Latent Losses	(288)	(2.28)	(216)	(2.12)
Gasifier Heat Losses	26	0.26	112	1.10
Power Block Losses	215	2.11	240	2.36
	10,303	100.15	10,173	99.78

ECONOMICS

Important economic results are summarized below.

TABLE EATC-6

SUMMARY OF ECONOMICS - CASES EATC AND EXTC

	Case EXTC	Case EATC
PRODUCTION AT DESIGN CAPACITY		
Net Power, MW (1)	1,156.8	1,138.2
Overall Plant Heat Rate, Btu/kWh	8,813	8,958
TOTAL CAPITAL (2)		
Total Capital @ $1/MM Btu Coal, $1,000	944,563	940,628
Total Capital @ $1/MM Btu, $/kW	816	826
Total Capital @ $2/MM Btu Coal, $1,000	961,681	947,747
Total Capital @ $2/MM Btu, $/kW	831	841
AVERAGE COSTS OF SERVICES (2)		
Annual Cost @ $1/MM Btu Coal, $1000/yr	262,088	264,078
Per Unit @ $1/MM Btu Coal, mills/kWh	37.21	37.84
Annual Cost @ $2/MM Btu Coal, $1000/yr	327,280	329,271
Per Unit @ $2/MM Btu Coal, mills/kWh	46.47	47.18

NOTES

(1) At 100% operating factor
(2) Mid-1976 dollars and 70% operating factor

CAPITAL COST COMPARISONS

Having discussed each Texaco technology individually, it is of interest to compare capital cost breakdowns for all sections of each system to point out major differences as well as to identify those areas within each system which would benefit most from development of new and/or advanced technologies. Such a comparison is shown in Table D-2 (see Discussion of Results).

It can be seen from this table that coal handling costs for case EATC are slightly higher than EXTC due to the multiplicity of gasifier trains,. The oxidant feed system cost for the oxygen blown Texaco case (EXTC) is almost five times that for the air blown system. This is mainly due to the fact that the oxygen blown gasifiers require air compressors, an air separation plant, and oxygen compressors, whereas the air blown gasifiers require only booster air compressors. Both cases have fuel gas expanders included in the capital cost of the combined cycle. However, in Case EATC the booster compressor drivers are these fuel gas expanders, which artificially lowers the cost of its oxidant feed unit.

Costs for the gasification and ash handling sections of each plant indicate that air blown gasification would benefit greatly from reduced gasifier residence time or from development of larger sized gasifiers. Gas cooling section costs, however, indicate that the air blown system pays part of the price of low oxidant feed cost by requiring extra surface area in the costly gas cooling equipment to remove the sensible heat carried by nitrogen.

Process condensate treating costs for ammonia recovery for both of the Texaco systems are not included since ammonia recovery is not anticipated. The higher cost of acid gas removal in the air blown case reflects primarily the additional train required to handle the extra gas volume.

The cost of the steam, condensate and BFW system and support facilities for the air blown case are slightly higher due to higher water flows for this case.

Combined cycle costs are lower for the air blown case partially due to one less gas turbine and HRSG being required and partially due to a larger fraction of the total power being produced in the steam turbine. The two small steam turbine generators included in Case EATC represent 8.3 percent of the combined cycle plant investment.

The total capital requirements are about one percent higher and average annual cost of services nearly 0.6 percent higher in the air blown case.

Tables EATC-7 and EATC-8 give detailed breakdowns of plant investment, capital charges and working capital for both cases at 70 percent operating factor and $1.00/MM Btu and $2.00/MM Btu coal HHV cost. The accuracy of the plant investment estimates is judged to be ±25%. Since other capital charges and working capital are keyed to elements of plant investment, this accuracy is reflected in other capital figures as well. This should be kept in mind when comparing cases. The comparison between Texaco gasifier types should be somewhat more accurate since the same estimating methods were used in both cases.

For all units, the air blown plant investment is higher or equal to the oxygen blown case except for the oxidant feed and combined cycle units. The latter units are less expensive in this case because there are no oxygen plants or compressors, and because the number of gas turbine/HRSG units is reduced compared to Case EXTC. The coal handling unit, waste water treating and steam, condensate and boiler feed water units are slightly higher in the air blown case. The coal handling units are essentially the same. In both cases process condensate can be returned to the gasifier, reducing waste water treating requirements. The size of the steam, condensate and boiler feed water system in the air blown case is larger because of the greater amount of sensible heat removed in the process units. This is reflected in the higher investment requirements for these units in the air blown case.

The contingency shown under plant investment is divided into two parts. First is a 15 percent project contingency which is intended to cover additional equipment that would result from a more detailed design of a definitive project at an actual site. The second is a process contingency which is applied to unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. Historically, as a new technology develops from the conceptual stage to commercial reality, a variety of technical

TABLE EATC-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $1.00/MM BTU COAL - CASES EXTC AND EATC

	Case EXTC			Case EATC		
	$1,000 (1)	$/kW (2)	Percent	$1,000 (1)	$/kW (2)	Percent
PLANT INVESTMENT						
Coal Handling	22,061	19.07	3.56	22,500	19.77	3.72
Oxidant Feed	117,389	101.48	18.95	23,716	20.84	3.92
Gasification and Ash Handling	24,261	20.97	3.92	60,138	52.85	9.94
Gas Cooling	66,986	57.91	10.81	119,139	104.69	19.71
Acid Gas Removal and Sulfur Recovery	28,585	24.71	4.61	42,966	37.76	7.10
Waste Water Treating	-	-	-	-	-	-
Steam, Condensate and BFW	827	0.71	0.14	879	.77	.15
Support Facilities	55,205	47.72	8.91	57,284	50.34	9.47
Combined Cycle	304,156	262.94	49.10	278,130	244.40	45.99
Subtotal	619,470	535.51	100.00	604,752	531.42	100.00
Contingency	118,160	102.14		127,696	112.21	
Total Plant Investment	737,630	637.65		732,448	643.63	
ILLINOIS SALES TAX	16,656	14.40		16,272	14.30	
CAPITAL CHARGES						
Preproduction Costs	46,342	40.06		46,458	40.82	
Paid-up Royalties	3,688	3.19		3,662	3.22	
Initial Catalyst and Chemical Charge	515	0.44		1,285	1.13	
Construction Loan Interest	92,130	79.64		91,483	80.39	
Total Capital Charges	142,675	123.33		142,888	125.56	
DEPRECIABLE CAPITAL	896,961	775.39		891,608	783.49	
WORKING CAPITAL	47,602	41.15		47,782	41.99	
TOTAL CAPITAL	944,563	816.54		939,390	825.48	

NOTE

(1) Mid-1976 Dollars
(2) Based on 100% Operating Load Factor

TABLE EATC-8

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $2.00/MM BTU COAL - CASES EATC AND EXTC

	Case EXTC			Case EATC		
	$1,000 (1)	$/kW (2)	Percent	$1,000 (1)	$/kW (2)	Percent
PLANT INVESTMENT						
Coal Handling	22,061	19.07	3.56	22,500	19.77	3.72
Oxidant Feed	117,389	101.48	18.95	23,716	20.84	3.92
Gasification and Ash Handling	24,261	20.97	3.92	60,138	52.85	9.94
Gas Cooling	66,986	57.91	10.81	119,139	104.69	19.71
Acid Gas Removal and Sulfur Recovery	28,585	24.71	4.61	42,966	37.76	7.10
Waste Water Treating	-			-	-	-
Steam, Condensate & BFW	827	0.71	0.14	879	.77	.15
Support Facilities	55,205	47.72	8.91	57,284	50.34	9.47
Combined Cycle	304,156	262.94	49.10	278,130	244.40	45.99
Subtotal	619,470	535.51	100.00	604,752	531.42	100.00
Contingency	118,160	102.14		127,696	112.21	
Total Plant Investment	737,630	637.65		732,448	643.63	
ILLINOIS SALES TAX	16,656	14.40		16,272	14.30	
CAPITAL CHARGES						
Preproduction Cost	48,202	41.67		48,319	42.46	
Paid-up Royalties	3,688	3.19		3,662	3.22	
Initial Catalyst and Chemical Charge	515	0.45		1,285	1.13	
Construction Loan Interest	92,130	79.64		91,483	80.39	
Total Capital Charges	144,535	124.95		144,749	127.20	
DEPRECIABLE CAPITAL	898,821	777.00		893,469	785.12	
WORKING CAPITAL	62,860	54.34		63,041	55.40	
TOTAL CAPITAL	961,681	831.34		956,510	840.52	

NOTE

(1) Mid-1967 Dollars
(2) Based on 100% Operating Load Factor

problems which were not considered during the early stages of the development emerge. Solution to these problems generally results in an increase in the cost of the technology due to the need for more expensive materials of construction, more complex equipment specifications and sometimes the need for additional processing equipment. A total plant process contingency is arrived at by applying a separate contingency to individual process units based on their state of development and accumulating the results.

Table EATC-9 summarizes cost of services for both cases based upon coal charged at $1.00/MM Btu and $2.00/MM Btu HHV. Costs are compiled in accordance with criteria furnished by EPRI (Criteria Section). They are presented as averages for the life of the plants.

Operating labor requirements are a function of the number of units and trains. Requirements are shown below on a shift basis.

	Case EXTC	Case EATC
"A" Operators	5	5
"B" Operators	17	20
Foremen	2	2
Lab and Instrument Technicians	4	4
Operating manpower	28	31

Catalyst and chemical costs are primarily for chemicals consumed in the demineralizer, cooling tower, and boiler feed water treating. There are some minor costs associated with making up solution losses in the acid gas removal, and tail gas treating units and replacement of catalyst in the sulfur recovery unit. Chemical consumption costs are included for process condensate treating in the dry feed case.

The operating charges are slightly higher for the air blown case and occur mainly in utilities and investment ratioed operating costs.

Operating charges constitute about 44 percent of cost of services with coal at $1.00/MM Btu and nearly 55 percent at a coal cost of $2.00/MM Btu. For both cases, coal is the largest single operating charge.

The relationship as percentage is summarized below:

	Case EXTC		Case EATC	
Cost of Coal, $/MM Btu, HHV	1.00	2.00	1.00	2.00
Coal as % of Operating Charges	54.5	70.5	53.2	69.5
Coal as % of Total Cost of Services	23.9	38.2	23.7	38.0
Operating Charges as % of Total Cost of Services	43.8	54.1	44.5	54.7
Capital Charges as % of Total Cost of Services	56.2	45.9	55.5	45.3

TABLE EATC-9

COST OF SERVICES AT 70% OPERATING LOAD FACTOR - CASES EATC AND EXTC

	Case EXTC		Case EATC	
COAL COST, HHV	$1/MM Btu	$2/MM Btu	$1/MM Btu	$2/MM Btu
NET PRODUCTION (1)				
Net Power, MW	1,156.8	1,156.8	1,138.2	1,138.2
By-product Ammonia ST/SD	0	0	0	0
By-product Sulfur ST/SD	301	301	309	309
OPERATING CHARGES, $1000/YEAR				
Coal	62,522	125,044	62,522	125,044
Operating Labor	2,692	2,692	2,987	2,987
Catalyst and Chemicals	262	262	358	358
Utilities	1,354	1,354	1,589	1,589
Maintenance, Labor	7,882	7,882	8,480	8,480
Maintenance, Materials	11,822	11,822	12,721	12,721
Administrative and Support Labor	3,172	3,172	3,440	3,440
General and Administrative Expenses	6,344	6,334	6,880	6,880
Ash Disposal	245	245	245	245
Property Tax/Insurance	18,441	18,441	18,311	18,311
By-product, Ammonia	(0)	(0)	(0)	(0)
By-product, Sulfur	(0)	(0)	(0)	(0)
Total Operating Charges, $1000/Year	114,736	177,258	117,553	180,055
CAPITAL CHARGES, $1,000/YEAR				
Total Capital Charges	147,352	150,022	146,545	149,216
COST OF SERVICES				
Total, $1,000/Year	262,088	327,280	264,078	329,271
Per Unit Production, mills/kWh	37.21	46.47	37.84	47.18

NOTES

(1) At 100% Operating Load Factor.

APPENDIX A

COMBINED CYCLE SYSTEM DETAILS

GENERAL

For each of the coal gasification processes studied, a similar combined cycle system was selected. For cases reported previously[1], the combined cycle system was designed by Westinghouse Electric Corporation, Lester, Pennsylvania based on interface conditions between the fuel processing and power section supplied by Fluor. For EATC, Fluor performed similar calculations based on previous Westinghouse data. Equipment design limitations imposed here were the same as those imposed by Westinghouse in the earlier work.

Each of the combined cycle systems, including the one reported here for EATC, consists of a set of gas turbines, heat recovery steam generators (HRSG), steam turbine and auxiliary power equipment to support the respective coal gasification processes. Approach temperatures, pressure losses and blade loadings used in the calculations all reflect current utility application criteria for lowest cost of power. Equipment performance criteria and overall combined cycle parameters are projected to represent the expected state of the art with 1985 delivery.

Case EATC also includes two low pressure steam turbine generators for recovery of low temperature process heat. These turbines contribute 1% of the electric power generated.

A summary of the calculated power output for the power block equipment and heat loads rejected to the station cooling tower for both the oxygen blown (EXTC) and the air blown (EATC) Texaco based systems is presented in Table CS-1. The power output is calculated at the generator terminals without margins for design or manufacturing tolerances. The calculated power outputs include approximately 2.0 percent deduction for mechanical and electrical losses which include lube and seal oil pumps, plus radiation losses of 3 Btu/lb/sec of flue gas.

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

TABLE CS-1

POWER BLOCK PERFORMANCE SUMMARY

GENERATION	EXTC	EATC
Gas Turbine, kW	744,470	548,162
Steam Turbines, kW	448,390	633,620
Total, Power Block, kW	1,192,860	1,181,782
HEAT REJECTION		
Process Cooling, M^2 Btu/Hr	379.8	868.3
Process Cooling Absorbed in C.C., M^2 Btu/hr	338.1	441.3
Process Cooling Rejection to Tower, M^2 Btu/hr	41.7	427.0
Power Block Heat Rejection,[1] M^2 Btu/hr	3,806.2	4,385.0
Total Heat Rejection to Tower, M^2 Btu/hr	3,847.9	4,812.8

Note:

1. Includes mechanical and electrical losses.

Gas Turbine

The fuel gas from the gasification process is delivered to the gas turbine trip valve at a pressure of 280 psig. In Case EATC, high pressure air is required by the coal gasification process and the source of this air is the gas turbine compressor discharge. The compression ratio of the gas turbine was selected to result in a pressure at the air extraction port of 225 psig, with ambient site conditions of 14.4 psia and 88°F dry bulb. The gas turbine employs a ceramic thermal barrier coating on the turbine vanes and blades for operation at 2400°F nominal inlet temperature.

Steam Conditions

Steam conditions used for the combined cycle system are:

Turbine Throttle	1450 psig
	900°F superheat
	1000°F reheat
Condenser	2.5" Hg. abs.

Heat Recovery Steam Generator (HRSG) Conditions

Steam production of the HRSG was calculated using a flue gas stack temperature of approximately 272°F. On the basis of demineralized boiler feedwater, the steady state boiler blowdown was assumed to be zero. The low pressure flash gas, available from the gasification process, has been burned as supplementary fuel in the HRSG.

Process Interface

The pertinent data regarding pressure, temperature and compositions of the fluids and their flow rates to the power block are based on the design of the process units for this case. Heat integration between the process units and the power block is considered whenever possible for the maximum utilization of energy.

Auxiliary Equipment

Auxiliary loads in excess of 5000 HP in both the gasification system and power block are steam-driven, including the boiler feed pumps.

POWER BLOCK

Gas Turbine-Generator Unit

Gas Turbine (50-1-GT-1)

The thermodynamic design of the gas turbine for each gasification cycle is different, taking into consideration the two major interface variations - available fuel gas and process air bleed requirements. The remaining operational constraints such as 2400°F turbine inlet temperature, site conditions of 88°F and 14.4 psia and a combustor shell pressure of 225 psig were also applied to Case EATC.

Overall engine performance was estimated by Fluor based on previous Westinghouse information. The performance included both inlet air and exhaust duct losses to account for pressure drops through air silencers, ducting, afterburner and HRSG heat recovery sections.

An air-to-water heat exchanger was used to cool the compressor discharge air for cooling the turbine rotating hot parts. The heat rejected to this heat exchanger was recovered and integrated with the intermediate pressure section of the HRSG.

Generator (50-1-G-1)

Each gas turbine drives a suitably rated, 0.9 power factor (pf), 0.58 short circuit ratio (SCR), three-phase, 60 hertz, 13.8 kV, 3600 rpm outdoor type, hydrogen cooled (30 psig) generator. These use water coolers for 95°F or lower water and direct connected, suitably rated enclosed air-cooled brushless exciters with a permanent magnet generator.

A summary tabulation of gas turbine performance and generator output is given in Table CS-2 as provided by Westinghouse for EXTC and as used or calculated by Fluor for EATC.

TABLE CS-2

GAS TURBINE PERORMANCE SUMMARY

DESCRIPTION	EXTC WESTINGHOUSE	EATC FLUOR
Compressor Inlet Air Duct Loss, Inches H_2O	4.0	4.0
Compressor Disch. Pressure, psia	239.1	239.1
Compressor Disch. Temp., °F	857	857
Turbine Inlet Pressure, psia	229	229
Rotor Cooling Air, % of Inlet Flow	2.9	2.9
Rotor Coolant Temperature °F	350	350
Turbine Exhaust System, Inches H_2O	20.5	20.5
Compressor Air Flow, lb/sec	4,818.86	4,174.89
Air to Process, lb/sec	0.0	1,046.56
Fuel Flow, lb/sec	399.14	1,196.77
Turbine Exhaust Temp., °F	1,140	1,135
Rotor Cooling Air Cooler Duty, MM Btu/hr	64.9	57.1
Power Output, kW[1]	744,470	548,162
Flash Gas Fuel Flow, lb/sec	5.08	5.16
Total Exhaust Gas Flow, lb/sec	5,223.10	4,330.25
Exhaust Gas Temp., Into HRSG, °F	1,147	1,139

Note:

1. At generator terminals

Steam Cycle Selection

HRSG

An HRSG, 51-1-B-1, is coupled with each gas turbine, 50-1-GT-1, to recover heat from the turbine exhaust gases. Flash gas from the process plant is also burned in the gas turbine exhaust before entry to the HRSG.

For Case EATC the HRSG generates saturated steam at one pressure level, high pressure (HP) at 1520 psia. An approach temperature greater than 30°F (temperature of gas leaving - saturation temperature of steam) was used in calculating the steam generated in the evaporator section. A 30°F approach temperature is in line with current HRSG design practice.

A typical arrangement of the heat recovery sections for an HRSG has been shown previously.[1] In the direction of exhaust gas flow the HRSG heat recovery sections for this case are as follows:

. Air heater	51-1-B-1:E-0
. Superheater	51-1-B-1:E-1
. Reheater	51-1-B-1:E-2
. Fuel Gas Heater Two	51-1-B-1:E-3
. HP Evaporator	51-1-B-1:E-4
. Economizer One	51-1-B-1:E-5
. Economizer Two	51-1-B-1:E-7
. Economizer Three	51-1-B-1:E-9A
. Economizer Four	51-1-B-1:E-9B

The high pressure saturated steam generated in the HP evaporator is heated to 900°F in the superheater. HP steam available from the process is combined with the HP steam from 51-1-B-1:E-1 after the superheater.

Saturated intermediate pressure (IP) steam produced in various process units is combined with cold reheat steam from the high pressure steam turbine, 51-T-1, and superheated to 1000°F by passing through the reheater, 51-1-B-1:E-2.

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

Low pressure (LP) steam generators located in process units supply deaerating steam to a tray type deaerator, 51-DA-1, and steam for two parallel 5.5 MW steam turbine generators. The deaerator operates at approximately 30 psia.

The boiler feedwater (BFW) from the deaerator at two pressure levels, HP and MP, is first preheated to the MP steam saturation temperature (357°F) in Economizers Three and Four. The LP BFW leaves the HRSG and flows to various process steam generators. The HP BFW is heated to the HP steam saturation temperature (595°F) in Economizers One and Two. A portion of this water is withdrawn, pumped through Fuel Gas Heater One, raising the fuel gas temperature to 550°F, and returns to the inlet of Economizer Two at about 347°F. Almost three quarters of the HP BFW is withdrawn and flows to the gasification unit for generation of 1520 psia, 900°F superheated steam. The balance of the BFW is evaporated in the HP steam recovery section at 598°F and flows to the HP superheater, where its temperature is raised to 900°F.

Each HRSG is provided with its own HP steam drum and corresponding BFW circulation pump.

The HRSG exhaust gas (stack) temperature of approximately 272°F allows the gas side surface of the economizers to operate safely above the sulfur dew point of the exhaust gas. The performance of the HRSG calculated by Fluor for case EATC is summarized in Table CS-3, next to data from Westinghouse for Case EXTC.

Steam Turbine - Generator Units

Steam Turbine (51-T-1 and 2)

A single steam turbine system consisting of HP and IP ends (51-T-1) and LP end (51-T-2) has been used for Case EATC. The turbine selected for the power block is a conventional tandem compound, reheat machine.

The HP end of 51-T-1 receives superheated HP steam at 1450 psig, 900°F from the HRSGs and the process and exhausts to the IP steam header operating at approximately 445 psig. The IP steam available from the process is combined with

TABLE CS-3

HRSG PERFORMANCE SUMMARY

	EXTC	EATC
Exhaust Gas Flow, lb/sec	5,223.10	4,330.25
HP EVAP., SH & RH SECTIONS		
Exhaust Gas Temperature In, °F (1)	1,147.0	1,139.0(2)
SH Temperature Out, °F	900.0	900.0
SH Pressure Out, psig	1,450.0	1,450.0
SH Enthalpy Out, Btu/lb	1,431.0	1,430.0
Sat. Steam Evap. lb/sec	362.3	82.0
Sat. Steam from Process, lb/sec	588.3	0.0
SH Outlet Flow, lb/sec	950.6	82.0
SH Duty, MM Btu/hr	896.60	77.5
HP Drum Temperature, °F	598.0	598.0
HP Drum Pressure, psia	1,520.0	1,520.0
HP Drum Water Enthalpy In, Btu/lb	614.0	610.0
HP Evap. Duty, MM Btu/hr	724.0	164.7
RH Enthalpy In, Btu/lb	1,303.9	1,296.8
RH Temperature Out, °F	1,000.0	1,000.0
RH Pressure Out, psig	385.0	385.0
RH Enthalpy Out, Btu/lb	1,523.0	1,523.0
RH Flow, lb/sec	1,016.6	1,061.2
RH Flow to Process, lb/sec	301.6	0.0
RH Duty, MM Btu/hr	801.9	866.4
Fuel Gas Heater Flow, lb/sec	399.14	1,196.77
Fuel Gas Heater Duty, MM Btu/hr	204.2	570.3
Air Heater Flow, lb/sec	-	1,046.6
Air Heater Duty, MM Btu/hr	-	194.4
Exhaust Gas Temperature Out, °F	630.0	688.2
ECONOMIZER NO. 1 SECTION		
Water Enthalpy In, Btu/lb	440.0	438.0
Water Flow, lb/sec	1,077.6	1,223.7
Water Flow to Process, lb/sec	588.3	863.9
Water Flow to Fuel Gas Heater, lb/sec	127.0	277.8
Duty, MM Btu/hr	675.0	756.0
IP EVAPORATOR SECTION		
IP Drum Temperature, °F	459.0	-
IP Drum Pressure, psia	460.0	-
IP Steam Enthalpy Out, Btu/lb	1,205.0	-
IP Steam Evap., lb/sec	5.2	-
IP Evap. Duty, MM Btu/hr	14.4	-
IP Steam from Air Cooler, lb/sec	17.2	-
IP Steam to (from) Process, lb/sec	(62.6)	-
IP Steam to Cold RH, lb/sec	85.0	-
Water Enthalpy In, Btu/lb	440.0	-
Exhaust Gas Temperature Out, °F	490.0	503.0

(1) Includes flash gas combustion heat
(2) Total radiation losses from HRSG would reduce this to 1128°F.

TABLE CS-3 (Continued)

ECONOMIZER NO. 2 SECTION	EXTC	EATC
Water Enthalpy In, Btu/lb	222	321.0
Water Flow, lb/sec	1,145.0	1,223.7
Outlet Flow to Process, lb/sec	62.6	0.0
Duty, MM Btu/hr	898.6	514.6
Exhaust Gas Temperature Out, °F	300.0	374.9
LP EVAPORATOR & DA SECTION		
LP Drum Temperature °F	246.0	-
LP Drum Pressure, psia	28.0	-
Cond. Flow In, lb/sec	1,061.4	-
Cond. Enthalpy In, Btu/lb	165.5	-
Process Flows In, lb/sec	160.3	-
FW Flow to Process, lb/sec	54.1	-
Duty, MM Btu/hr	133.5	-
Exhaust Gas Temperature Out, °F	272.0	-
ECONOMIZERS NO. 3 & 4 SECTIONS		
HP Water Enthalpy In, Btu/lb	-	225.0
HP Water Flow, lb/sec	-	945.9
LP Water Enthalpy In, Btu/lb	-	221.0
LP Water Flow, lb/sec	-	216.3
LP Water Flow to Process, lb/sec	-	216.3
HP Duty, MM Btu/hr	-	328.7
LP Duty, MM Btu/hr	-	76.4
Exhaust Gas Temperature Out, °F	-	272.0
DEAERATOR		
LP Steam Enthalpy in, Btu/lb	-	1,155.7
LP Steam Flow, lb/sec	-	49.2
Cond. Enthalpy in, Btu/lb	-	180.4
Cond. Flow in, lb/sec	-	1,188.3
Vent Steam Enthalpy out, Btu/lb	-	1,164.3
Vent Steam Flow, lb/sec	-	0.3
LP Water Enthalpy, Btu/lb	-	219.0
LP Water Flow, lb/sec	-	1,237.2

HP turbine exhaust and reheated to 1000°F in the HRSG reheaters (51-1-B-1:E-2), and then flows to the IP end of 51-T-1. The inlet and exhaust conditions for the IP end of 51-T-1 are 385 psig, 1000°F and 110 psig respectively.

The LP end (51-T-2) is a condensing type unit receiving steam at 110 psig and exhausting at 2-1/2" Hg abs. The surface condenser, 51-E-8, associated with 51-T-2 is designed for cooling water (CW) flow in two tube side passes with 80°F CW inlet temperature and 20°F CW temperature rise.

Generator - Exciter (51-1-G-1)

The steam turbine (51-T-1 and 2) drives a suitably rated, 0.9 pf, 0.58 SCR, 3-phase, 60 hertz, 24.0 kV, 3600 rpm outdoor type, hydrogen inner-cooled generator with water coolers for 95°F or lower water and direct connected suitably rated enclosed air-cooled brushless exciter with permanent magnet generator.

Steam Turbines (51-T-3A and B) and Generators

Two steam turbines consisting of 50 psig inlets and 15 psig admittance steam have been provided to recover additional low temperature process heat. The exhaust conditions are 2-1/2" Hg absolute.

A summary tabulation of steam turbine performance and generator ouput is given in Table CS-4.

TABLE CS-4

STEAM TURBINES

PERFORMANCE SUMMARY

HP ELEMENT (T1)	EXTC	EATC
Throttle Conditions:	1,450 psig/900°F TT	
Steam Enthalpy In, Btu/lb	1,431.0	1,430.0
Throttle Flow from HRSG, lb/sec	950.6	82.0
Throttle Flow from Process, lb/sec	0.0	863.9
Total Throttle Flow, lb/sec	950.6	945.9
Exhaust Flow to Process, lb/sec	0.0	0.0
Exhaust Enthalpy, Btu/lb	1,313.0	1,308.0
IP ELEMENT (T1)		
Inlet Conditions:	385 psig/1,000°F TT	
Inlet Enthalpy, Btu/lb	1,523.0	1,523.0
Inlet Flow, lb/sec	715.0	1,061.2
Exhaust Enthalpy, Btu/lb	1,383.0	1,387.0
Exhaust Flow to Process, lb/sec	18.7	0.6
Exhaust Flow to Pump Turbines, lb/sec	0.0	25.9
LP ELEMENT (T2)		
Turbine Exh. Flow, lb/sec	715.3	1,093.1
Exhaust Enthalpy, Btu/lb	1,060.0	1,060.0
Total Flow to Condenser, lb/sec	759.0	1,190.0
Power Output, kW at Gen. Terminals	448,390.0	622,146.0
AUXILIARY TURBINES (T3A & B)		
Inlet Conditions	-	50 psig/sat'd.
Inlet Enthalpy, Btu/lb	-	1,180.0
Inlet Flow, lb/sec	-	36.7
Admittance Conditions	-	15 psig
Admittance Enthalpy, Btu/lb	-	1,156.0
Admittance Flow, lb/sec	-	25.8
Total Flow to Condenser, lb/sec	-	62.5
Power Output, kW at Gen. Terminals	-	11,474.0

EQUIPMENT STATE OF THE ART

Information given here is repeated from previous work[1] and was originally supplied by Westinghouse.

Gas Turbine

The major equipment assembled and described as part of the power block combined cycle for this study, contain some operating parameters uncommon to current industry practice. These uncommon areas are:

- gas turbine compressor pressure ratio
- high temperature turbine operation

Although not in current commercial production, these areas are not beyond the state of the art for 1985 base load operation.

Gas Turbine Compressor Pressure Ratio

At 2400°F turbine inlet temperature, the specified high compressor pressure ratio of approximately 17 to 1 on an 88°F ambient day approaches an equivalent of 19 to 1 at ISO conditions (59°F, 14.7 psia). This is much higher than current design practice (12 and 14 to 1) by Westinghouse on large single spool axial flow compressors incorporated into single shaft, gas turbines. Single spool engines rated at 20 MW and 17 to 1 are commercially available today. However, these units employ several (6-7) stages of variable geometry compressor stators at the inlet end of the compressor.

Performance studies on combined cycles operating with gas turbine inlet temperatures in the 2400°F range have shown the optimum pressure ratio to be near 14 to 1. Because of this no one has yet undertaken development of large single spool gas turbines having fixed (maximum of two variable stage) compressor geometry for these higher compression ratios.

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

High Temperature Turbine Operation

Gas turbines for operation at base load with an inlet temperature of 2400°F and fitted with hot parts having thermal barrier coatings are not commercially available at the present time.

To date, test aircraft type engines fitted with plasma-sprayed ceramic coated turbine blades have been operated successfully by NASA. On this basis, 2400°F operation with thermal barrier coatings could be considered as being commercially available in the 1981 to 1985 period with appropriate development plans.

HRSG

The equipment in this section of the power block is commercially available.

Steam Turbines

Although the selected throttle steam conditions of 1450 psig/900°F/1000°F reheat for the large steam turbine present no problem to the state of art, current industry practice with machines in the size range of this study would have throttle pressures of 1800 and 2400 psig. The two small low pressure steam turbines are commercially available.

APPENDIX B

AREA AND UNIT NUMBERING

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

AREA/UNIT NUMBERING SYSTEM

AREA	AREA DESCRIPTION	UNIT	UNIT DESCRIPTION
10	Feed Systems	10	Coal Preparation
		11	Oxidant Feed
20	Onsite Units	20	Gasification and Ash Handling
		21	Gas Cooling, Char Recovery and Particulate Removal
		22	Acid Gas Removal
		23	Sulfur Recovery and Tail Gas Unit
		24	Process Condensate Treating
30	Utility Systems	30	Steam, Condensate and Boiler Feedwater System
		32	Cooling Water System
		33	Plant and Instrument Air System
		34	Potable and Utility Water
		35	Fuel Gas System
		36	Nitrogen System
40	Offsite Facilities	40	Effluent Water Treating
		41	Flare System
		42	Firewater System
		43	Buildings
		44	Railroad Loading and Unloading
		45	Electrical Distribution
50	Combined Cycle System	50	Gas Turbine and Power Generation
		51	Heat Recovery and Power Generation