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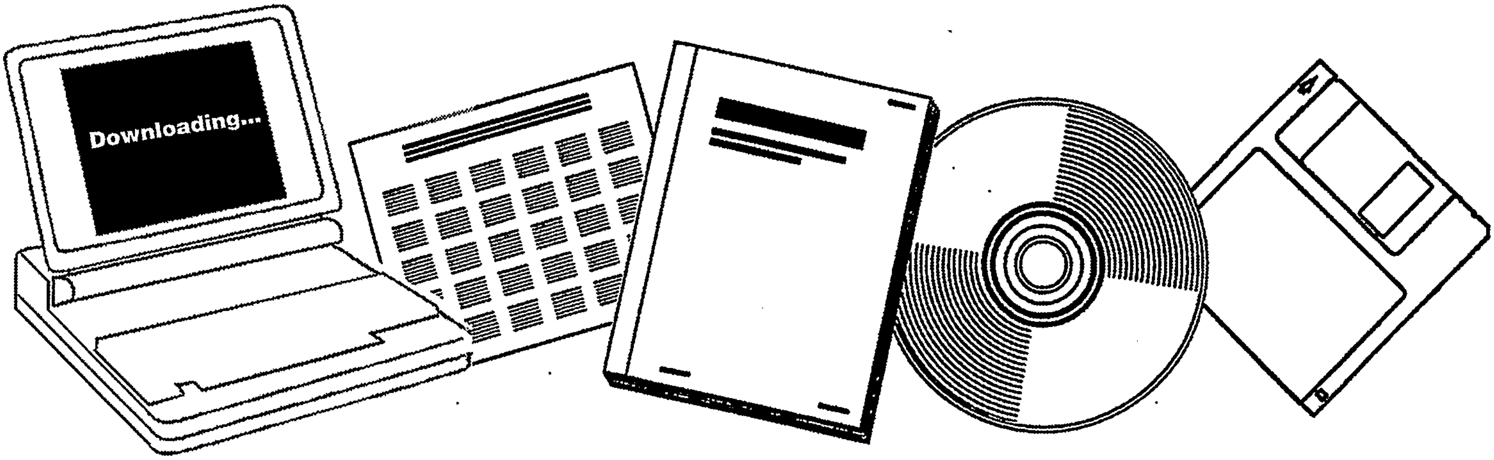
ECONOMIC EVALUATION OF THE COPRODUCTION OF METHANOL AND ELECTRICITY WITH TEXACO GASIFICATION-COMBINED-CYCLE SYSTEMS. FINAL REPORT

FLUOR ENGINEERS AND CONSTRUCTORS, INC.
IRVINE, CA

JAN 1982



Economic Evaluation of the Coproduction of Methanol and Electricity With Texaco Gasification-Combined-Cycle Systems

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Prepared by

FLUOR ENGINEERS AND CONSTRUCTORS, INC.
3333 Michelson Drive
Irvine, California 92730

Principal Investigators
R. E. Brown
R. C. Delaney
W. W. Hsu
R. H. Ravikumar
S. C. Smelser
R. M. Stock

Prepared for

Electric Power Research Institute
3412 Hillview Avenue
Palo Alto, California 94304

EPRI Project Managers
E. L. Force
M. J. Gluckman

Engineering and Economic Evaluations Program
Advanced Power Systems Division

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ABSTRACT

This study evaluates process and cost reduction options for the production of methanol fuel for peaking and intermediate load generating units. Methanol is coproduced with electricity in this study from medium BTU gas generated in an oxygen blown Texaco-based Gasification Combined-Cycle (GCC) Plant firing Illinois No. 6 coal. The Chem Systems liquid phase methanol process is used in a "once-through" configuration where the unconverted reactor effluent is used for gas turbine fuel.

The cost of methanol was determined by first calculating the revenue requirements of a Texaco-based GCC plant without methanol coproduction. The methanol coproduction case was then credited with those electricity revenues and the remaining revenues required (in excess of the electricity credit) then represented the revenue requirement for methanol coproduction.

The gasification facilities are similar in most respects to Case EXTC-79 published in EPRI Report No. AP-1624, except the gasification plant pressure has been increased from 600 psig used previously to 1000 psig in order to match the methanol synthesis requirements. This change in operating pressure was found not to be a significant factor on system efficiency or capital cost.

Coproduction of methanol in a Chem Systems "once-through" plant configuration shows a potential 30 percent first year savings over the nonregulated, dedicated coal-to-methanol plant employing currently commercial technology. The "once-through" scheme described in this report exists only at the experimental level.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This final report, Economic Evaluation of the Coproduction of Methanol and Electricity with Texaco Gasification Combined-Cycle Systems, presents a detailed assessment of the potential financial benefits to be derived from utility ownership of a baseloaded, Illinois No. 6 coal gasification combined-cycle (GCC) power plant coproducing methanol by the once-through technique.

Within recent years, it has become apparent to the electric utility industry that acquisition of critically needed petroleum derived liquid fuels and natural gas could become exceedingly difficult by the end of this decade. Political uncertainties in the Middle East coupled with rapidly escalating liquid and gaseous fuel prices have made the vulnerability of the electric utility industry clear.

The potential difficulties associated with the acquisition of petroleum derived liquids for power generation have stimulated interest on the part of the electric utility industry in synthetic liquid fuels produced from coal. It seems clear, however, from actions taken to date, that coal derived liquid fuels production will be dominated by the existing petroleum and chemical industries. In particular, it appears unlikely that the electric utility industry will participate to any major extent in the capitalization or the operation of these large and highly capital intensive liquid fuel production facilities. If synthetic liquid fuels are to be produced by the private nonregulated sector, they will be sold at prices that are competitive with current prices for petroleum derived fuels.

All of this leaves the electric utility industry in a particularly vulnerable and uncomfortable position with respect to future supplies of liquid fuels. The question that must be addressed is how can the utility industry control its own destiny by ensuring a long term supply of clean liquid fuels at a reasonable cost without having to make major capital investments in facilities producing large fractions of their output as nonutility fuels.

The purpose of this study was to investigate the potential economic advantages to be associated with the coproduction of relatively small quantities of methanol by the once-through technique in a baseloaded gasification combined-cycle power plant. It is important to point out the fact that coproduction of methanol and electricity is not being presented as the only solution to the liquid fuel supply problem. It should rather be considered as one option that potentially appears to offer attractive economics, as well as supply security.

PROJECT OBJECTIVES

The major objectives of this study were:

- To determine the cost of producing once-through methanol using the Chem Systems liquid phase synthesis reactor in conjunction with a Texaco-based gasification combined-cycle power plant employing currently available (2,000°F) gas turbines.
- To compare the cost of once-through methanol produced by a regulated electric utility company with the potential selling price of methanol produced in a dedicated coal-to-methanol plant owned and operated by a nonregulated energy company.

PROJECT RESULTS

Two detailed plant designs were performed by Fluor. The first design was for a standard Texaco-based GCC plant with no methanol production capability. The second design was of a similar Texaco-based GCC power plant incorporating a Chem Systems liquid phase methanol synthesis system. In the once-through concept employed, all of the desulfurized fuel gas is passed through the methanol synthesis reactor without pre-shifting, CO_2 removal or recycle of the unconverted gas. By this procedure, only a small fraction of the synthesis gas is converted to methanol (i.e., 7% - 17%). The unconverted gas, instead of being recycled to the methanol reactors, is sent to the combined-cycle power plant.

The cost of producing once-through methanol was determined by crediting the revenue requirements for electricity production (determined from the first GCC design with no methanol production capability) to the total revenue requirements of the coproduction plant, thereby determining the incremental revenues required for methanol production. Anticipated required selling prices for methanol produced in a dedicated Texaco-based coal-to-methanol plant owned and operated by a nonregulated producer are based on a recent Fluor design of such a facility (EPRI AP-1962, August, 1981).

All financial parameters, economic criteria, detailed economic results and sensitivity studies are presented in Section 7 of this report. All readers interested in understanding the results are urged to study Section 7. A brief summary of the financial results is presented in the following table:

	Regulated Utility Owned Once-Through Methanol/Electricity Coproduction Plant		Nonregulated Company Owned Dedicated Coal-To-Methanol Plant[#]	
Net System Power, MW	1,106.52		0	
Methanol Produced, $FOEB^{\delta}$/day	10,520		36,154	
Efficiency of Methanol Production, % of Coal HHV	68.80		57.86	
	Current Dollars[*]	Mid-1980 Dollars	Current Dollars[*]	Mid-1980 Dollars
Total Capital Requirement for 1990 Start-Up, $\$/FOEB^{\delta}$/day	79,545	32,165	103,142	41,666
Methanol Cost/Price				
First Year (1990) $\$/10^{6}$ BTU	15.18	5.58	21.41	7.87
Fifth Year (1994) $\$/10^{6}$ BTU	19.10	4.80	31.35	7.87
Tenth Year (1999) $\$/10^{6}$ BTU	26.83	4.18	50.49	7.87
Twentieth Year (2009) $\$/10^{6}$ BTU	60.30	3.62	130.96	7.87
Levelized $\$/10^{6}$ BTU	25.24	4.32	36.26	7.87

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

[*] Assuming 10%/year inflation.

[#] Assuming a minimum required after tax return on common equity of 20%. If this after tax return requirement is increased to 30%, the mid-1980 required selling price for dedicated methanol would increase to $\$12.72/10^{6}$ BTU.

Comparing the cost of producing methanol in a utility owned once-through methanol/electricity coproduction plant with the anticipated selling price of methanol produced in a nonutility owned dedicated coal-to-methanol facility results in a number of interesting conclusions:

- The first year cost of once-through methanol has the potential to be 30% lower than the expected selling price of methanol produced by a nonregulated company. These lower production costs are due primarily to the increased efficiency and lower unit capital requirements of once-through methanol as a result of eliminating shift conversion, CO_2 removal and gas recycle. Such a saving translates into a first-year saving of $50 million (1980 dollars) for a utility consuming 10,000 bbl/day of liquid fuel.

- After the first year of operation, the cost of methanol produced by a utility in a once-through methanol/electricity coproduction plant decreases (in constant dollars) with time from $\$5.58/10^6$ BTU in 1990 to $\$3.62/10^6$ BTU (both in mid-1980 dollars) in 2009. The nonutility produced methanol, however, will, at best, maintain its constant dollar price of $\$7.87/10^6$ BTU for the twenty year period. If liquid fuels escalate in price at a rate higher than the general inflation rate, the constant dollar $\$7.87/10^6$ BTU will increase proportionately.

- The average (levelized) constant dollar cost of the coproduced methanol ($\$4.32/10^6$ BTU) represents a saving of 45% over the average constant dollar selling price of $\$7.87/10^6$ BTU for nonutility produced methanol. This translates into an annual average saving of $78 million (1980 dollars) for a utility consuming 10,000 bbl/day of liquid fuel.

The final conclusion to be derived from this analysis is that the potential benefits to the utility industry to coproducing once-through methanol and electricity could be extremely large. It is critical to keep in mind, however, the fact that a Texaco-based Chem Systems once-through methanol coproduction plant described in this study <u>is not yet ready to be commercially deployed</u>. The Texaco coal gasification process has only been operated at the 150 ton/day scale. Demonstration of this technology at large scale (i.e., the 1,000 ton/day scale of the Cool Water Coal Gasification Project) is essential prior to being able to consider full scale commercial operation. Similarly, the Chem Systems once-through methanol synthesis process has only been conducted in small scale laboratory equipment. Large scale demonstration of this technology is required before utilities can consider investing in commercial once-through methanol coproduction plants. A possible alternative that EPRI is currently investigating is to determine if one of the existing commercial methanol synthesis processes can be modified to operate in the once-through mode.

Michael J. Gluckman, Project Manager
Engineering and Economic Evaluations
Advanced Power Systems Division

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SUMMARY

This study continues the evaluation of Texaco coal gasification, combined-cycle (GCC) power plants and examines process options and costs for coproducing methanol.

Oxygen-blown coal gasification systems such as the Texaco process produce a medium BTU gas which can be utilized as raw synthesis gas feedstock for a methanol plant as well as a gaseous fuel for use in an integrated GCC power plant. The latter option has been investigated previously in a series of engineering and economic evaluations. Methanol, from an environmental viewpoint, is an attractive fuel option for use by the U.S. electric power industry in both peaking and intermediate load generating units. However, high projected costs for coal-derived methanol are a potential hinderance to future use of this desirable fuel. A need therefore exists to examine process options which could lower the cost of coal-derived methanol. One such option is to produce methanol as a by-product in an integrated GCC plant.

The objective of this study is to determine the economic potential of "once through" coproduction of methanol using the Chem Systems liquid phase methanol process in an integrated GCC plant. In this configuration the entire gas make from the gasification system after sulfur removal to protect the methanol catalyst flows through the methanol reactor. The unconverted gas leaving the reactor is used as fuel for the gas turbines. This process configuration eliminates the necessity for shift conversion, excess CO_2 removal and recycle of unconverted synthesis gas typical of a coal-derived methanol plant.

The Chem Systems process is ideally suited to the "once-through" configuration because of its high achievable single pass conversion to methanol and its ability to absorb the heat released by the synthesis reaction by an inert oil.

The plant design, excluding the methanol synthesis unit, utilizes an integrated Texaco based GCC system using a currently available (2,000°F) gas turbine. It is similar in most respects to Case EXTC-79 published in EPRI Report No. AP-1624. The major difference is the gasification plant pressure which has been increased

from the 600 psig used in EXTC-79 to an average pressure of 1000 psig in order to match methanol synthesis requirements.

The process equipment used in each plant design consist primarily of commercially demonstrated units and processes with the following exceptions:

- The gasifiers employed in this study are extensions of existing technology with capacities and slurry concentrations higher than those currently proposed for the first generation of Texaco-based gasification systems.
- The Chem Systems process, chosen for the methanol plant in this study, is in the early stages of development. Research activities are being focused on abrasion-resistant, active, low-cost catalyst, and switching to a smaller catalyst size.
- The COS hydrolysis unit located upstream of the acid gas removal unit is yet to be proven on a commercial scale.
- The centrifugal oxygen compressor supplying the oxidant at 1120 psig is believed to be within state of the art; however, it has been commercially operated only up to 950 psig.
- Equipment suitable for high temperature gasifier effluent heat recovery requires further development work, especially in handling entrained ash.

The costs of electricity and methanol for all cases (with one exception to be noted later) evaluated in this study were calculated on the basis of criteria outlined in Table S-1 which are a new set of financial criteria, to be used in future studies for facilities owned, operated and financed by a regulated private utility. The split of revenue requirements between products for a multiproduct plant can be performed in an infinite number of ratios. The methodology employed in this study to determine revenue requirements for the methanol coproduction plant was developed by the Advanced Power Systems Division of EPRI. Here the revenue required for electricity production from a Texaco-based GCC power plant without methanol coproduction was calculated. The methanol coproduction plant was then credited with those revenues calculated for GCC plant without coproduction. The remaining revenue required (in excess of the electricity credit) then represents the revenue requirement for the methanol coproduct.

Table S-1

FINANCIAL CRITERIA USED FOR INVESTOR OWNED UTILITY REVENUE REQUIREMENT CALCULATIONS

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

The performance and economic results of the Texaco-based GCC and methanol coproduction plants developed in this study were compared to Case EXTC-79 to assess the impact of gasifier operating pressure. Additionally, for comparative purposes, the economic results were compared to a previously prepared conventional coal fired steam plant with flue gas desulfurization. In order to assess the cost of methanol produced in the "once-through" mode with the alternative of producing methanol from coal in a dedicated mode, the results were compared to a recently prepared cost estimate of a dedicated coal-to-methanol plant (based on Texaco gasification of Illinois No. 6 coal and the ICI methanol synthesis process). The cost of methanol produced by the latter case was calculated using nonregulated company owned economic analyses, the most probable method of producing and distributing liquid fuel from a dedicated facility.

The system performances for the 1000 psig Texaco-based GCC plant, designated Case A2, and the "once-through" methanol coproduction plant, Case B2, are compared in Table S-2 to the Texaco-based GCC plant (EXTC-79) published in EPRI Report No. AP-1624. This summary shows that an increase in gasifier pressure has a slight advantageous impact on overall system efficiency, 36.3 percent versus 37.0 percent. Inclusion of the methanol synthesis unit decreases the net system power from 1,106.5 MW to 810.3 MW (26.8 percent) due to the conversion of the fuel (synthesis) gas to methanol. The efficiency of converting coal to methanol is 68.8 percent (see page 2-1 for verification) for the "once-through" process configuration which increases the overall system efficiency to 45.5 percent.

A summary of the capital requirements and costs of electricity and methanol developed in this study is presented in Table S-3. Included in this table are the results of previous studies on Texaco-based GCC dedicated coal to methanol and conventional coal-fired steam plants. One assumption made in the economic analysis of the methanol producing cases is that the facility would operate at a 90 percent capacity factor while the GCC and conventional coal-fired steam plants operate at a 70 percent capacity factor. EPRI has confirmed this differential by performing an economic dispatch study for the three (GCC, conventional coal-fired steam and coproduction) types of plants. Availability analyses have shown GCC systems to be capable of the 90 percent on stream factor while historical records show that conventional coal-fired steam plant availability to be approximately 70 percent.

Table S-2

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

CASE DESCRIPTION	Conventional Texaco Based GCC Power Plant With No Methanol Production	Texaco GCC Plant With Once-Through Methanol Coproduction	High Efficiency Oxygen-Blown Texaco GCC System Power Only (EPRI AP-1624)
CASE DESIGNATION	A2	B2	EXTC-79
GASIFICATION SYSTEM			
Coal Feed Rate, lbs/hr m.f.	798,333	798,333	806,667
Oxygen/Coal Feed*, lb/lb m.f.	0.858	0.858	0.8639
Oxidant Temperature, °F	287	287	300
Slurry Water/Coal Ratio, lb/lb m.f.	0.503	0.503	0.503
Gasification Section Average Pressure, psig	1,000	1,000	600
Crude Gas Temperature, °F	2,300-2,600	2,300-2,600	2,300-2,600
Crude Gas HHV (Dry Basis), BTU/SCF**	281.1	281.1	282.8
POWER SYSTEM			
Temperature of Fuel Gas to Gas Turbine	339	372	600
Gas Turbine Inlet Temperature, °F	2,000	2,000	2,000
Steam Conditions, psig/°F/°F	1450/900/900	1450/900/900	1450/900/900
Surface Condenser Pressure, inches Hg abs.	2.5	2.5	2.5
Stack Gas Exit Temperature, °F	290	290	295
Gas Turbine Power#, MW	692.25	519.19	731.12
Steam Turbine Power#, MW	561.43	450.54	539.16
Fuel Gas Expander Power#, MW	36.64	26.37	--
Oxygen Plant Power##, MW	1.72	1.72	1.75
Power Consumed, MW	185.52	187.48	176.28
Net System Power, MW	1,106.52	810.34	1,095.75
METHANOL			
Methanol Produced, 10^3 gal/day	--	688.0	--
tons/day	--	2,283.4	--
FOEBδ/day	--	7,705	--
10^6 BTU/day	--	45,072	--
OVERALL SYSTEM			
Process and Deaerator Makeup Water, gpm	602	1,245††	--
gpm/1000 MW	544	1,536	1,270.4
Cooling Tower Makeup Water†, gpm	8,299	6,875	--
gpm/1000 MW	7,495	8,478	7,031
Cooling Tower Heat Rejection, % of coal HHV	42.9	35.5	38.71
Air Cooler Heat Rejection, % of coal HHV	0.30	1.28	0.397
Net System Heat Rate, BTU/kWh	9,214	--	9,404
Overall System Effieciency (Coal → Power and Methanol), % of coal HHV	37.0	45.5	36.29
Efficiency of Methanol Production, % of coal HHV	--	68.8%	--

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* Dry basis, 100 percent oxygen

** Excluding HHV of H_2S, COS, NH_3

\# At generator terminals

\#\# From power recovery expander in 11-ME-1

† Includes condensate from oxidant feed in Unit 11

†† Includes water for steam injection into gas turbine

δ FOEB = Barrels of distillate fuel oil (5.85 x 10^6 BTU/BBL) with high heating value equivalent to methanol produced

The results show that there is a relatively small impact on the cost of electricity for changes in the operating pressure of the gasification system, 600 psig to 1,000 psig; system efficiency increases about 2 percent while capital costs go up 7 percent, resulting in only a 4.2 percent increase in the cost of electricity.

Compared to the coal-fired steam plant, there is a 10 percent reduction in the cost of electricity primarily attributed to the higher overall system efficiency and lower O&M costs of the GCC systems.

The first year cost of "once-through" methanol has the potential to be 30 percent lower than the expected selling price of methanol produced by a nonregulated company. Forty percent of this differential methanol cost is due to the different methods of financing the two plants (i.e., regulated vs. nonregulated). The remaining 60 percent of the differential is due to more efficient conversion of the coal to methanol via the "once-through" route. The cost advantage increases after the first year where methanol cost decreases from $\$5.58/10^6$ BTU (first year) to a levelized cost of $\$4.32/10^6$ BTU compared to $\$7.87/10^6$ BTU for the dedicated plant; a savings of 45 percent.

A series of sensitivity analyses conducted during the course of this study have shown that a decrease in the assumed 10 percent general inflation rate increases the first year differential in cost of methanol from the two facilities. Likewise, a two year startup delay has a similar effect in first year methanol cost.

The efficiency of methanol production for the "once-through" concept (68.8 percent) is much higher than the dedicated coal-to-methanol concept (57.86 percent). This is due to the elimination of the detrimental impact of shift conversion, excess CO_2 removal and recycle of unconverted synthesis gas thus confirming the original premise under which the study was performed.

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

TABLE S-3
SUMMARY OF ECONOMIC RESULTS ****
METHANOL COPRODUCTION AND TEXACO-BASED GCC PLANTS

	Conventional Texaco Based GCC Power Plant With No Methanol Production	High Efficiency Oxygen Blown Texaco GCC (EPRI AP-1624)	Coal Fired Steam Plant FGD-Subscritical	Texaco GCC Plant With Once-Through Methanol Coproduction	Dedicated Coal to Methanol Plant (EPRI AP-1962)
CASE DESIGNATION	A2	EXTC-79	2400/1000/1000	B2	
CAPACITY FACTOR, %	70	70	70	90	90
PRODUCTION AT DESIGN CAPACITY					
Net System Power, MW	1,106.52	1,095.75	987.18	810.34	0
Overall Plant Heat Rate, BTU/kWh	9,214	9,404	9,981	N.A.	N.A
Methanol Produced, Tons/day	0	0	0	2,283.4	10,927
Overall System Efficiency)Coal ⟶ Power and Methanol), % of coal HHV	37.0	36.29	34.19	45.5	57.86
TOTAL CAPITAL REQUIRED*, $/kW	1,077	1,009	1,030	N.A.	N.A.
, $/FOEB**/day	N.A.	N.A.	N.A.	32,165	41,666
COST OF ELECTRICITY ***					
First Year, Mills/kWh	47.36	45.45	52.64	39.85	N.A.
Last Year, Mills/kWh	23.30	23.24	26.41	22.08	N.A.
Levelized, Mills/kWh	33.14	32.23	36.15	29.11	N.A.
COST OF METHANOL ***					
First Year, $/10⁶ BTU	⟵	N.A.	⟶	5.58	7.87
Last Year, $/10⁶ BTU	⟵	N.A.	⟶	3.55	7.87
Levelized, $/10⁶ BTU	⟵	N.A.	⟶	4.32	7.87

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* Mid-1980 basis

** FOEB = Barrels of distillate fuel oil (5.85×10^6 BTU/BBL) with high heating value.equivalent to methanol produced

*** In constant mid-1980 dollars

**** The dedicated coal to methanol plant economics have been based on nonregulated company ownership. All other cases employ regulated utility financing.

Section 1

INTRODUCTION

PURPOSE OF THE STUDY

Coal gasification combined-cycle (GCC) power plants based on second generation gasifiers continue to show promise for future utilization by the U.S. electric power industry. Of the three second-generation coal gasification processes presently supported by EPRI, the oxygen-blown Texaco entrained gasifier is currently being funded by EPRI for scale-up to a commercial baseload GCC power plant. A series of engineering and economic evaluations have been conducted to assess different process configurations for Texaco-based GCC power plants.

The Texaco coal gasification process technology is derived from Texaco Development Corporation's established commercial process for partial oxidation of heavy petroleum fractions. EPRI is committed to funding extensive pilot plant scale studies of the Texaco coal gasification process. Development of this process has reached the stage where a detailed engineering design of a demonstration-scale GCC facility for the electric power industry is now in preparation.

In July 1979, Texaco Inc. and Southern California Edison (SCE) jointly initiated the Cool Water Project. EPRI became a formal financial participant in the Cool Water Project in February 1980. The goal of this project is the design, construction, and operation of a 1000 MW demonstration-scale, coal gasification, combined-cycle power plant. This demonstration plant will employ both the oxygen-blown Texaco process and a 2000°F combustion turbine and will be located at SCE's Cool Water Station near Barstow, California. Field construction of the plant is scheduled to begin in 1981. Initial plant operations are targeted for 1984. Strong incentives exist, therefore, for EPRI to continue with engineering and economic studies of large (1000 MW) Texaco-based GCC systems which may ultimately be constructed as baseload power plants starting in the mid- to late 1980s.

An oxygen-blown coal gasification system, such as the Texaco process, is capable of providing a gaseous fuel for use in an integrated GCC baseload power plant; and can also produce raw synthesis gas feedstock for production of methanol. From an environmental viewpoint, methanol is an attractive fuel option to be used in both peaking and intermediate load generating units. Methanol is an easily transportable and storable liquid fuel containing no sulfur, nitrogen, or particulate matter. In the past, high prices for methanol as well as an extremely limited supply have been a hindrance to the use of this desirable fuel by the U.S. electric power industry. Therefore, a need exists to continue to examine process options which could lower the cost of methanol.

Some of the major factors which have contributed to the high cost of producing methanol in a dedicated coal gasification-based plant are as follows:

- The gasifier produces a raw synthesis gas (H_2 + CO) which must undergo shift conversion to ultimately provide a methanol synthesis feed gas with the necessary 2:1 stoichiometric ratio of H_2:CO. The use of the shift conversion process increases the CO_2 content of the raw synthesis gas. The excess CO_2 must be removed before the methanol synthesis reaction to avoid the need for processing and separating a significant quantity of "inerts." Removal of the excess CO_2 decreases the overall thermal efficiency of the process.

- Thermodynamic equilibrium limits the methanol synthesis reaction, such that only partial conversion can be achieved in a single pass through the reactor. Unconverted gas must be separated from the product methanol, recompressed, and recycled as feed to the reactor. The need for recycle operation increases capital costs due to the need for larger volume reactors and separation equipment; and increases energy requirements for separation and recompression operations, thus reducing overall thermal efficiency.

This study used a process for coproducing methanol in a GCC power plant which eliminated the detrimental impact of the above noted factors on the cost of methanol. This process eliminated the need to employ shift conversion, CO_2 removal, and recycle of unreacted synthesis gas.

EPRI specified the use of Chem Systems methanol process for this study. This process uses an ebullated-bed reactor, which allows for the operation of an isothermal reactor with the heat of reaction being essentially removed by an inert oil. Development of this process has progressed under EPRI Research Project 317. Results of this development program show that methanol can be produced from coal-derived synthesis gas via the Chem Systems process, with further development necessary to produce an abrasion resistant, active, low-cost catalyst.

The major goal of this study was to determine if the cost of producing methanol could be reduced by producing it as a by-product in an integrated gasification combined-cycle plant. The designs used were for grass roots facilities, based on the oxygen-blown Texaco gasification process and the Chem Systems liquid phase, once-through reactor-type methanol synthesis process; integrated with currently available gas turbine combined-cycle power plants. Designs for the Texaco gasification unit, the Chem Systems methanol synthesis unit, and the Selexol acid gas removal unit were based on information provided by the appropriate licensors.

DESCRIPTION OF THE CASE STUDIES

An initial screening evaluation of four cases was conducted. The first of these cases was a Texaco-based GCC plant with no methanol coproduction to provide a basis for estimating the cost of electric power to be applied as a credit in the methanol coproduction cases. The other three cases all coproduced methanol together with electricity. All of these initial designs employed advanced 2,400°F gas turbines in the power block and three of the designs employed high temperature gas coolers raising superheated steam in the gasification section of the plant.

At about the same time that these screening studies were completed, results from other evaluation efforts had demonstrated that Texaco-based GCC plants employing currently available (2,000°F) gas turbines had the potential to be competitive with coal-fired steam plants with stack gas scrubbers. It had also been shown that raising high pressure superheated steam instead of saturated steam in the Texaco gas coolers contributed very little to GCC system efficiency while adding substantially to the plant capital investment. It was therefore judged that these initial screening designs did not provide a useful framework for evaluation of the coproduction of methanol and electricity using the "once-through" synthesis concept.

For this reason, two additional cases were evaluated which form the basis for this report. Both of these cases employ currently available, 2,000°F gas turbines in the power block and produce saturated high pressure steam in the high temperature gas cooling section of the plant.

Case A2 represents a design for a conventional Texaco-based GCC plant producing no methanol to provide a basis for estimating the cost of electricity. For this case, the gasification section of the plant was designed to operate at an average pressure of 1000 psig. High-temperature gasifier effluent is cooled in heat exchange equipment by generation of 1500 psig saturated steam. This gas contains particulate matter, which is then removed in the particulate scrubbing unit. The gas is further cooled in gas cooling unit (Unit 21) and then enters the acid gas removal system, where 90 percent of the sulfur compounds are removed to meet environmental standards. The treated gas is reheated, expanded to recover energy in a fuel gas expander, and then combusted in a gas turbine firing at 2,000°F. Turbine exhaust gas is then sent to the heat recovery steam generators (HRSGs).

The second design, coproducing methanol and electricity, is represented as Case B2. For this design, the gasification and power production sections of the plant were kept as similar as possible to equivalent sections in the electricity only plant (Case A2). Modifications were made to accommodate the inclusion of a "once-through" methanol synthesis reactor between the acid gas removal unit and the gas turbines. Therefore, in the Case B2 design, 1500 psig saturated steam is generated by cooling the hot gasifier effluent. The gas stream is scrubbed for the removal of entrained particulate matter, and is then reheated and processed through a COS hydrolysis unit, which shifts COS to H_2S, in order to reduce energy requirements and equipment sizes in the acid gas removal unit. Effluent from the COS hydrolysis unit is cooled in the gas cooling section, and is processed for the removal of sulfur compounds to a level of 5 ppmv in the acid gas removal unit. The treated gas is reheated before being sent to the Chem Systems once-through methanol synthesis reactor operating at a relatively low conversion. Methanol product is recovered and sent to storage, while the unconverted gases are reheated and expanded in the gas expander. The unconverted gas is then combusted in 2,000°F gas turbines. Exhaust gases from the turbines flow to the HRSGs.

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, and general plant requirements.

Gasifier heat and material balances were those used for a previous EPRI report (1) and are typical of a Texaco single-stage, entrained-bed oxygen-blown gasifier operating at 600 psig. It is important to realize that these gasifier system designs are based on a mature and well developed gasification technology. Gasifier capacities and coal slurry concentrations are higher than those currently being proposed for the first generation of Texaco-based gasification systems. As noted in a previous EPRI report (2), pressure has little effect on gasifier yields. Therefore, the yields at 600 psig were deemed to be acceptable for use with the 1000 psig gasifier used in this study.

Gas turbine performance was estimated by EPRI for commercially available equipment operating at a 2000°F combustor outlet temperature.

Methanol plant heat and material balances were supplied by Chem Systems, Inc. for a liquid phase, once-through-type reactor methanol synthesis process.

The coal used is Illinois No. 6 and is defined by the analysis given in Table 1-1. The coal was assumed to be delivered to the site washed and sized. If experience were to demonstrate that this assumption was not reliable, then each of the cases presented here would require additional coal handling equipment. This would slightly affect overall plant costs but would not alter the comparison between cases.

The site for the plant is the Chicago area, and Table 1-2 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, Table 1-3,

Table 1-1

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	-
	100.0
HEATING VALUE - AS RECEIVED	
Higher Heating Value (HHV)	12,235 Btu/lb
Net Heating Value (LHV)	11,709 Btu/lb
AS PURCHASED	
Washed, sized 1-1/2" x 0, delivered to plant battery limits by unit train	

Table 1-2

SITE CONDITIONS

LOCATION	Chicago, Illinois
ELEVATION	600 feet
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 1-3

WATER ANALYSIS

(ppmw)

Silica (SiO_2)	1.8
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

of finished water from the South District filtration plant. These data were extracted from a previous EPRI report (3).

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

- Cooling tower
- Plant and instrument air
- Potable and utility water
- Fuel gas and nitrogen systems
- Fire water
- 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

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 on-stream 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.

Multiple processing trains have been necessitated by the large size of the plant. The number of trains is generally established by the limiting size of major process equipment that will be available in the near future; and by shipping size limitations for pressure vessels ("shop fabricated").

REFERENCES

1. "Economic Studies of Coal Gasification Combined Cycle Systems for Electric Power Generation," EPRI AF-642, January 1978.
2. "Effects of Sulfur Emission Controls on the Cost of Gasification Combined Cycle Power Systems," EPRI AF-916, October 1978.
3. "Economics of Current and Advanced Gasification Processes for Fuel Gas Production," EPRI AF-244, July 1976.

Section 2

DISCUSSION OF TECHNICAL RESULTS

GENERAL DISCUSSION

Table 2-1 presents a summary of operating results for the two designs prepared for this study. The first two columns represent results for each plant as designed. These designs were based on the same coal feed rate to each of the plants in order to minimize the cost of performing the study. As plant B2 (the methanol coproduction Case) converts some of the fuel gas to methanol, it produces 296.18 MW less power than Case A2. In order to make comparisons between Cases A2 and B2, it has been deemed expedient to scale up the B2 plant such that it produces the same amount of electric power as the dedicated GCC power plant (Case A2). This scale-up has been conducted in a linear manner and has been done simply for comparison purposes. All engineering design and cost estimating work has been conducted on the unscaled B2 plant feeding the same amount of coal as Case A2.

The most interesting result of Table 2-1 is the 68.8 percent conversion efficiency of coal to methanol in this "once-through" mode of operation. This efficiency is calculated by comparing the scaled-up coproduction plant results with those for the conventional GCC system (Case A2). These two plants have the same 1,106.52 MW capacity. The coproduction plant, however, produces an additional 10,520 fuel oil equivalent barrels per day of methanol and consumes an additional 3,502 tons per day of moisture free coal. The 68.8 percent efficiency represents the efficiency with which the higher heat content of the additional 3,502 tons per day (MF) coal has been converted to 10,520 FOE barrels per day of methanol. In a dedicated coal to methanol plant, the net conversion efficiency could be expected to be in the range 53 to 58 percent. Therefore, the above result indicates that the "once-through" synthesis concept evaluated in this study provides certain real process advantages that translate directly into increased production efficiency. In general, the major efficiency improvements over dedicated methanol production resulting from "once-through" synthesis in a GCC plant are due to the following process differences:

Table 2-1

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

CASE DESCRIPTION	Conventional Texaco Based GCC Power Plant With No Methanol Production	Texaco GCC Plant With Once-Through Methanol Coproduction (As Designed)	Methanol Coproduction Case (Same Design As Case B2), Scaled To Produce Same Quantity Of Electricity As Case A2
CASE DESIGNATION	A2	B2	B2-Scaled-Up
GASIFICATION SYSTEM			
Coal Feed Rate, lbs/hr m.f.	798,333	798,333	1,090,124
Oxygen/Coal Feed*, lb/lb m.f.	0.858	0.858	0.858
Oxidant Temperature, °F	287	287	287
Slurry Water/Coal Ratio, lb/lb m.f.	0.503	0.503	0.503
Gasification Section Average Pressure, psig	1,000	1,000	1,000
Crude Gas Temperature, °F	2,300-2,600	2,300-2,600	2,300-2,600
Crude Gas HHV (Dry Basis), BTU/SCF**	281.1	281.1	281.1
POWER SYSTEM			
Temperature of Fuel Gas to Gas Turbine, °F	339	372	372
Gas Turbine Inlet Temperature, °F	2,000	2,000	2,000
Steam Conditions, psig/°F/°F	1450/900/900	1450/900/900	1450/900/900
Surface Condenser Pressure, inches Hg abs.	2.5	2.5	2.5
Stack Gas Exit Temperature, °F	290	290	250
Gas Turbine Power#, MW	692.25	519.19	708.95
Steam Turbine Power#, MW	561.43	450.54	615.21
Fuel Gas Expander Power#, MW	36.64	26.37	36.01
Oxygen Plant Power##, MW	1.72	1.72	2.35
Power Consumed, MW	185.52	187.48	256.00
Net System Power, MW	1,106.52	810.34	1,106.52
METHANOL			
Methanol Produced, 10^3 gal/day	--	688.0	939.5
tons/day	--	2,283.4	3,118.0
FOEBδ/day	--	7,705	10,520
10^6 BTU/day	--	45,072	61,546
OVERALL SYSTEM			
Process and Deaerator Makeup Water, gpm	602	1,245††	1,700††
gpm/1000 MW	544	1,536	2,097
Cooling Tower Makeup Water†, gpm	8,299	6,875	9,388
gpm/1000 MW	7,495	8,478	11,577
Cooling Tower Heat Rejection, % of coal HHV	42.9	35.5	35.5
Air Cooler Heat Rejection, % of coal HHV	0.30	1.20	1.28
Net System Heat Rate, BTU/kWh	9,214	--	--
Overall System Efficiency (Coal → Power and Methanol), % of coal HHV	37.0	45.5	45.5
Efficiency of Methanol Production, % of coal HHV	--	68.8%	68.8%

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*Dry basis, 100 percent oxygen

#At generator terminals

†Includes condensate from oxidant feed in Unit 11

**Excluding HHV of H_2S, COS, NH_3

##From power recovery expander in 11-ME-1

††Includes water for steam injection into gas turbine

δFOEB = Barrels of distillate fuel oil (5.85 x 10^6 BTU/BBL) with higher heating value equivalent to methanol produced

- In the "once-through" synthesis mode, unconverted synthesis gas is not recycled to the methanol reactors, thereby eliminating the power required to operate the recycle gas compressors.

- In the concept evaluated for this study, the fuel gas produced by the gasifiers does not need to be shifted to produce a 2:1 hydrogen:carbon monoxide ratio, nor does it have the bulk of the CO_2 removed prior to entering the methanol synthesis reactors, thereby eliminating the energy requirements for gas shifting as well as for CO_2 removal.

- By integrating the "once-through" synthesis system into a GCC power plant, maximum use can be made of the heat liberated during methanol synthesis. In a dedicated methanol plant, this liberated heat (at 450°F to 500°F) is at too low a temperature to be efficiently utilized. If the methanol synthesis unit is integrated with a GCC power plant, however, this heat can be used to raise intermediate and low pressure steam which can be efficiently utilized in the power plant's steam system.

The conventional GCC plant design (Case A2) employed for this study parallels, in large part, other Texaco-based GCC plant designs developed for EPRI. Therefore, discussion of the Case A2 design will be limited, as it can be found in other EPRI reports (i.e., EPRI report number AP-1624, November, 1980). The only significant difference between the Case A2 design and those appearing in earlier EPRI reports is the operating pressure of the plant. The Case A2 system has a gasification section average pressure of 1,000 psig whereas other EPRI studies have been based on an average gasification section pressure of 600 psig. Increasing the gasifier operating pressure appears to have minor impact on GCC system performance or cost (system efficiency increases by approximately 2½ percent, but capital costs in $/kW also increase by approximately 7 percent). Therefore, the net impact of GCC system pressure on the cost of electricity is negligibly small.

In considering the "once-through" coproduction of methanol (Case B2), some of the more important design considerations follow:

a. Methanol Reaction System: For this case, the Chem Systems Liquid Phase Methanol Process, currently under development, has been evaluated. It is important to point out the fact that this methanol synthesis system (under development since 1975; see EPRI report number AF-1291, December, 1979) has a long way to go before it can be considered to be commercially available. Initial evaluations were conducted in a one inch diameter tube capable of handling approximately 40 SCFH of feed gas. This bench scale unit was scaled up to a 3.62 inch diameter reactor process development unit capable of synthesis gas feed rates of 1500 SCFH (equivalent to between 0.25 ton per day and 0.5 ton per day methanol). Plans to demonstrate a scaled-up version (a two foot diameter reactor feeding 62,500 SCFH of gas) of this process development unit are currently underway by Chem Systems and Air Products. For the designs presented in this study, "once-through" liquid phase methanol synthesis reactors having a capacity of approximately 5×10^6 SCFH of synthesis gas have been assumed.

The temperature of the methanol reactor effluent is limited by the catalyst and because lower temperature favors methanol reaction equilibrium. This temperature limitation results in the production of steam at no higher than 295 psig pressure. Since the IP steam level used in the steam turbine is at 445 psig, 295 psig steam generated in the methanol plant is best suited only as injection steam in the gas turbine combustor. Injection not only controls the NO_x emissions in the gas turbine exhaust, but also slightly reduces the fuel gas flow per unit of power produced, thereby marginally improving the gas turbine efficiency.

The methanol reaction system design employed for Case B2 minimizes the conversion of feed gas to methanol as the recycle oil contains a substantial quantity of dissolved methanol. The disadvantage that this implies is that all of the gas produced in the plant needs to be desulfurized to the 5 ppmV level to protect the synthesis catalyst. If after exiting the reactor, the recycle oil is flashed to recover dissolved methanol, the methanol produced per unit of feed gas could be substantially increased. For a plant employing this alternate "high conversion" design to produce the same quantity of methanol as the Case B2 design, only 59 percent of the gasifier product would have to be fed to the methanol synthesis reactors. Therefore, the remaining 41 percent of the product gas could be split from the main gas stream and separately desulfurized only to the extent that the total power plant flue gas would comply with environmental control requirements. This procedure would reduce the overall cost of sulfur removal somewhat and would also decrease the cost of the methanol reaction section. The initial screening studies indicated that this high conversion approach to "once-through" synthesis could reduce the cost of methanol production by approximately 7 percent.

b. Desulfurization of the Syngas: The feed gas to the methanol synthesis reactors should contain no more than 5 ppmV sulfur compounds. The raw gas from the gasifiers contains approximately 13,000 ppm H_2S and 740 ppm COS. In order to get down to the 5 ppm level, all of the H_2S and 99.3 percent of the COS must be removed. The Selexol plant can achieve these removal requirements only if very high solvent circulation rates are employed. For this reason, the initial screening studies indicated that it would be economically attractive to install a COS hydrolysis unit upstream of the Selexol plant to convert the COS to H_2S. This has the effect of substantially decreasing the solvent circulation rate in the Selexol absorber, thereby reducing the cost of the overall acid gas removal system by 20 percent and reducing its power consumption by 40 percent. The major negative impact associated with the COS hydrolysis unit is that gas exiting the particulate scrubber at 393°F must be heated to 440°F by heat exchange with raw fuel gas prior to entering the hydrolysis beds. This decreases the system's capability to generate some low pressure steam, thereby reducing power generation capability somewhat.

Final gas polishing to remove small quantities of sulfur compounds is accomplished in ZnO beds. The syngas is heated to 646°F against gasifier effluent before being processed in the ZnO beds. ZnO beds can operate at ambient temperature if only H_2S and COS removal is required, at the expense of much more ZnO consumption. But, as a precautionary measure, ZnO is operated at about 650°F so that any other trace sulfur compounds like mercaptans, if present, can be desulfurized. If tests prove the nonexistence of other sulfur compounds, then the ZnO can operate at ambient temperature. Operating at lower temperature will be more efficient because feed gas to the methanol plant need only be heated to about 450°F, thereby liberating sensible heat in the raw gas for raising medium pressure steam and increasing the total power generation capability of the system.

POTENTIAL IMPROVEMENTS

Consideration should be given to optimizing the cooling water system. The cooling tower approach, surface condenser area and LP power turbine back pressure are all related parameters which should be optimized to produce the lowest cost of electricity. In addition to the above, the cooling tower best suited for this plant should be investigated. So far, these designs consider only mechanical draft towers, but use of natural draft (hyperbolic) may become common in the face of higher energy prices.

Savings may be made in equipment costs, if the number of air separation plants are reduced by increasing the unit train capacity. For example, the particular designs for both Cases A2 and B2 in this study employ five operating air separation and oxygen compression trains, each producing approximately 1,700 tons per day of oxygen at an installed cost of $18,300 per tons per day of oxygen. Currently there are in operation a number of air separation plants having capacities of 2,200 tons per day of oxygen (see EPRI report number AP-1674, January, 1981). This report, prepared by the Linde Division of Union Carbide Corporation, indicates that increasing train capacity from 1,700 tons per day to 2,200 tons per day would result in a plant investment reduction per ton of oxygen of approximately 10 percent. This would reduce the total capital requirement for the case A2 GCC power plant by $20/kW leading to an overall reduction in the levelized cost of electricity of approximately 1 percent. In addition, if a lower oxygen concentration was used, the cold box inlet pressure could be reduced. Then the air compressor ratio and power could also be reduced. The oxygen compressor power would increase because of increased flow of nitrogen and argon, but this percentage flow increase required less additional power than saved on the air compressors. All of the air and oxygen compressors are motor driven which will make turn down more power consuming. Reduction in the purity of oxygen leaving the cold box could also be considered as a means of reducing capital investment.

The balances included in this study do not carefully follow trace components like ammonia. A more detailed mass and energy balance around the gasifiers, including ammonia content in the recycle water streams and more definition in process condensate blowdown would be helpful.

The ZnO beds can operate at ambient temperature instead of at 650°F if tests prove no other sulfur compounds besides H_2S and COS are present even in trace amounts. Operation at 650°F is done as a precautionary measure. Of course, the consumption of ZnO will increase for ambient temperature operation. Further analysis should be done for an optimum ZnO-bed operating temperature.

Temperature limits in the methanol plant dictate production of steam at no higher than 295 psig. This steam is used as injection steam in the gas turbines. More analysis needs to be done to compare using this steam as injection steam against using it in the steam turbine.

The cost of the gasifier HP steam generator appears to be very sensitive to the cold end approach. Significant savings can be realized by optimizing the temperature approaches for this exchanger.

Section 3

PLANT DESCRIPTIONS - CASE A2
BASE CASE FOR COST OF ELECTRICITY WITH 1500 PSIG/900°F PROCESS STEAM AND 2000°F GAS TURBINES

GENERAL

A grass roots plant for electric power generation based on single-stage entrained oxygen-blown gasifiers of the Texaco type, integrated with current state-of-the-art combined-cycle generating equipment, is shown schematically on Block Flow Diagram EXTC(ME-A2)-1-1 for Case A2. Each block indicates the area and unit numbering, as well as the number of operating trains in each unit. The plant consumes 10,000 short tons per day of Illinois No. 6 coal, fed to the gasifiers in a water slurry containing 66.5 weight percent solids.

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

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

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

Table 3-1 shows the number of operating and spare trains for major sections of Case A2.

Table 3-1

TRAINS OF EQUIPMENT IN MAJOR PLANT SECTIONS - CASE A2

Unit No.	Unit Name	Operating	Spare
10	Coal Handling	1	0
10	Coal Grinding	2	0
10	Slurry Preparation	1	0
11	Oxidant Feed	5	0
20	Gasification	3	1
20	High-Temperature Gas Cooling and Gas Scrubbing	3	1
20	Ash Handling	1	0
21	Gas Cooling	2	0
22	Acid Gas Removal	2	0
23	Sulfur Recovery	2	1
24	Tail Gas Treating	2	1
30	Steam, BFW and Condensate System		
	• Condensate Collection and and Deaeration	1	0
	• Water Treating	1	0
32	Cooling Water System	1*	0
40	Effluent Water Treating	1	0
40	Process Condensate Treating	1	0
50	Gas Turbine/Generator	7	0
51	Heat Recovery Steam Generator	14	0
51	Steam Turbine/Generator	1	0

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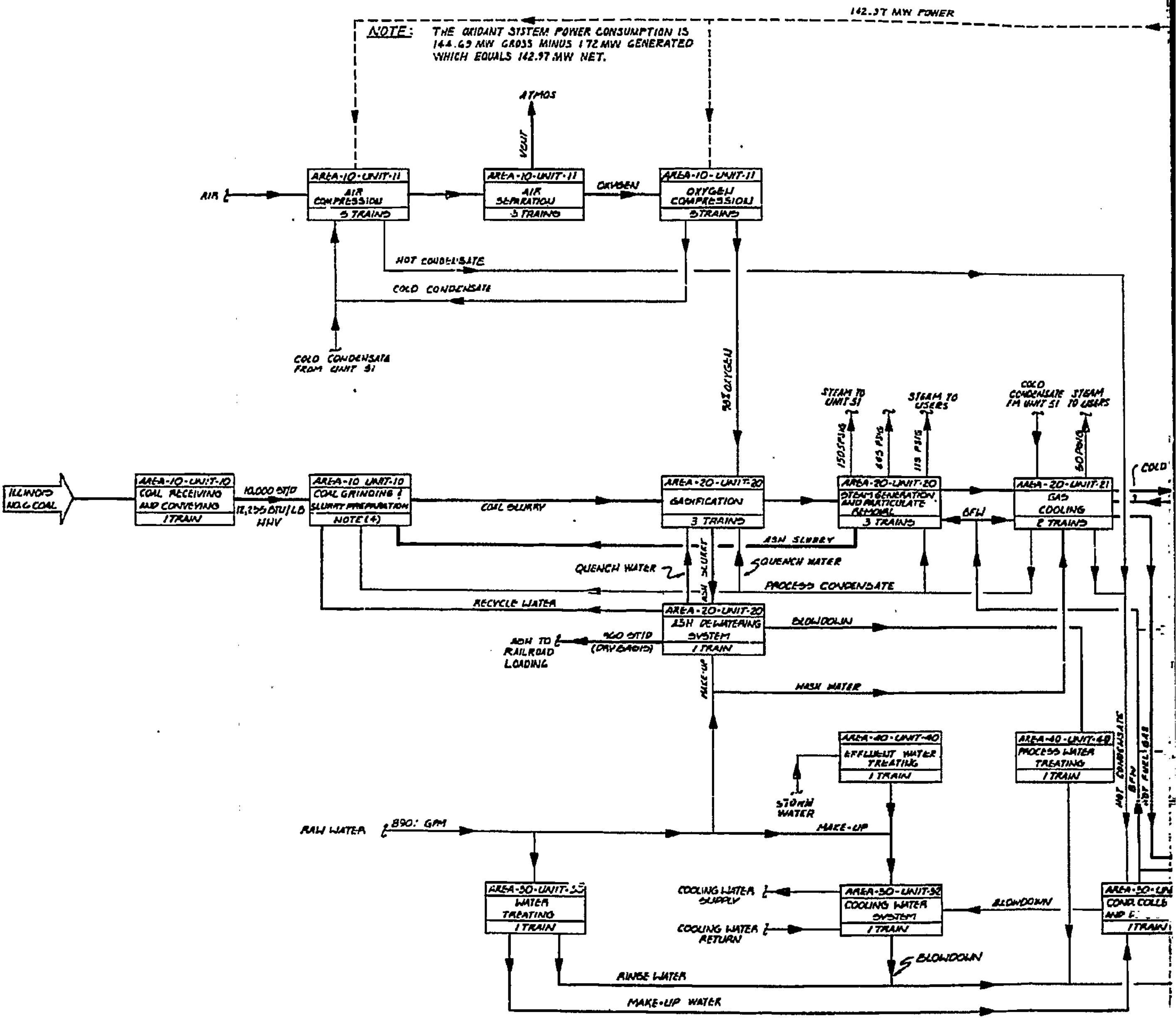
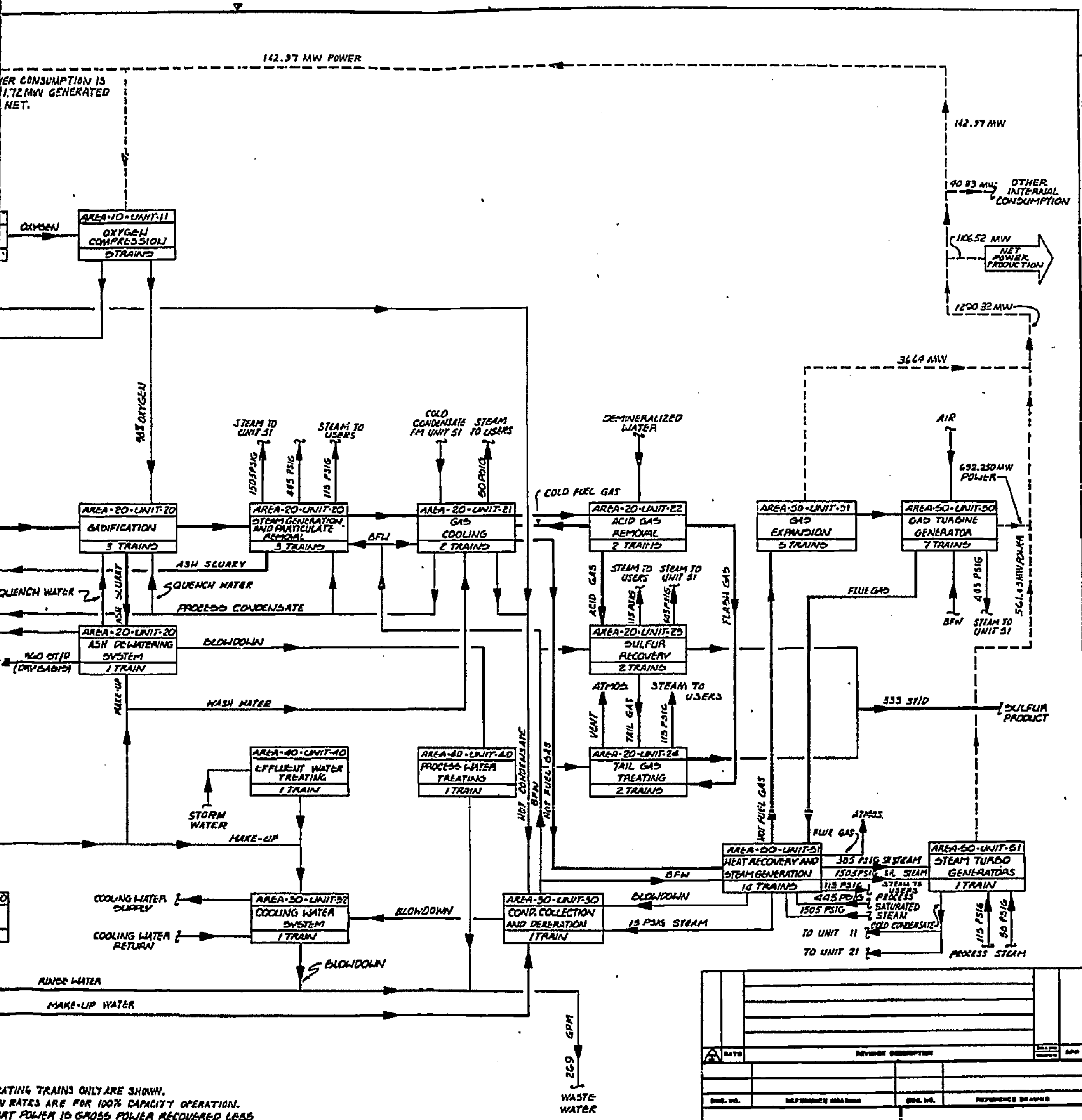
*The cooling tower dedicated to the process plant sections is separate from the towers dedicated to the steam turbogenerator condenser



NOTE: THE OXIDANT SYSTEM POWER CONSUMPTION IS 144.69 MW GROSS MINUS 1.72 MW GENERATED WHICH EQUALS 142.97 MW NET.

1 OPERATING TRAINS ONLY ARE SHOWN.
2 FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3) EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
4) TWO TRAINS OF COAL GRINDING & ONE TRAIN OF SLURRY PREPARATION ARE PROVIDED.



…ATING TRAINS ONLY ARE SHOWN.
…N RATES ARE FOR 100% CAPACITY OPERATION.
…RT POWER IS GROSS POWER RECOVERED LESS
…IT POWER CONSUMPTION.
…TRAINS OF COAL GRINDING & ONE TRAIN OF SLURRY
…ARATION ARE PROVIDED.

DATE	REVISION DESCRIPTION		

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

T. LEE

D'DELMITO

RAVIKUMAR

SHELTON

EPRI

OVERALL BLOCK FLOW DIAGRAM
CASE A2-BASE CASE
(NO METHANOL 1500 PSIG STM)

EPRI
PALO ALTO, CALIF.

NONE
448334- EXTC (ME-A2)-1-1
04

COAL HANDLING, PULVERIZATION, AND SLURRY PREPARATION

Process Flow Diagram EXTC-(ME)-10-1 depicts the arrangement of equipment which incorporates one train of coal unloading, stacking, reclamation, and conveying, followed by two trains of grinding and one of slurry preparation. This section is common for both cases.

Washed, 1-1/2 inch by zero Illinois No. 6 coal is received at the plant site by unit train. The coal is unloaded from 100-ton bottom dump cars into unloading hopper 10-BN-1, at 1250 tons per hour. The unloading and stacking system is designed to handle a one day supply in eight hours. Four vibrating feeders 10-FE-1A-D withdraw coal from the hopper and place it on receiving conveyor 10-CV-1, while belt scale 10-SC-1 measures the actual conveyor transport rate. After passing a magnetic separator 10-MS-1 for protection of downstream equipment from miscellaneous metal fragments, the coal travels on sample tower conveyor 10-CV-2, which houses the sampling system 10-SA-1. From 10-CV-2, storage conveyor 10-CV-3 transports the coal to a tripper which supplies the double boom stacker 10-ME-1. The stacker travels on tracks and forms up to 3-1/2 day (35,000 tons) live storage piles on either side. Total live storage is limited to seven days to reduce the possibility of spontaneous ignition.

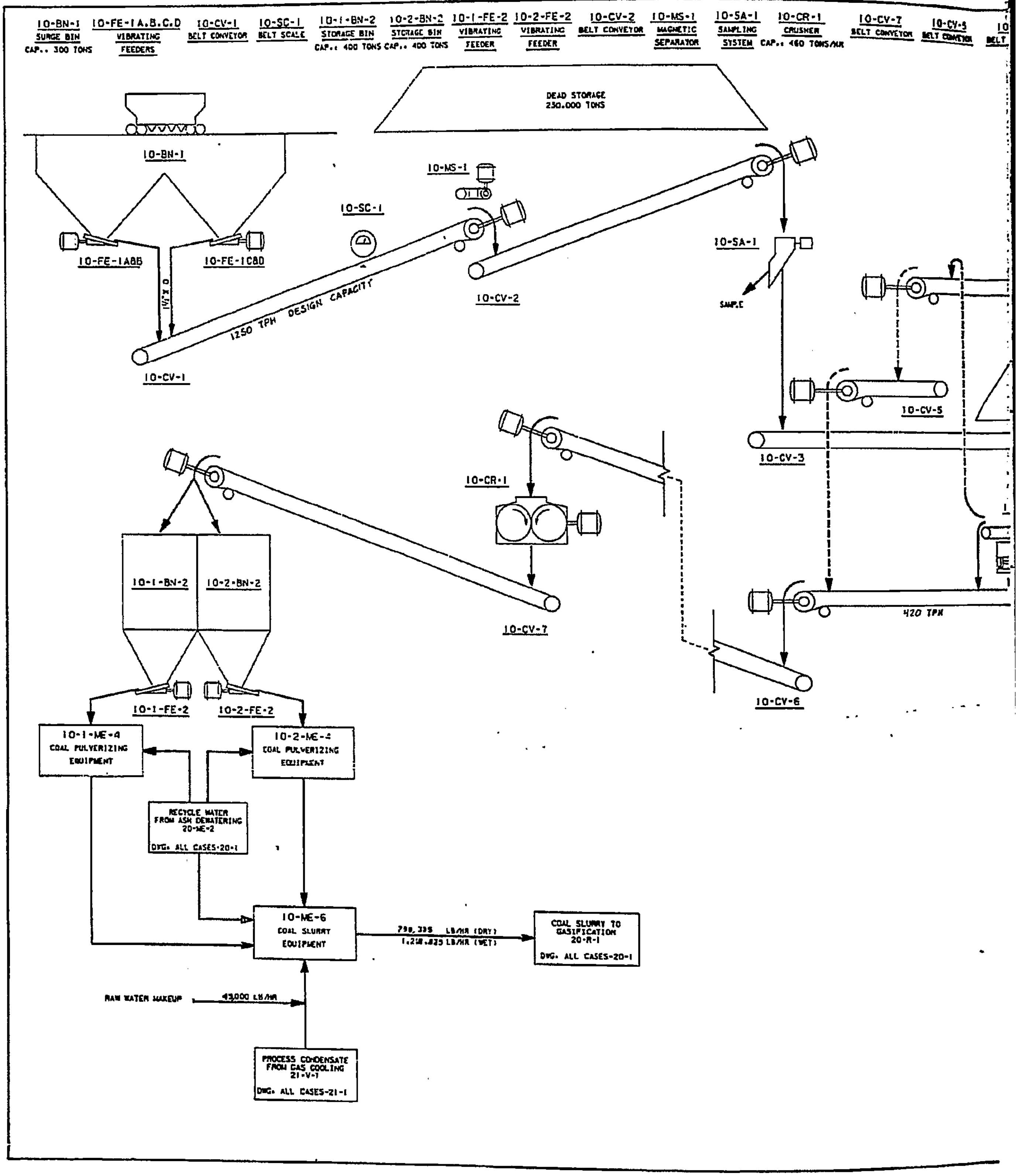
Space for a reserve dead pile of up to 60 days storage is provided adjacent to the rail unloading station. The amount of coal in the dead pile is assumed to be a 23-day supply. Total capital requirement is based on 30 days of coal inventory (7 days live and 23 days dead). The dead pile is sodded to minimize coal entrainment in rain water. Nevertheless, rain water runoff from this coal pile is collected and used in slurry preparation.

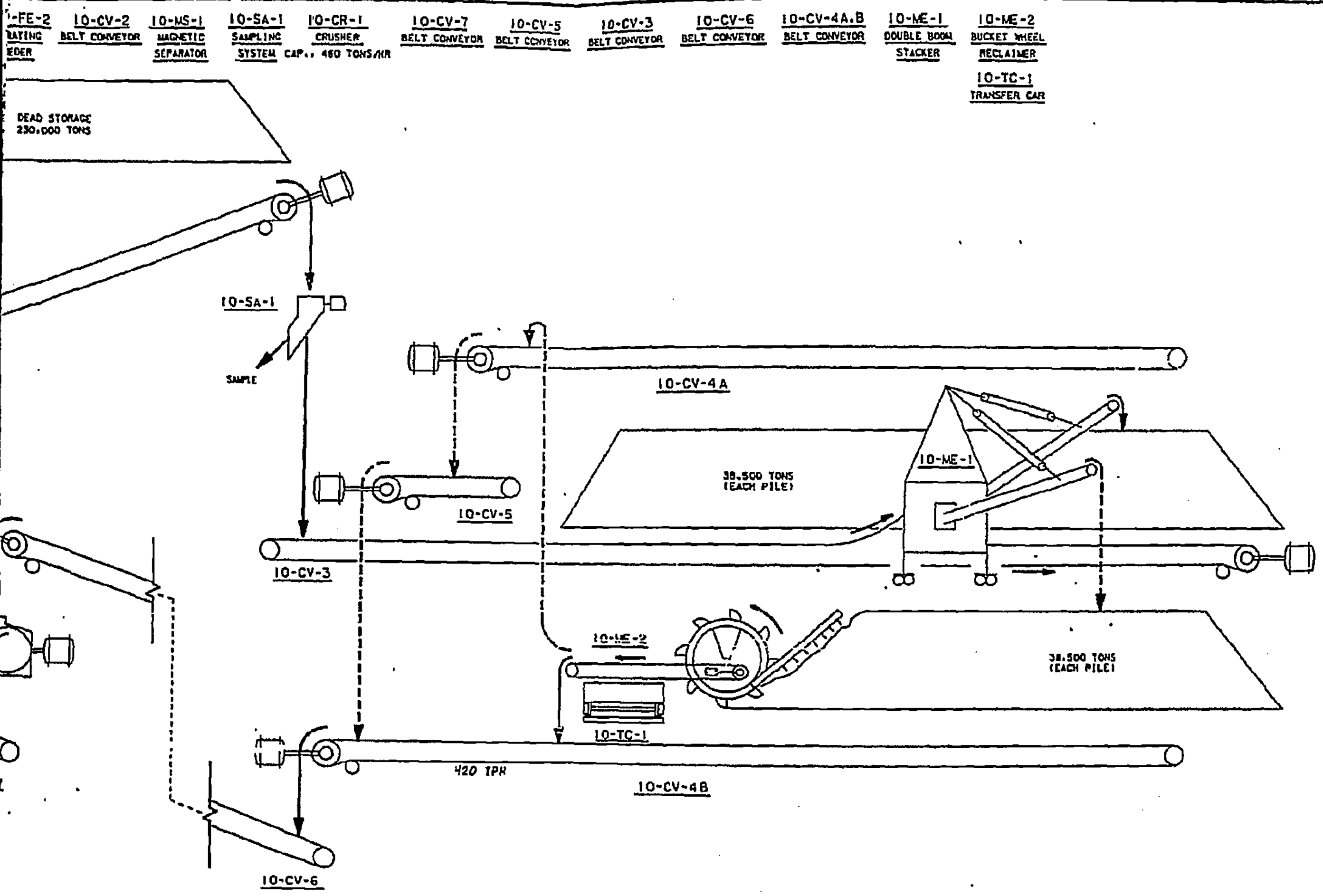
Coal is reclaimed from the storage piles by a bridge-type bucket wheel reclaimer 10-ME-2, rated at 420 tons per hour. This machine is moved between live storage piles as necessary by transfer car 10-TC-1. The wheel moves across the face of the pile, making an angle of repose cut across the many layers of coal, thereby blending the coal fed to the gasification plant. This blending provides more uniform gasifier operation. The reclaimer continuously moves ahead, reclaimed coal being carried on the bucket wheel conveyor to one of the two reclaim conveyors 10-CV-4A&B. Cross conveyor 10-CV-5 is employed when 10-CV-4A is in service, to deliver coal to crusher conveyor 10-CV-6, which is located near 10-CV-4B.

A Nittle crusher 10-CR-1 is used to break down 1-1/2 inch by zero inch coal, which would require longer residence times in the pulverizing equipment. Crushed coal conveyor 10-CV-7 delivers 3/4 inch by zero coal to storage bins 10-BN-2, which provide storage capacity equivalent to 1-3/4 hours of downstream throughput.

Vibrating feeders 10-FE-2, at the bottoms of the storage bins, supply two trains of the grinding mills, which pulverize the coal in two stages. Coal is then slurried with recycle process water and raw makeup water, and stored in a tank of about 24-hour capacity. The 65.5 percent solids slurry is then pumped by three parallel charge pumps to the three operating gasifiers.

The unloading system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. Local environmental regulations may seriously impact this area of design.


10-BN-1
SURGE BIN
CAP.. 300 TONS
10-FE-1A.B.C.D
VIBRATING FEEDERS
10-CV-1
BELT CONVEYOR
10-SC-1
BELT SCALE
10-1-BN-2
STORAGE BIN
CAP.. 400 TONS
10-2-BN-2
STORAGE BIN
CAP.. 400 TONS
10-1-FE-2
VIBRATING FEEDER
10-2-FE-2
VIBRATING FEEDER
10-CV-2
BELT CONVEYOR
10-MS-1
MAGNETIC SEPARATOR
10-SA-1
SAMPLING SYSTEM
10-CR-1
CRUSHER
CAP.. 460 TONS/HR
10-CV-7
BELT CONVEYOR
10-CV-5
BELT CONVEYOR
10
BELT
DEAD STORAGE
250.000 TONS
10-BN-1
10-FE-1A&B
10-FE-1C&D
1½" x 0
1250 TPH DESIGN CAPACITY
10-CV-1
10-SC-1
10-MS-1
10-CV-2
10-SA-1
SAMPLE
10-CV-5
10-CV-3
10-CR-1
10-CV-7
10-1-BN-2
10-2-BN-2
420 TPH
10-CV-6
10-1-FE-2
10-2-FE-2
10-1-ME-4
COAL PULVERIZING EQUIPMENT
10-2-ME-4
COAL PULVERIZING EQUIPMENT
RECYCLE WATER FROM ASH DEWATERING 20-ME-2
DWG. ALL CASES-20-1
10-ME-6
COAL SLURRY EQUIPMENT
798,335 LB/HR (DRY)
1,228,825 LB/HR (WET)
COAL SLURRY TO GASIFICATION 20-R-1
DWG. ALL CASES-20-1
RAW WATER MAKEUP
43,000 LB/HR
PROCESS CONDENSATE FROM GAS COOLING 21-V-1
DWG. ALL CASES-21-1



COAL SLURRY TO
GASIFICATION
20-R-1

G. ALL CASES-20-1

NOTE:

ALL FLOW RATES AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY UNLESS OTHERWISE NOTED.

G. CWIEKA
B.ASH
T.C.O'DELMITO
R.RAVIKUMAR
E.C.SHELSER
EPRI

PROCESS FLOW DIAGRAM
COAL PREPARATION
TEXACO PROCESS-OXYGEN BLOWN, SAT.STEAM CASE
EPRI PALO ALTO,CALIFORNIA

NONE
448334-EXTC-(ME)-10-1
02
483SK0201

OXIDANT FEED

Process Flow Diagram EXTC(ME-A2)-11-1 shows the oxidant feed system design used for Case A2. There are five parallel trains each consisting of one air compression system, one air separation plant and one oxygen compression system. No spare train is provided in this section.

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

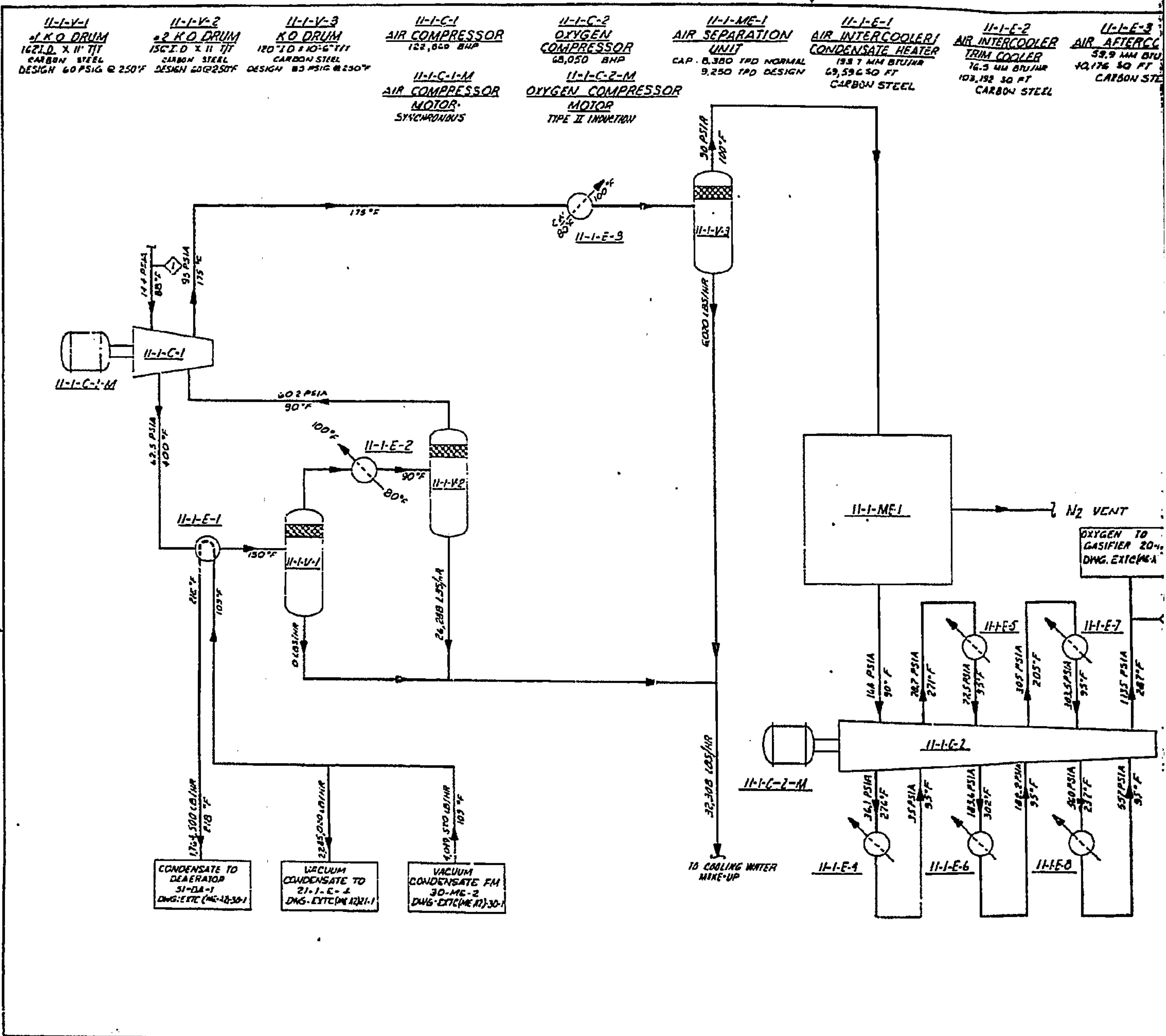
The 122,900 total hp required by the air compressors is supplied by electric motors. The compressed air at 90 psia, 100°F is processed in air separation unit 11-1-ME-1, to produce a total of 8380 tons per day (100 percent O_2 basis) of 98 volume percent oxygen. The air separation unit operating parameters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

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

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

Equipment Notes

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



MATERIAL BALANCE

COMPONENT	PROCESS STREAM 1 AIR MPH	MOL %	5 OXIDANT MPH	MOL %
O2	22,714	20.48	21,397.5	98.00
N2	84,320	76.07	330.0	1.50
Ar	1,087	0.98	108.5	0.50
H2O	2,739	2.47	—	—
TOTAL MPH	110,910	100.00	21,836.0	100.00
LB/HR	3,188,138		698,250	
MOL WT	28.70		31.98	

11-1-C-2
OXYGEN COMPRESSOR
68,050 BHP
11-1-C-2-M
OXYGEN COMPRESSOR MOTOR
TYPE II INDUCTION

11-1-ME-1
AIR SEPARATION UNIT
CAP. 8,380 TPD NORMAL
9,250 TPD DESIGN

11-1-E-1
AIR INTERCOOLER/ CONDENSATE HEATER
193.7 MM BTU/HR
69,596 SQ FT
CARBON STEEL

11-1-E-2
AIR INTERCOOLER TRIM COOLER
76.5 MM BTU/HR
103,197 SQ FT
CARBON STEEL

11-1-E-3
AIR AFTERCOOLER
59.9 MM BTU/HR
40,176 SQ FT
CARBON STEEL

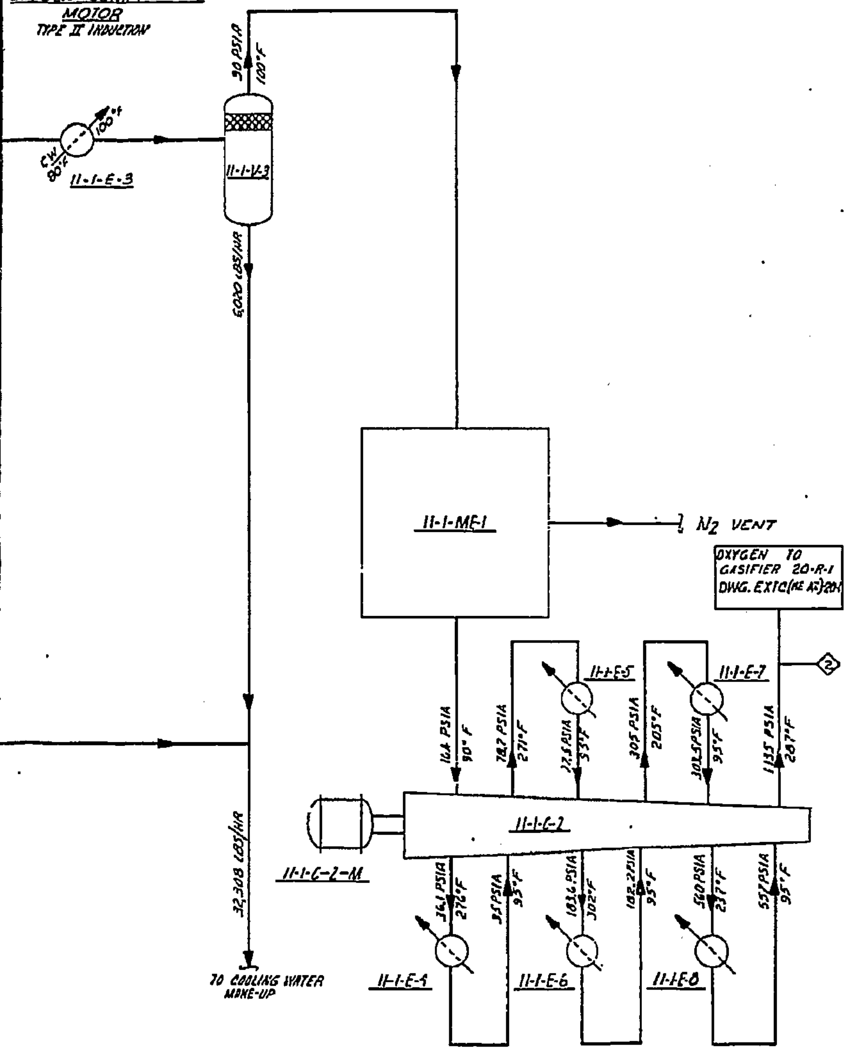
11-1-E-4
1ST STAGE O_2 COOLER
29.1 MM BTU/HR
25,177 SQ. FT.
CARBON STEEL

11-1-E-5
2ND STAGE O_2 COOLER
28.3 MM BTU/HR
24,788 SQ. FT
CARBON STEEL

11-1-E-6
3RD STAGE O_2 COOLER
33.4 MM BTU/HR
26,923 SQ. FT.
CARBON STEEL

11-1-E-7
4TH STAGE O_2 COOLER
17.5 MM BTU/HR
20,733 SQ. FT
CARBON STEEL

11-1-E-8
5TH STAGE O_2 COOLER
22.7 MM BTU/HR
22,551 SQ. FT.
CARBON STEEL



NOTES:

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

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

DATE	REVISION DESCRIPTION

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

C ABESAMIS
D BELMINO
KAHNWAIL
SPILLER
EPRI

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
CASE A2-BASE CASE (NO METHANOL, 1500 PSIG STEAM)
EPRI PALO ALTO, CALIFORNIA

SCALE NONE
DRAWING NUMBER 44833 4-EXTC(ME-A2)-11-1
REVISION 02

GASIFICATION AND ASH HANDLING

Process Flow Diagram EXTC(ME-A2)-20-1 shows the gasification, raw gas cooling, and particulate removal steps for Case A2. Three operating trains and one spare train are provided. The ash handling system is a single 100 percent capacity train. The 20-ME-2 box represents proprietary sections of the Texaco coal gasification process containing many units of equipment.

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

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

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

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

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

Energy Recovery

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

Raw gas leaving the high-pressure saturated steam generator is further cooled by heat exchange to generate saturated intermediate-pressure (IP) steam at 445 psig and saturated medium pressure (MP) steam at 115 psig in 20-1-E-2 and 20-1-E-3, respectively. The ash containing raw gas flows on the tube side to reduce solids deposition. Hot boiler feedwater at HP steam saturation temperature (598°F), and boiler feedwater streams at 459°F and 347°F are supplied from heat recovery steam generation (HRSG) units located in Unit 51. Exchangers 20-1-E-2 and 20-1-E-3 are kettle-type boilers with boiler feed water fed to the shell side.

Particulate Removal

The particulate bearing raw gas leaves the cooling unit and flows to the gas scrubbing unit 20-1-V-4. Ammonia absorber bottoms and hot process condensate from the gas cooling area (Flow Diagram EXTC(ME-A2)-21-1) are used for gas scrubbing. Water from 20-1-V-4 is recycled to 20-ME-2. The solids-free raw gas from 20-1-V-4 flows to the gas cooling section Unit 21. In subsequent sections of this report dealing with economics, the reader's attention is called to the fact that the costs for equipment included in the proprietary gas cooling and scrubbing units are included in the gas cooling system (Unit 21) costs.

Equipment Notes

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

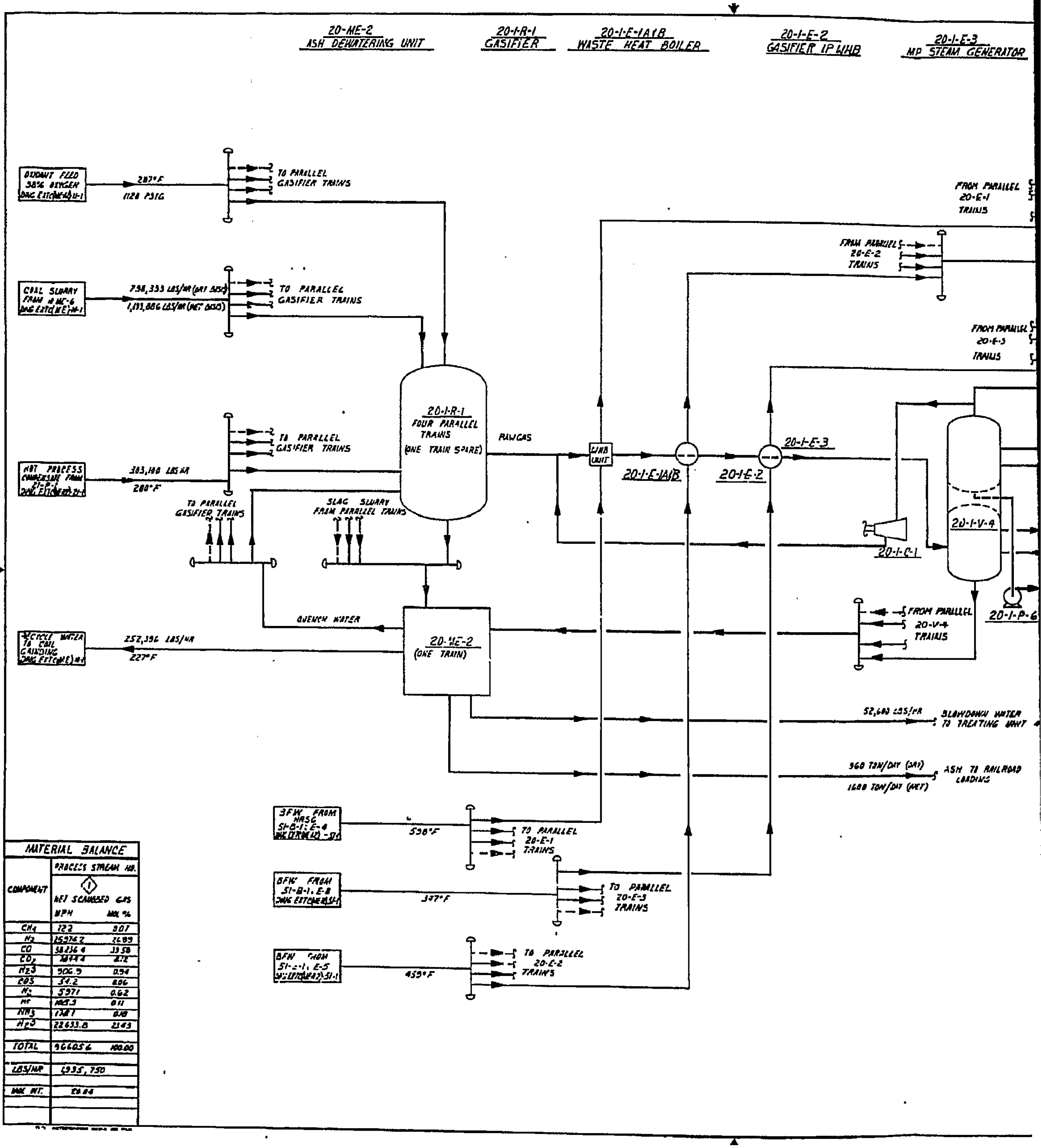
A coal gasifier having approximately one-half of the capacity of the gasifiers used for this study (when corrected for pressure effects) is currently in the final engineering design stages. This gasifier, to be constructed as part of the Cool Water Coal Gasification-Combined Cycle Demonstration Plant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

The slag dewatering system is composed of commercially proven equipment.

The gas scrubbing unit equipment is commercially available.

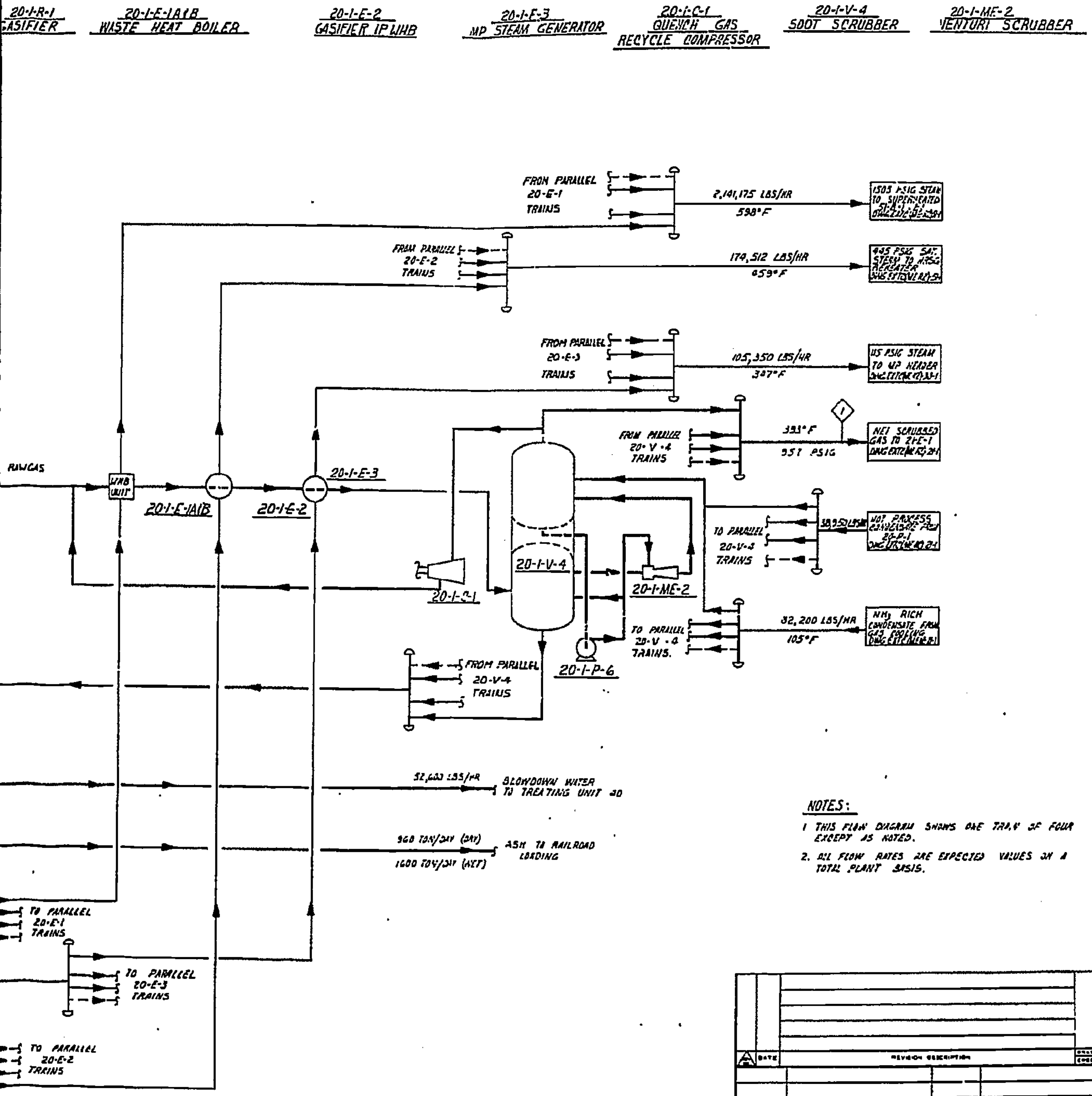
The key features in these designs center on the heat transfer equipment used for high-level sensible heat recovery. 1500 psig, saturated steam is generated in an unconventional fire-tube boiler, which is wholly conceptual at this point. A gasification process which is similar to the Texaco process has successfully superheated 750 psig steam for a very limited time in a pilot plant unit. The designs and cost estimates adopted in this study were developed by a major waste heat boiler manufacturer. It is also important to realize that the gas cooler design employed for this study is different from those being designed for the Cool Water Demonstration Plant.

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



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NO. ① NET SCRUBBED GAS MPH	MOL %
CH_4	72.2	0.07
H_2	25974.2	26.89
CO	38236.4	39.58
CO_2	[illegible]	[illegible]
H_2S	906.5	0.94
COS	54.2	0.06
N_2	597.1	0.62
Ar	105.3	0.11
NH_3	172.1	0.18
H_2O	22633.8	23.43
TOTAL	96605.6	100.00
LBS/HR	1,935,750	
MOL WT.	20.04	



NOTES:

1 THIS FLOW DIAGRAM SHOWS ONE TRAIN OF FOUR EXCEPT AS NOTED.

2. ALL FLOW RATES ARE EXPECTED VALUES ON A TOTAL PLANT BASIS.

DATE | REVISION DESCRIPTION

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

V FRIAS

O'BELMITO

RAVIEMAR

SHIRSER

EPRI

PROCESS FLOW DIAGRAM

COAL GASIFICATION / ASH HANDLING

CASE A2-BASE CASE (NO METHANOL, 1500 PSIG STM)

EPRI PALO ALTO, CALIFORNIA

NONE

448334 - EXTC (ME-A2) - 20-1

02

REDUCED PRINT SCALES

GAS COOLING

Process Flow Diagram EXTC(ME-A2)-21-1 shows one of the two parallel trains in the gas cooling section for Case A2. No spare train is provided.

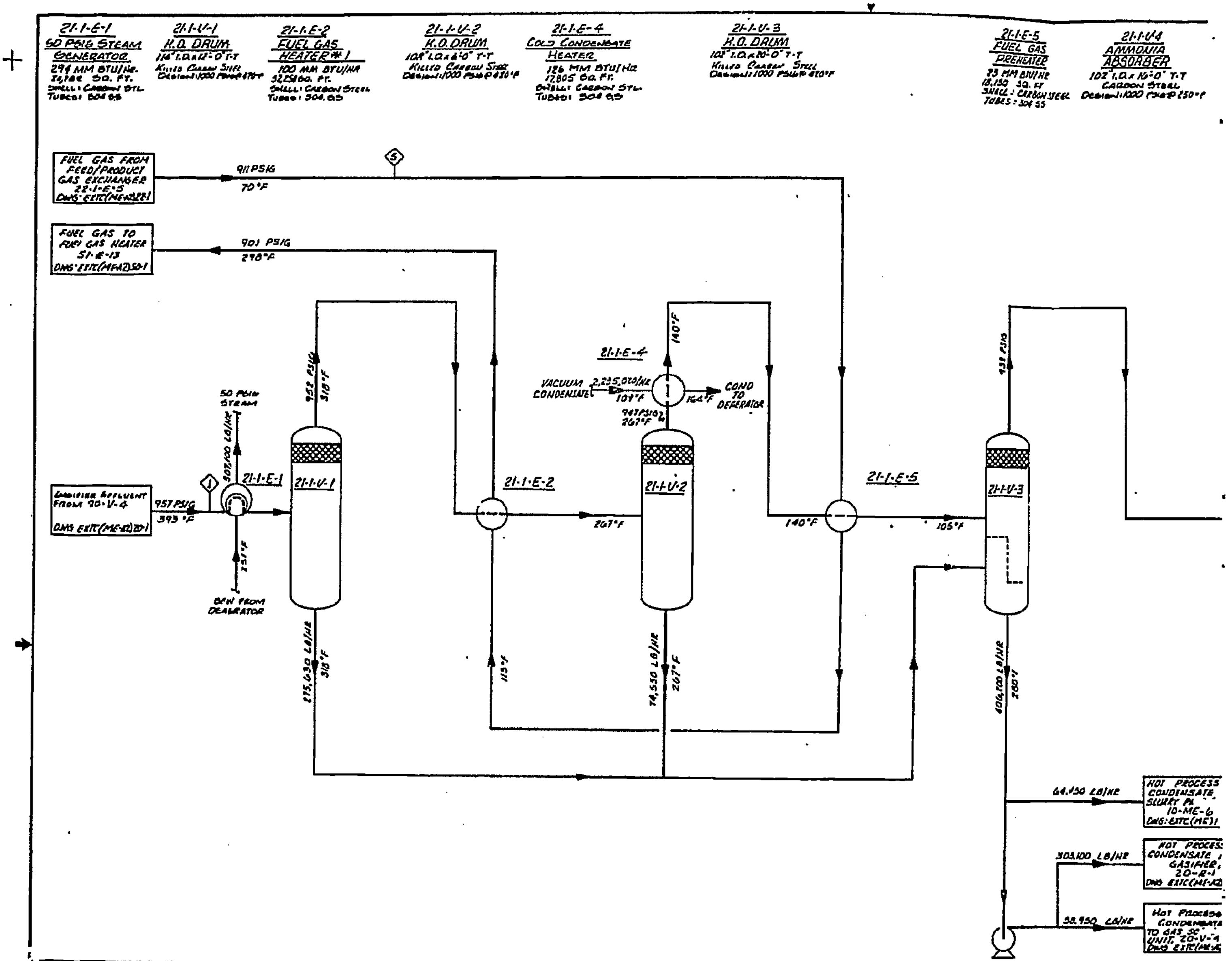
Clean gasifier effluent from the particulate scrubber 20-1-V-4 is cooled to 105°F on the tube side of a series of exchangers 21-1-E-1, 21-1-E-2, 21-1-E-4, and 21-1-E-5. Heat is recovered in exchanger 21-1-E-1 by the generation of saturated 50 psig steam. The effluent, after separation of condensate in the knockout drum 21-1-V-1, is then cooled by exchanging heat against fuel gas in 21-1-E-2. The condensate produced in cooling is separated in 21-1-V-2. Further gas cooling is obtained in exchanger 21-1-E-4 by heating vacuum condensate. The gas is then cooled in 21-1-E-5 against fuel gas from the acid gas removal unit. The resultant condensate is separated in knockout drum 21-1-V-3.

Condensate from knockout drums 21-1-V-1 and 21-1-V-2 flows to 21-1-V-3. Some of the combined hot condensate from 21-1-V-3 flows to the slurry preparation unit 10-ME-6, the remainder is pumped to the particulate scrubber 20-1-V-4 and to the gasifier 20-1-R-1 (Flow Diagram EXTC(ME-A2)-20-1).

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

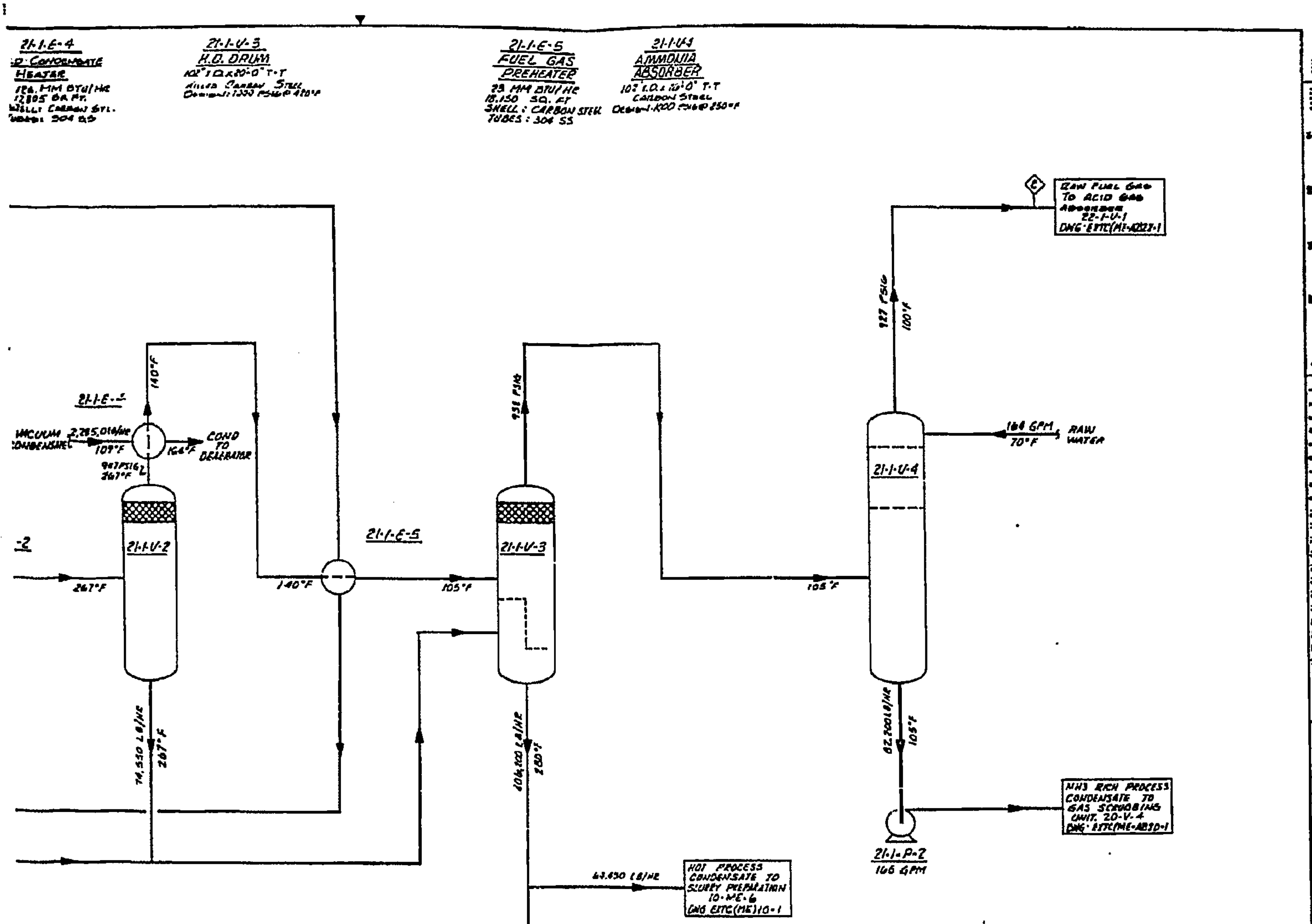
Equipment Notes

All equipment is commercially available.



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER ① PARTICULATE [illegible] MPH	MOL%	② AMMONIA SCRUBBER [illegible] MPH	MOL%	⑤ FUEL GAS MPH	MOL%
CH_4	72.1	0.07	77.8	0.10	71.5	0.10
H_2	25,971.8	26.87	25,174.2	33.16	25,971.9	36.80
CO	38,236.4	39.58	38,236.4	51.76	38,157.5	53.16
CO_2	7,884.4	8.19	7,884.4	10.38	6,743.4	9.41
H_2S	906.9	0.94	906.9	1.19	57.0	0.08
COS	54.2	0.06	54.2	0.07	38.1	0.05
N_2	597.1	0.62	597.1	0.81	595.5	0.83
Ar	108.3	0.11	108.3	0.15	108.0	0.15
NH_3	174.1	0.18	1.8	—	0.1	—
H_2O VAPOR	22,683.8	23.43	74.8	0.10	4.6	0.01
TOTAL VAPOR MPH	96,656.6	100.00	75,903	100.00	71,680.0	100.00
TOTAL LB/HR	1,935,750		1,576,947		1,441,545	
MOL WT VAPOR	20.04		20.66		20.13	



NOTES:

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

V. FRIAS
D'BELMONTE
KAVANAGH
SMELSER
EARL

PROCESS FLOW DIAGRAM
GAS COOLING
CASE A2-BASE CASE (NO METHANOL, 1500 PSIG STEAM)
EPRI
PALO ALTO, CALIFORNIA
NONE
418334- EXTC(ME-A2)21-1
02

ACID GAS REMOVAL

Process Flow Diagram EXTC(ME-A2)-22-1 depicts one of the two parallel acid gas removal trains for Case A2. No spare train is provided.

The acid gas removal system employs Allied Chemical Corporation's Selexol process for selective removal of hydrogen sulfide (H_2S) and carbonyl sulfide (COS). The H_2S and COS in the crude gas are absorbed in Selexol solvent to the extent that sulfur in the treated gas is reduced by 90 percent. The bulk of these sulfur emissions exit in the combined-cycle stack gas. A trace quantity is also discharged from the fuel gas treating unit stack.

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

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

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

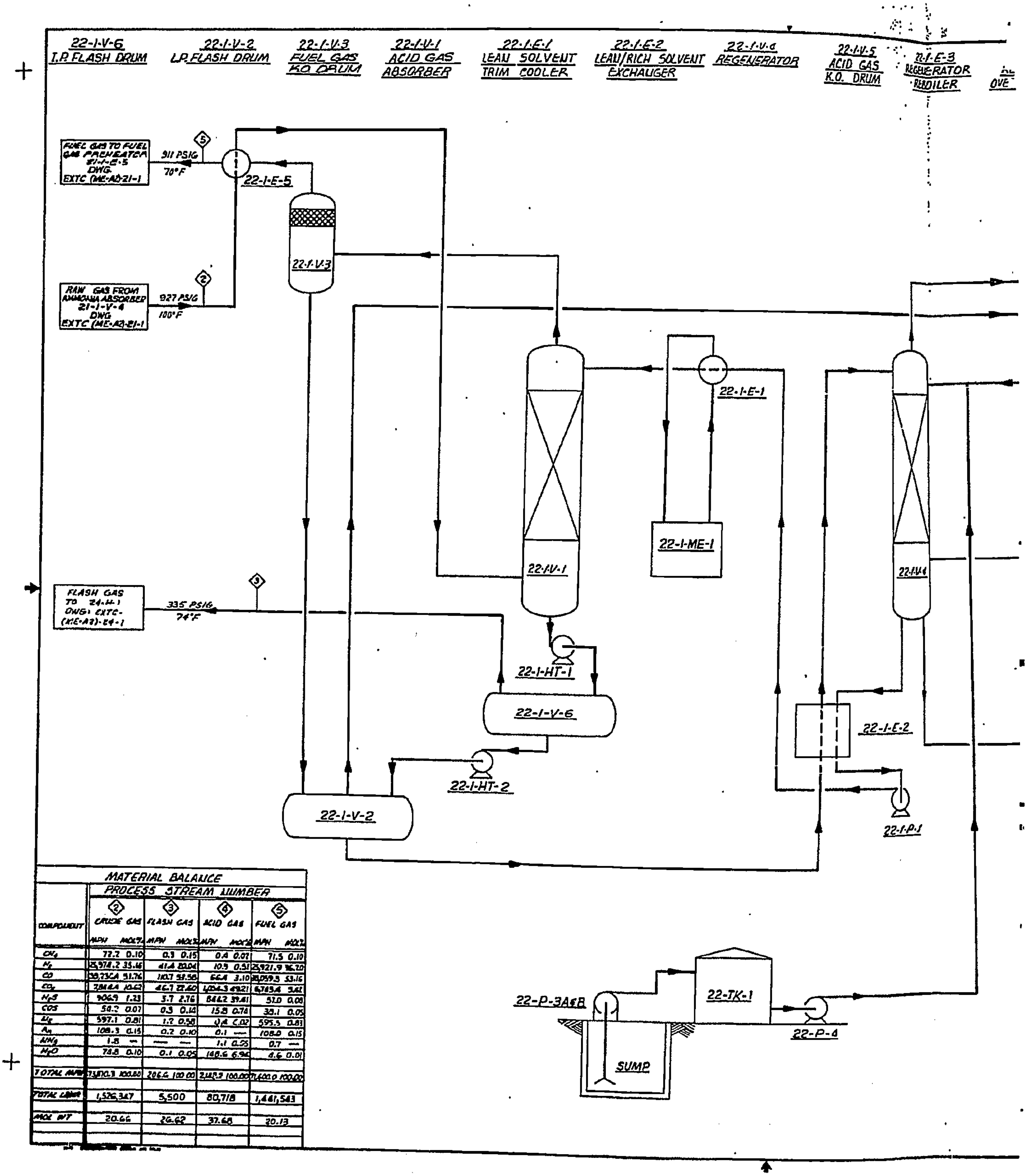
Acid gas from the regenerator overhead is cooled to 120°F in regenerator overhead condenser 22-1-E-4. Condensate resulting from this cooling step is separated in knockout drum 22-1-V-5 and then pumped back to the regenerator by 22-1-P-2. A small stream of demineralized water is added to the condensate at the discharge of 22-1-P-2 to maintain the water balance in the absorption system. The cooled acid gas from 22-1-V-5 contains about 40 percent H_2S on a volume basis and flows to the sulfur recovery unit for further processing.

Refrigeration System

The refrigeration system employed is a typical packaged fluorocarbon unit. The compressor, receiver, and condensing equipment are fabricated on skids and installed near lean solvent chiller 22-1-E-1. The capacity of the unit in each train is about 1900 tons of refrigeration duty.

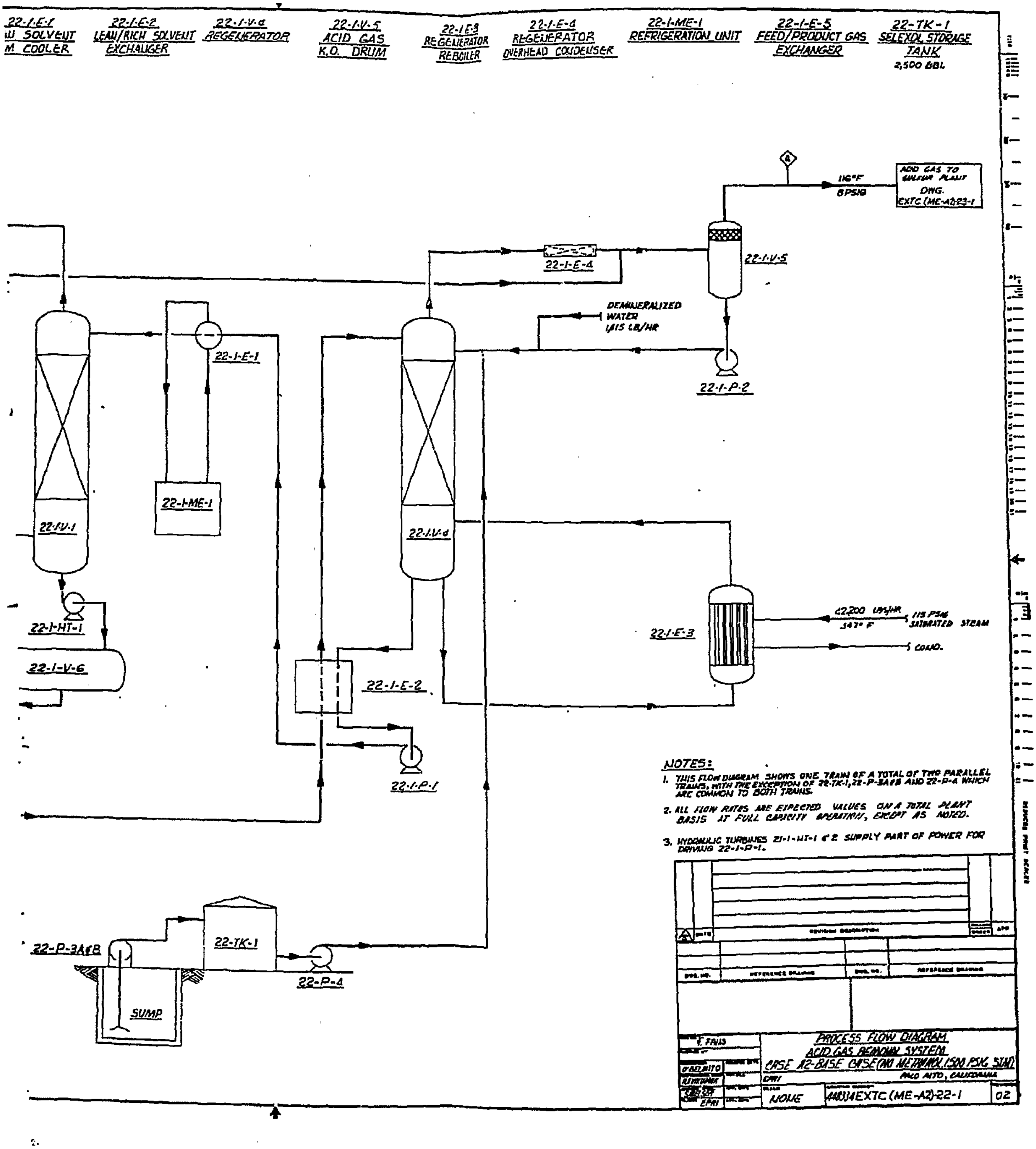
Equipment Notes

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



MATERIAL BALANCE

COMPOUND	PROCESS STREAM NUMBER ② CRUDE GAS MPH	MOL%	③ FLASH GAS MPH	MOL%	④ ACID GAS MPH	MOL%	⑤ FUEL GAS MPH	MOL%
CH_4	72.2	0.10	0.3	0.15	0.4	0.02	71.5	0.10
N_2	25,974.2	35.16	41.4	20.04	10.9	0.51	25,921.9	36.20
CO	38,236.4	51.76	110.7	53.58	66.4	3.10	38,059.5	53.16
CO_2	7,844.4	10.62	46.7	22.60	1,054.3	49.21	6,783.6	9.47
H_2S	906.9	1.23	5.7	2.76	844.2	39.41	57.0	0.08
COS	54.2	0.07	0.3	0.14	15.8	0.74	38.1	0.05
H_2	597.1	0.81	1.2	0.58	0.4	0.02	595.5	0.83
Ar	108.3	0.15	0.2	0.10	0.1	—	108.0	0.15
NH_3	1.8	—	—	—	1.1	0.05	0.7	—
H_2O	74.8	0.10	0.1	0.05	148.6	6.94	4.6	0.01
TOTAL MPH	73,870.3	100.00	206.6	100.00	2,142.2	100.00	71,600.0	100.00
TOTAL LB/HR	1,526,347		5,500		80,718		1,441,543	
MOL WT	20.66		26.62		37.68		20.13	



NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF TWO PARALLEL TRAINS, WITH THE EXCEPTION OF 22-TK-1, 22-P-3A&B AND 22-P-4 WHICH ARE COMMON TO BOTH TRAINS.
2. ALL FLOW RATES ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.
3. HYDRAULIC TURBINES 21-1-HT-1 & 2 SUPPLY PART OF POWER FOR DRIVING 22-1-P-1.

PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
CASE A2-BASE CASE (NO METHANOL, 1500 PSIG SIM)
EPRI
PALO ALTO, CALIFORNIA
NONE
448334EXTC (ME-A2)-22-1
02

SULFUR RECOVERY

Process Flow Diagram EXTC(ME-A2)-23-1 describes the basic sulfur plant design used. The entire sulfur plant system for Case A2 is composed of three parallel, 50 percent capacity, sulfur recovery trains. Two operating trains and one spare train are provided for increased reliability due to the important environmental requirements this unit fulfills. Sulfur recovery is 154.1 short tons per day per train.

The sulfur recovery unit is a two-stage acid gas bypass type Claus unit. About one-third of the 120°F gas from the acid gas removal (Selexol) unit is burned in a sulfur furnace 23-1-H-1, to convert H_2S to SO_2. Air for combustion in the furnace is supplied by blower 23-1-BL-1. Heat from the combustion products is recovered by generating 445 psig steam, in waste heat boiler 23-1-E-1. The 900°F exhaust gas from the sulfur furnace is mixed with the acid gas bypass stream and the resultant 606°F gas is fed to the sulfur converter No. 1, 23-1-R-1. The amount of acid gas bypassing the furnace is controlled to maintain a ratio of H_2S to SO_2, slightly more than the 2:1 stoichiometric ratio required for the sulfur formation reactions.

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

$$2\ H_2S + 1\ SO_2 \rightarrow 3\ S + 2\ H_2O \tag{3-1}$$

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

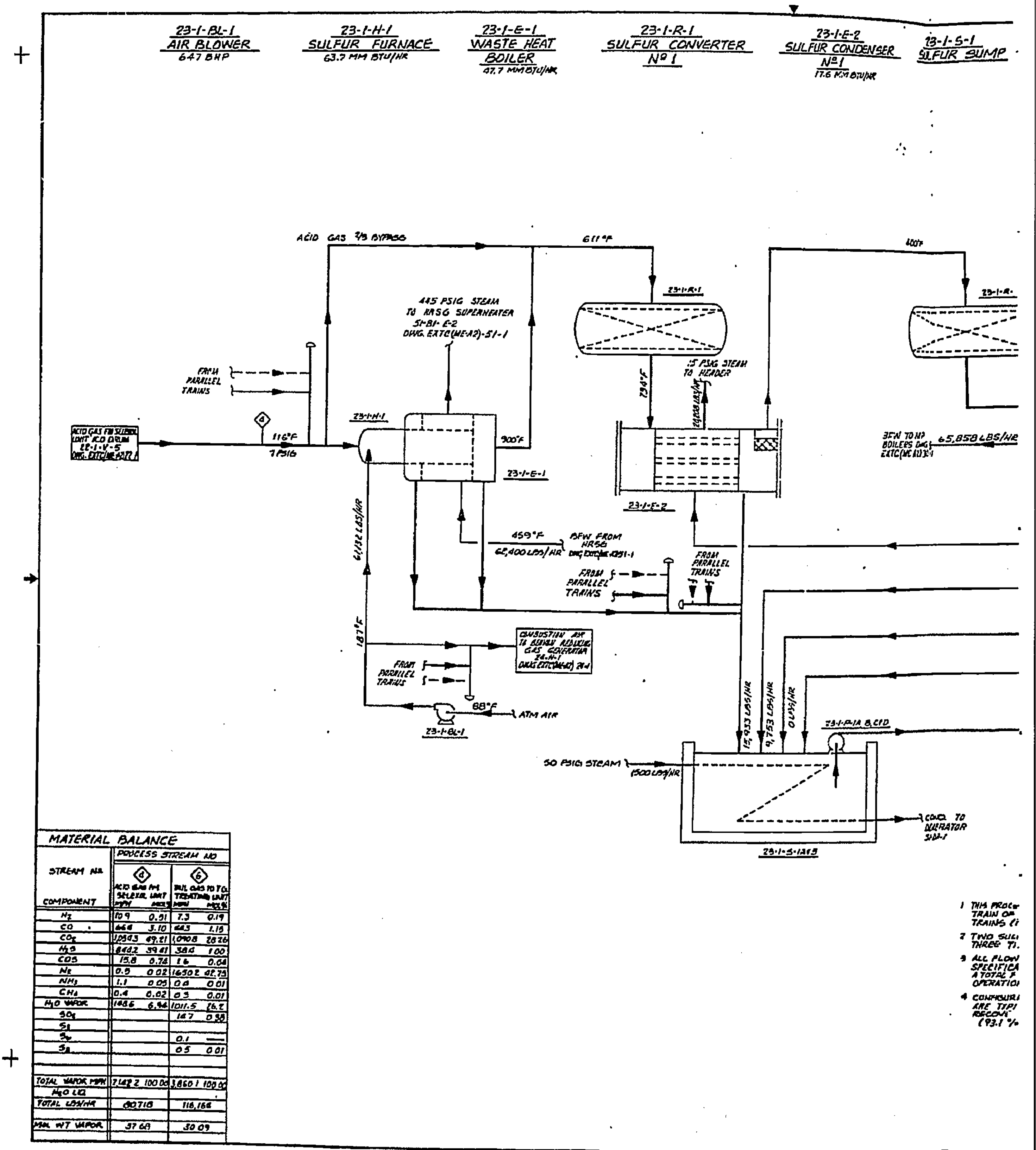
Converter effluent gas is cooled below its sulfur dew point in sulfur condenser 23-1-E-2 by generating 115 psig steam from boiler feedwater. Condensed sulfur flows by gravity to a concrete sulfur sump 21-S-1A&B. Since sulfur is a solid at ambient temperature, it must be heated in the sump to take advantage of liquid phase transport to loading facilities. The sump contains low-pressure steam coils to maintain sulfur in its molten state.

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

Tail gas at 285°F, still containing about 1900 lb/hr sulfur (mainly as H_2S, with smaller amounts of SO_2, COS, and elemental sulfur), flows through coalescer 23-1-V-1 and then enters Beavon/Stretford Unit 24 for final sulfur recovery to preserve air quality.

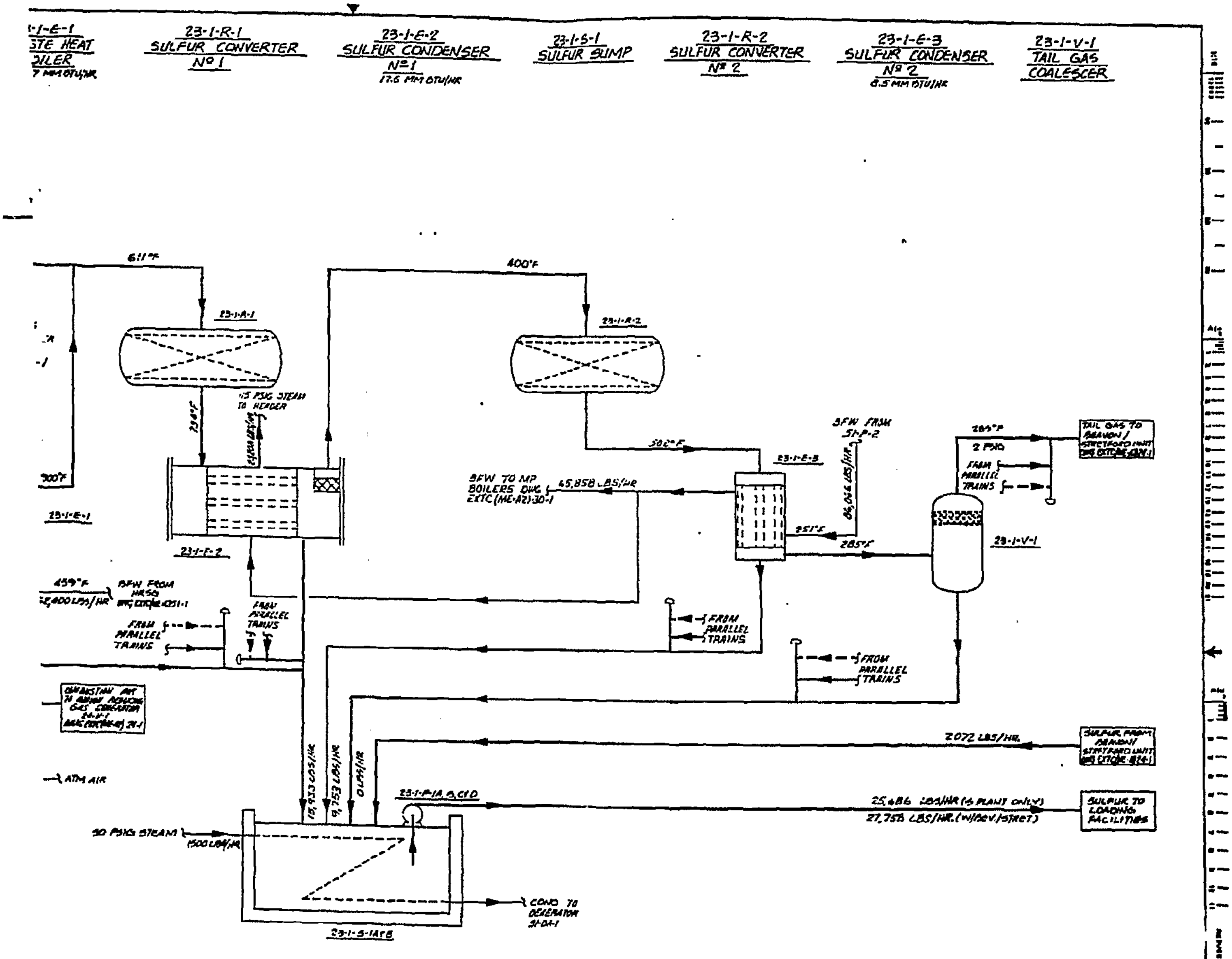
Equipment Notes

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



MATERIAL BALANCE

STREAM NO. COMPONENT	PROCESS STREAM NO 4 ACID GAS FM SELEXOL UNIT MPH	 MOL%	 6 TAIL GAS TO TG TREATING UNIT MPH	 MOL%
N_2	10.9	0.51	7.3	0.19
CO	664	3.10	443	1.15
CO_2	10543	49.21	10908	28.26
H_2S	8482	39.41	38.6	1.00
COS	15.8	0.74	1.6	0.04
N_2	0.5	0.02	1650.2	42.75
NH_3	1.1	0.05	0.4	0.01
CH_4	0.4	0.02	0.3	0.01
H_2O VAPOR	1486	6.94	1011.5	26.2
SO_2			14.7	0.38
S_2				
S_6			0.1	—
S_8			0.5	0.01
TOTAL VAPOR MPH	21422	100.00	3860.1	100.00
H_2O LIQ				
TOTAL LBS/HR	80710		116,166	
MOL WT VAPOR	37.68		30.09	



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

2 TWO SULFUR SUMPS WILL SERVICE THE THREE TRAINS

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

4 CONFIGURATION AND SIZE OF EQUIPMENT ARE TYPICAL FOR AN EXPECTED SULFUR RECOVERY RATE OF 154.1 ST/D PER TRAIN (93.1 % RECOVERY)

V FRIAS
G BELMIRO
RAYMUNDO
SHELTER
EPRI

PROCESS FLOW DIAGRAM
SULFUR PLANT (TYPICAL)
CASE A2-BASE CASE (NO METHANOL, 1500 PSIG STEAM)
EPRI
PALO ALTO, CALIF.
NONE
448334-EXTC(ME-A2)-23-1
01

TAIL GAS TREATING

Process Flow Diagram EXTC-(ME-A2)-24-1 describes the Beavon/Stretford system design used for the two oxygen-blown GCC plants in the study. As in the sulfur recovery unit, two 50 percent parallel operating trains and a third identical spare train are provided.

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

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

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

$$SO_2 + 3\ H_2 \rightarrow H_2S + 2\ H_2O \tag{3-2}$$

$$COS + H_2O \rightarrow CO_2 + H_2S \tag{3-3}$$

$$S_6 + 6\ H_2 \rightarrow 6\ H_2S \tag{3-4}$$

$$S_8 + 8\ H_2 \rightarrow 8\ H_2S \tag{3-5}$$

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

$$CO + H_2O \rightarrow CO_2 + H_2 \tag{3-6}$$

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

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

$$2\ Na_2CO_3 + 2\ H_2S \rightarrow 2\ NaHCO_3 + 2\ NaHS \tag{3-7}$$

$$2\ NaHS + 2\ NaHCO_3 + 4\ NaVO_3 \rightarrow 2\ Na_2CO_3 + H_2O + S_2 + Na_2V_4O_9 + 2\ NaOH \tag{3-8}$$

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

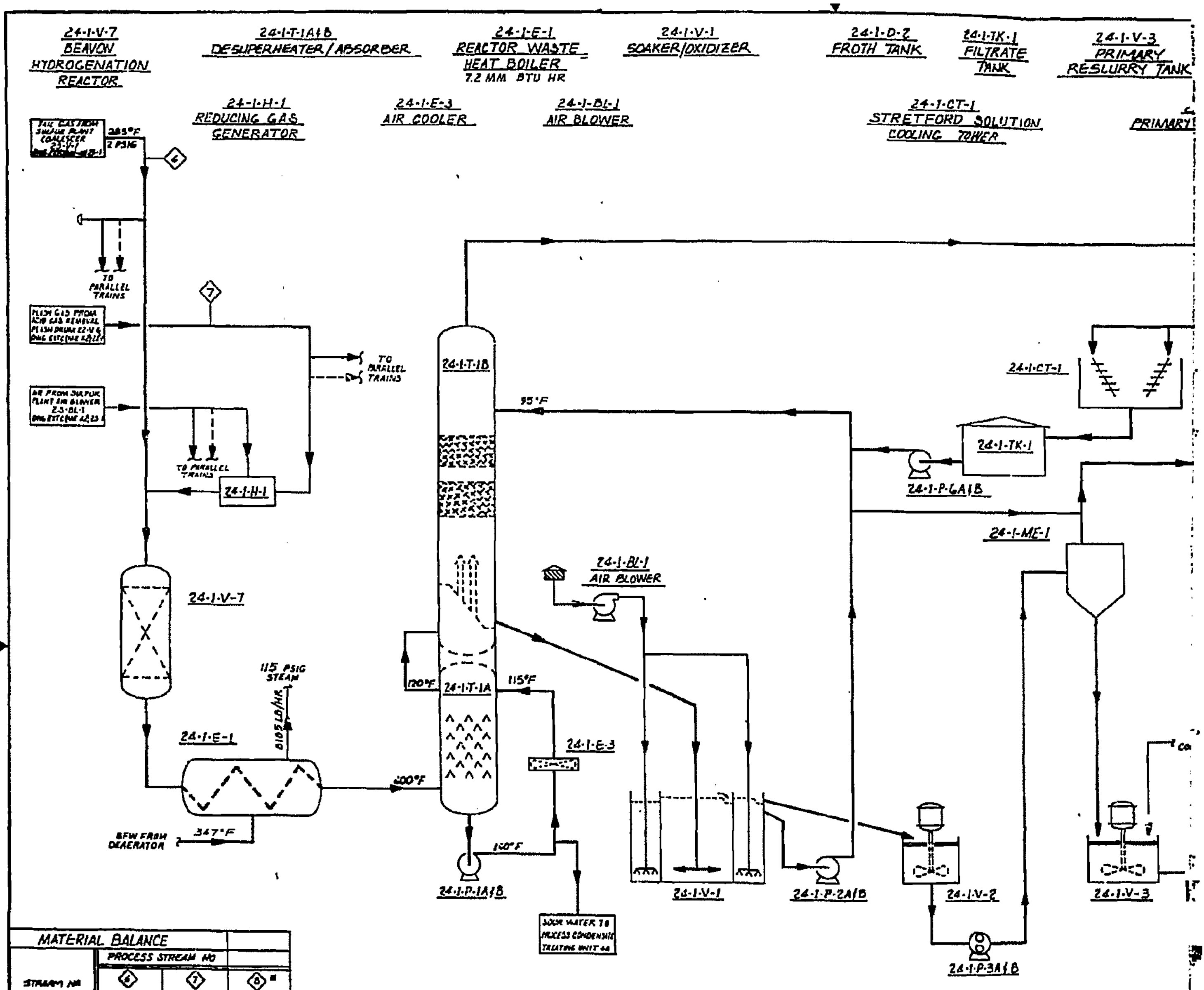
The reacted Stretford solution flows to soaker/oxidizer 24-1-V-1, where the reduced vanadate ($Na_2V_4O_9$) is oxidized to its original form by anthraquinone disulfonic acid (ADA) in the solution. The reduced ADA is subsequently regenerated by air sparged into the tank by blower 24-1-BL-1. The air also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

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

The sulfur from the secondary centrifuge is reslurried in 24-1-V-4 and pumped to the sulfur separator 24-1-EJ-1, where sulfur is melted with heat supplied by 115 psig steam in coils. Molten sulfur (2081 lb/hr) is separated from the slurry medium (primarily water) in sulfur separator 24-1-V-5. From 24-1-V-5 it flows by gravity, into one of the two sumps located in Unit 23. The decanted water flows to flash drum 24-1-V-6 and then back to the secondary reslurry tank. Because certain side reactions degrade the Stretford solution, a small stream of liquid is continuously discarded from the system and pumped to effluent water treating, Unit 40.

Equipment Notes

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



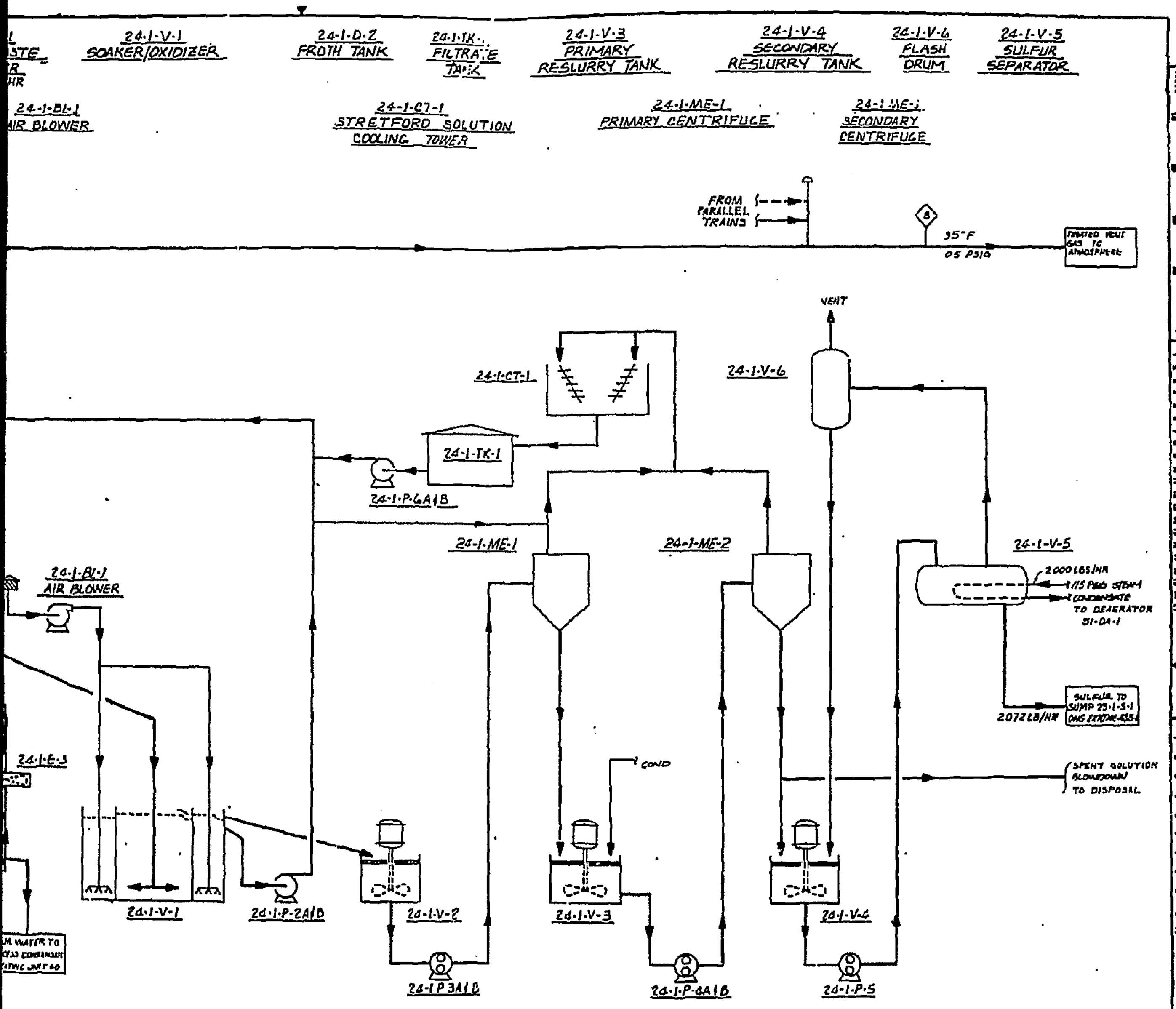
MATERIAL BALANCE

STREAM NO / COMPONENT	PROCESS STREAM NO 6 TAIL GAS FM SULFUR PLANT MPH	MOL %	7 FLASH GAS MPH	MOL %	8* VENT GAS MPH	MOL %
H_2	7.3	0.19	41.4	20.04	50.5	1.53
CO	44.5	1.15	110.7	53.58	2.5	0.08
CO_2	1070.8	28.76	46.7	22.60	1292.1	39.12
H_2S	38.4	1.00	5.7	2.76	0.4	0.01
COS	1.6	0.04	0.3	0.14		
N_2	1630.2	42.35	1.4	0.66	1651.6	50.01
NH_3	0.4	0.01				
CH_4	0.3	0.01	0.3	0.15	0.4	0.01
H_2O VAPOR	1011.5	26.3	0.1	0.05	305.2	9.24
SO_2	14.1	0.38				
S_6						
S_7	0.1	–				
S_8	0.5	0.01				
TOTAL VAPOR MPH	3840.1	100.00	206.6	100.00	3307.4	100.00
H_2O LIQ						
TOTAL LB/HR	116,164		5,500		108,826	
MOL WT VAPOR	30.09		26.62		32.9	

* STATED AS H_2S BUT CONSISTS OF TOTAL SULFUR.

NOTES:

1. THIS PROCESS FLOW DIAGRAM SHOWS O
TRAIN OF A TOTAL OF THREE 50%
(TWO OPERATING AND ONE SPARE).
2. ALL FLOW RATES AND EQUIPMENT
ARE TYPICAL VALUES ON A TOTAL .
AT FULL CAPACITY OPERATION EXCEPT
3. CONFIGURATION AND SIZE OF EQUIPMENT
FOR AN EXPECTED RATE OF 12.4
TRAIN SULFUR RECOVERY.



NOTES:

1. THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE 50% PARALLEL TRAINS (TWO OPERATING AND ONE SPARE).
2. ALL FLOW RATES AND EQUIPMENT SPECIFICATIONS ARE TYPICAL VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
3. CONFIGURATION AND SIZE OF EQUIPMENT ARE TYPICAL FOR AN EXPECTED RATE OF 12.4 ST/D PER TRAIN SULFUR RECOVERY.

V FALLS
D'ALLURITO
RAVINDRA
SALSER
EPRI

PROCESS FLOW DIAGRAM
BEAVON/STRETFORD UNIT (TYPICAL)
CASE A2-BASE CASE (NO METHANOL, 1500 PSIG SIM)
EPRI
PALO ALTO, CALIFORNIA

NONE
448334-EXTO(ME-A2)24-1
03

STEAM, BOILER FEEDWATER, AND CONDENSATE

Process Flow Diagram EXTC(ME-A2)-30-1 schematically represents the steam, boiler feedwater (BFW), and condensate systems for Case A2.

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 at the 51-T-1A turbine inlet
● Intermediate Pressure (IP)	445 psig, 900°F at the 51-T-1B turbine inlet
● Medium-Pressure (MP)	115 psig
● Low-Pressure (LP)	50 psig at the 51-T-3 turbine inlet
● Very Low Pressure (VLP)	15 psig for consumption in the deaerator

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

Saturated intermediate pressure (IP) steam at 445 psig is obtained from the IP steam generators located in the sulfur plant, the gas cooling unit 20-1-E-2, and from the gas turbine air cooler 50-1-E-1. The saturated IP steam, together with the exhaust steam from 51-T-1A is superheated to 900°F in the HRSG reheater 51-1-B-1:E-2. The superheated IP steam at 385 psig, 900°F is then used in the IP end 51-T-1B, a machine of 85.4 percent isentropic efficiency. The low-pressure end of 51-T-1B exhausts steam at 93.8 psig.

Steam for the 115 psig header is obtained from steam generator 20-1-E-3 in gas cooling unit, from the sulfur plant, and from the tail gas treating unit. A portion of the 115 psig steam is supplied to the sulfur heater and the acid gas removal unit reboiler. The remainder is combined with 51-T-1B exhaust at

93.8 psig for consumption in MP turbine 51-T-2, and in BFW pump turbine 51-T-4. The MP turbine and the BFW pump driver are condensing turbines exhausting at 2-1/2 inches Hg absolute. 51-T-2 has an isentropic efficiency of 87.4 percent.

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

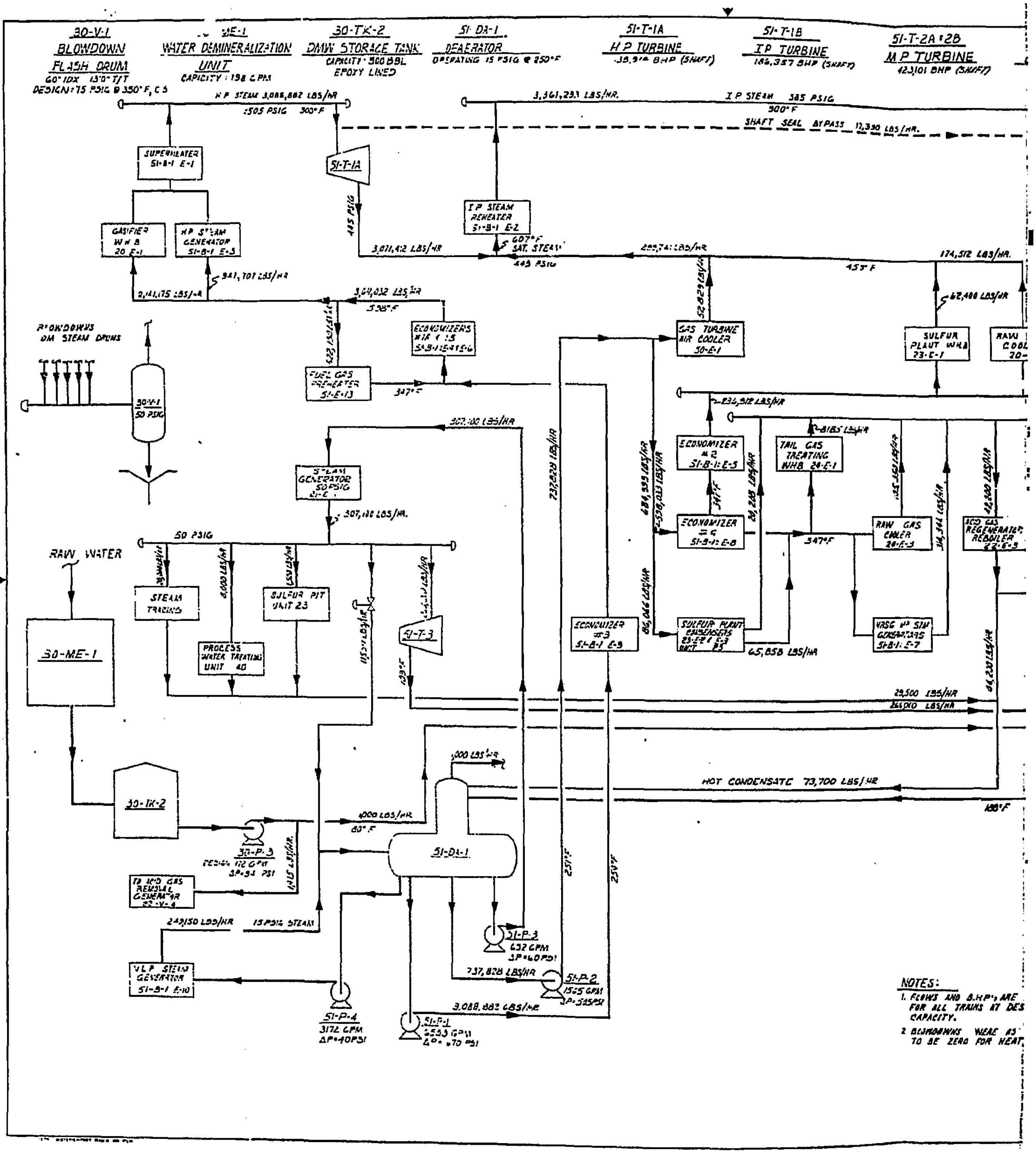
15 psig steam is supplied by steam generation in HRSG coil 51-1-B-1:E-10. This very low-pressure steam is used entirely in deaerator 51-DA-1.

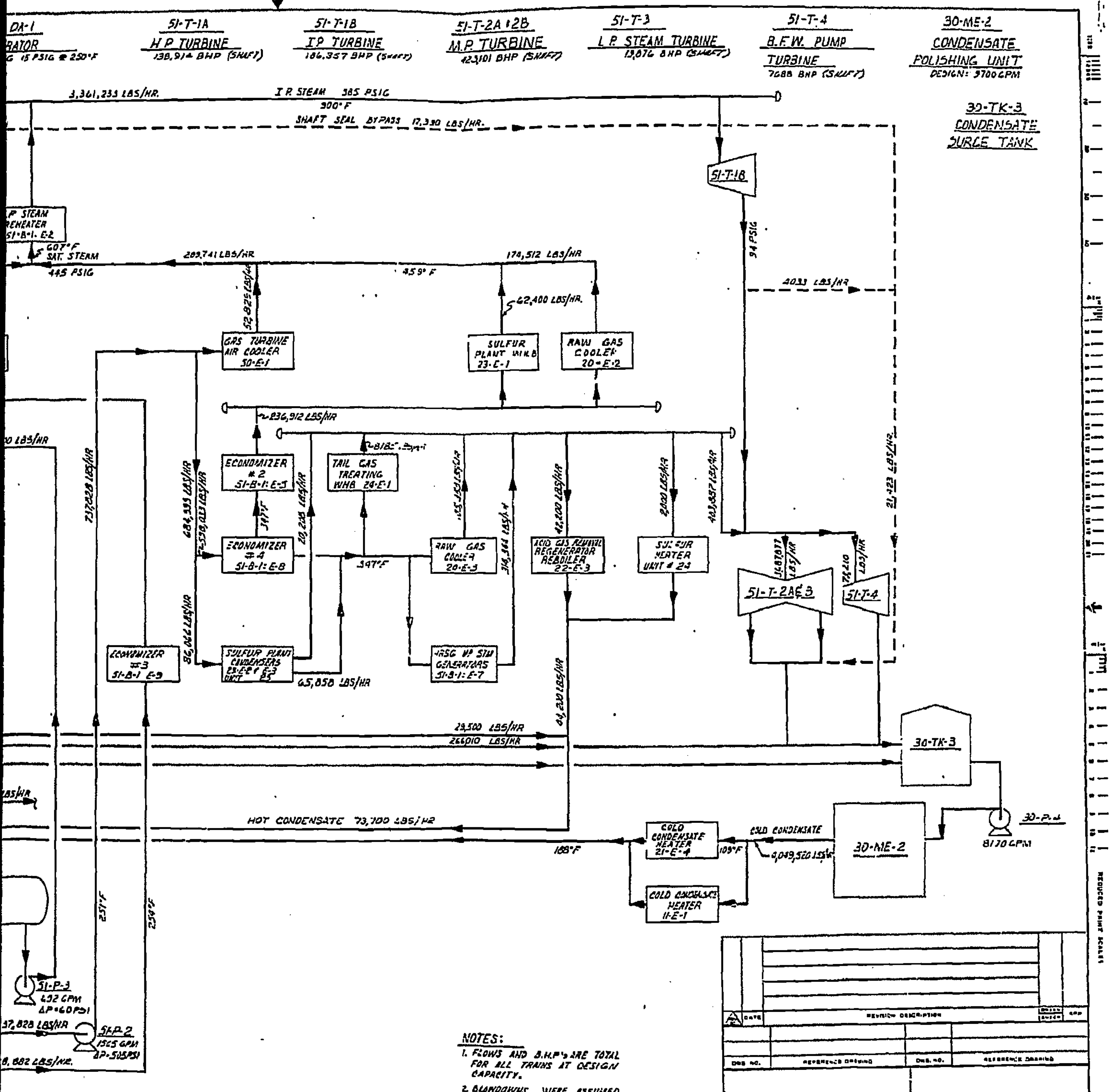
Raw water is treated in an automatic ion exchange demineralizer 30-ME-1, consisting of three strong-acid cation columns; one degasifier (with 10-minute holdup vessel), and three strong-base anion columns. Two of the three cation and anion columns can handle the design flow of raw water, either for the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

The turbine surface condensers 51-E-11, 51-E-12 and 51-E-14 are single-shell single-pass units with divided water boxes. The tubes are 90/10 copper/nickel, 7/8 inch OD, 22 BWG wall thickness. The noncondensable gas removal and priming equipment includes positive displacement rotary vacuum pumps and a recirculating ball-type condenser tube cleaning system. Motor-driven condensate pumps 51-P-8, 51-P-9 and 51-P-10 respectively transport the condensate to condensate storage tank 30-TK-3, which is sized for 30-minute capacity at design flow rate.

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

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


30-V-1 BLOWDOWN FLASH DRUM 60" IDX 13'0" T/T DESIGN: 75 PSIG @ 350°F, C.S
ME-1 WATER DEMINERALIZATION UNIT CAPACITY: 198 GPM
30-TK-2 DMW STORAGE TANK CAPACITY: 500 BBL EPOXY LINED
51-DA-1 DEAERATOR OPERATING 15 PSIG @ 250°F
51-T-1A H P TURBINE
51-T-1B I P TURBINE 186,357 BHP (SHAFT)
51-T-2A & 2B M P TURBINE 423,101 BHP (SHAFT)
H P STEAM 3,088,882 LBS/HR 1505 PSIG 900°F
3,361,233 LBS/HR.
I P STEAM 385 PSIG 900°F
SHAFT SEAL BYPASS 11,350 LBS/HR.
SUPERHEATER 51-B-1 E-1
51-T-1A
445 PSIG
I P STEAM REHEATER 51-B-1 E-2
607°F SAT. STEAM 445 PSIG
3,071,412 LBS/HR
289,741 LBS/HR
455°F
174,512 LBS/HR.
GASIFIER W H B 20 E-1
H P STEAM GENERATOR 51-B-1 E-3
947,707 LBS/HR
2,141,175 LBS/HR
3,611,032 LBS/HR
538°F
BLOWDOWNS OM STEAM DRUMS
ECONOMIZERS #1 & 3 51-B-1 E4 & 6
FUEL GAS PREHEATER 51-E-13
347°F
30-V-1 50 PSIG
GAS TURBINE AIR COOLER 50-E-1
62,488 LBS/HR
SULFUR PLANT WHB 23-E-1
RAW GAS COOLER 20-
236,512 LBS/HR
307,100 LBS/HR
STEAM GENERATOR 50 PSIG
307,100 LBS/HR.
ECONOMIZER #2 51-B-1 E-5
TAIL GAS TREATING WHB 24-E-1
8185 LBS/HR
ECONOMIZER #5 51-B-1 E-8
347°F
RAW GAS COOLER 20-E-3
ACID GAS REGENERATOR REBOILER 22-E-3
50 PSIG
RAW WATER
STEAM TRACING
8,000 LBS/HR
SULFUR PIT UNIT 23
PROCESS WATER TREATING UNIT 40
51-T-3
30-ME-1
ECONOMIZER #3 51-B-1 E-9
SULFUR PLANT CONDENSERS 23-E-2 & E-3 UNIT 25
65,858 LBS/HR
VRSC LP STM GENERATORS 51-B-1 E-7
737,828 LBS/HR
29,500 LBS/HR
265,010 LBS/HR
HOT CONDENSATE 73,700 LBS/HR
1000 LBS/HR
30-TK-2
30-P-3 DESIGN 112 GPM ΔP=94 PSI
4000 LBS/HR 60°F
51-DA-1
188°F
TO ACID GAS REMOVAL GENERATOR 22-V-4
249,150 LBS/HR
15 PSIG STEAM
VLP STEAM GENERATOR 51-B-1 E-10
51-P-3 632 GPM ΔP=60 PSI
737,828 LBS/HR
51-P-2 1525 GPM ΔP=325 PSI
3,088,882 LBS/HR
51-P-4 3172 GPM ΔP=40 PSI
51-P-1 6553 GPM ΔP=670 PSI
251°F
250°F
NOTES:
1. FLOWS AND B.H.P's ARE FOR ALL TRAINS AT DES CAPACITY.
2. BLOWDOWNS WERE AS TO BE ZERO FOR HEAT



NOTES:

1. FLOWS AND B.H.P's ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. BLOWDOWNS WERE ASSUMED TO BE ZERO FOR HEAT BALANCE.

DATE	REVISION DESCRIPTION		
DWG. NO.	REFERENCE DRAWING	DWG. NO.	REFERENCE DRAWING

V FRIAS
OBELMILO
Raviraopal
Sanser
EPRI

PROCESS FLOW DIAGRAM
STEAM B.F.W. AND CONDENSATE SYSTEM
CASE-A2-BASE CASE (NO METHANOL, 1500PSIG STEAM)
EPRI
NONE
448334 EXTC (ME-A2)-30-1
02

COMBUSTION GAS TURBINE

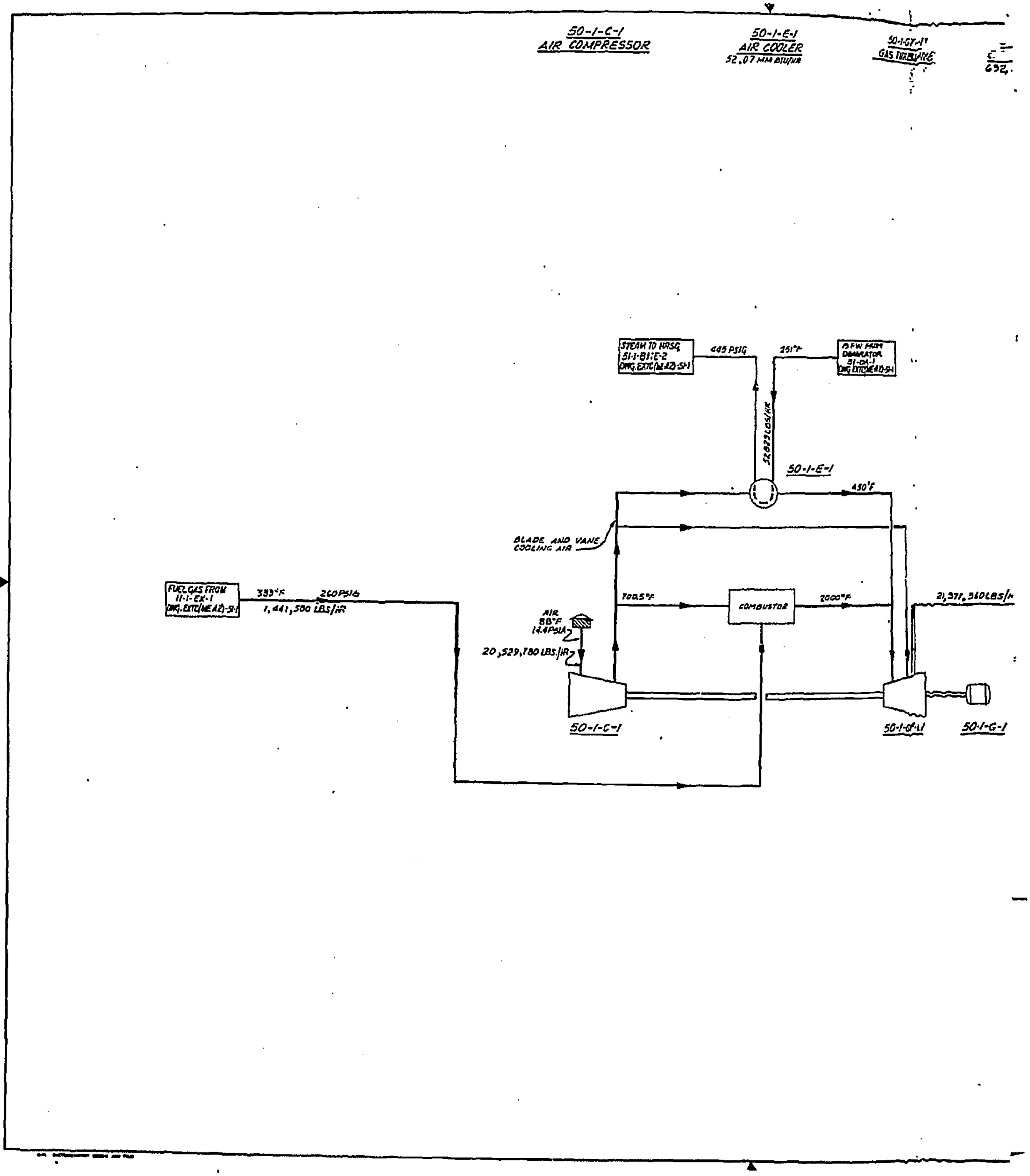
Process Flow Diagram EXTC(ME-A2)-50-1 shows one of the seven parallel combustion gas turbines for Case A2. No spare turbine is provided.

Fuel gas from the fuel gas expander, 51-1-EX-1, at 245 psig and 339°F flows to the gas turbine combustor at 245 psig where it is burned with excess air supplied by air compressor 50-1-C-1. Effluent gases exit the combustor at 2000°F and flow to the combustion gas turbine 50-1-GT-1. A small fraction of compressed air is cooled by IP steam generation in 50-1-E-1 before being injected into the turbine to cool the rotors.

The combustion gases are expanded in the combustion gas turbine, producing 692.25 MW net power in generator 50-1-G-1. The effluent gases at 967°F flow to the heat recovery steam generator (HRSG) in Unit 51. The turbine drives the air compressor and electric generator 50-1-G-1. Detailed performance information on the combustion gas turbine is presented in Appendix A.

Equipment Notes

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

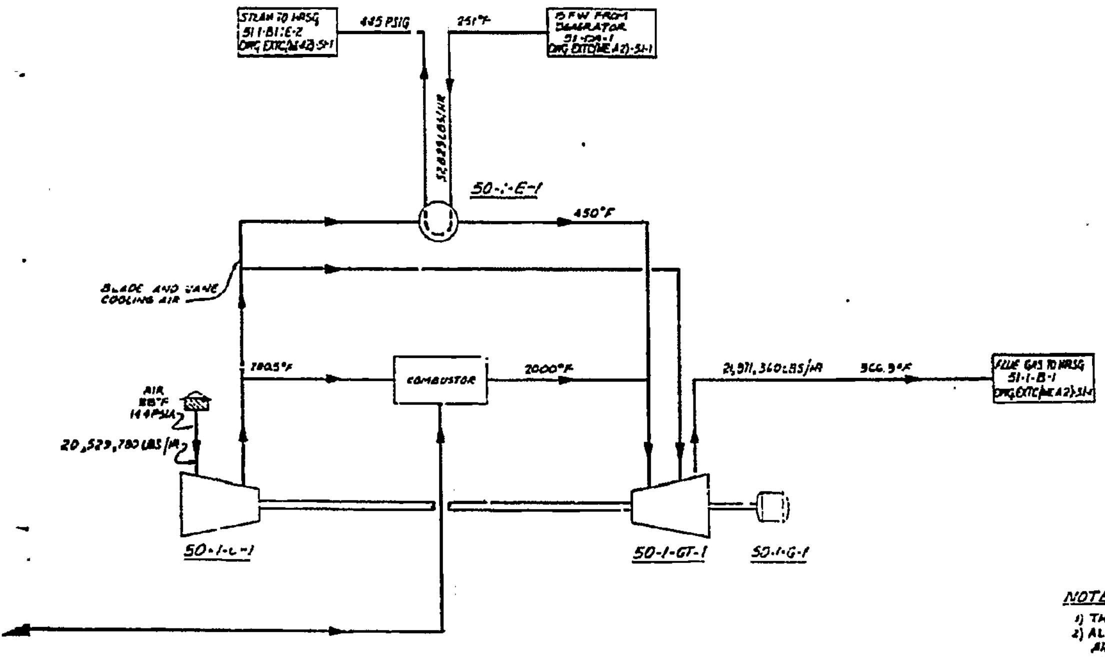

50-1-C-1
AIR COMPRESSOR
50-1-E-1
AIR COOLER
52.07 MM BTU/HR
50-1-GT-1
GAS TURBINE
692.
STEAM TO HRSG
51-1-B1:E-2
DWG. EXTC(ME AZ)-SH1
445 PSIG
251°F
BFW FROM
DEAERATOR
51-DA-1
DWG EXTC(ME AZ)-SH1
52823 LBS/HR
50-1-E-1
450°F
BLADE AND VANE
COOLING AIR
FUEL GAS FROM
11-1-EX-1
DWG. EXTC(ME AZ)-SH1
333°F
260PSIG
1,441,580 LBS/HR
700.5°F
COMBUSTOR
2000°F
21,371,360 LBS/H
AIR
88°F
14.4PSIA
20,529,780 LBS./HR
50-1-C-1
50-1-GT-1
50-1-G-1

50-1-C-1
AIR COMPRESSOR

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

50-1-GT-1
GAS TURBINE

50-1-G-1
GENERATOR
632,250 KW(NET)



NOTES:

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

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

V FRIAS
D'BELMIRO
RAYMOND
SHRISER
EPRI

PROCESS FLOW DIAGRAM
COMBUSTION GAS TURBINE
CASE A2-BASE CASE (NO METHANOL, 1200 PSIG STM
EPRI
PALO ALTO, CALIFORNIA

NONE
448334-EXTC(ME-A2)-50-1
0

HEAT RECOVERY STEAM GENERATOR AND STEAM TURBINES

Process Flow Diagram EXTC-(ME-A2)-51-1 shows the heat recovery steam generators (HRSG) 51-1-B-1 and the steam turbines for Case A2. There are fourteen operating HRSG units, one primary steam turbine 51-T-1A, 1B and 2, and one secondary steam turbine 51-T-3. The primary and secondary steam turbines drive generators for production of electric power. Steam turbine 51-T-4 is used to drive the high-pressure boiler feedwater pump. Additional electric power is generated by the expansion of high-pressure fuel gas to 245 psig in seven fuel gas expanders 51-1-EX-1. No spare turbines or HRSGs are provided.

Two HRSGs are coupled with each gas turbine to recover heat from the turbine exhaust gas, which leaves the turbines at 967°F. Radiation heat losses occur throughout the HRSG and are conservatively assumed to be realized immediately following the gas turbines, such that the HRSG flue gas inlet temperature is 960°F. The HRSG performs superheating, high-pressure (HP), medium-pressure (MP), very low-pressure (VLP) steam generation, and boiler feedwater heating. The arrangement of the heat recovery sections of the HRSG in the direction of flue gas flow is as follows:

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

Saturated HP steam from 20-1-E-1 and saturated HP steam from the HP evaporator is superheated to 900°F in the HRSG superheater 51-1-B-1:E-1. The HRSG superheater outlet supplies the HP feed of back-pressure steam turbine 51-T-1A. Expanded steam from 51-T-1A combines with process generated saturated IP steam and is reheated to 900°F in 51-1-B-1:E-2. This steam supplies the feed to IP back-pressure turbine 51-T-1B. Saturated MP steam generated in process areas and in MP evaporator 51-1-B-1:E-7 combines with the IP turbine exhaust to drive both the MP power turbine 51-T-2 and the HP BFW pump turbine 51-T-4. These are condensing turbines exhausting at 2-1/2 inches Hg absolute.

HP BFW is preheated to 347°F in economizer 4. Heated to 410°F in economizer 1B, and further heated to saturation temperature 598°F, in economizer 1A. Both HP steam generator 51-1-B-1:E-3 and the gasifier waste heat boiler are supplied by this 598°F boiler feedwater. Part of the outlet from economizer 1A is pumped to the fuel gas preheater 51-1-E-13 and returned to economizer 1B inlet at 347°F. The fuel gas preheater 51-1-E-13 heats fuel gas from 298°F to 580°F, before the gas is routed to the fuel gas expander, 51-1-EX-1. The heating of fuel gas is accomplished outside the HRSG for safety considerations. Expansion of the fuel gas to 245 psig produces electric power. The expanded gas at 339°F is routed to the gas turbine 51-1-GT-1.

Both the IP BFW and MP BFW needs are met by boiler feedwater pump 51-P-2. A portion of the pump discharge stream supplies IP steam generators in the process areas, while the balance is "let down" to supply MP process area steam generators and HRSG MP evaporator 51-1-B-1:E-7. BFW heating to MP saturation temperature (347°F) is accomplished in economizer 3, while heating to IP saturation (450°F) is done in economizer 2.

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

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

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

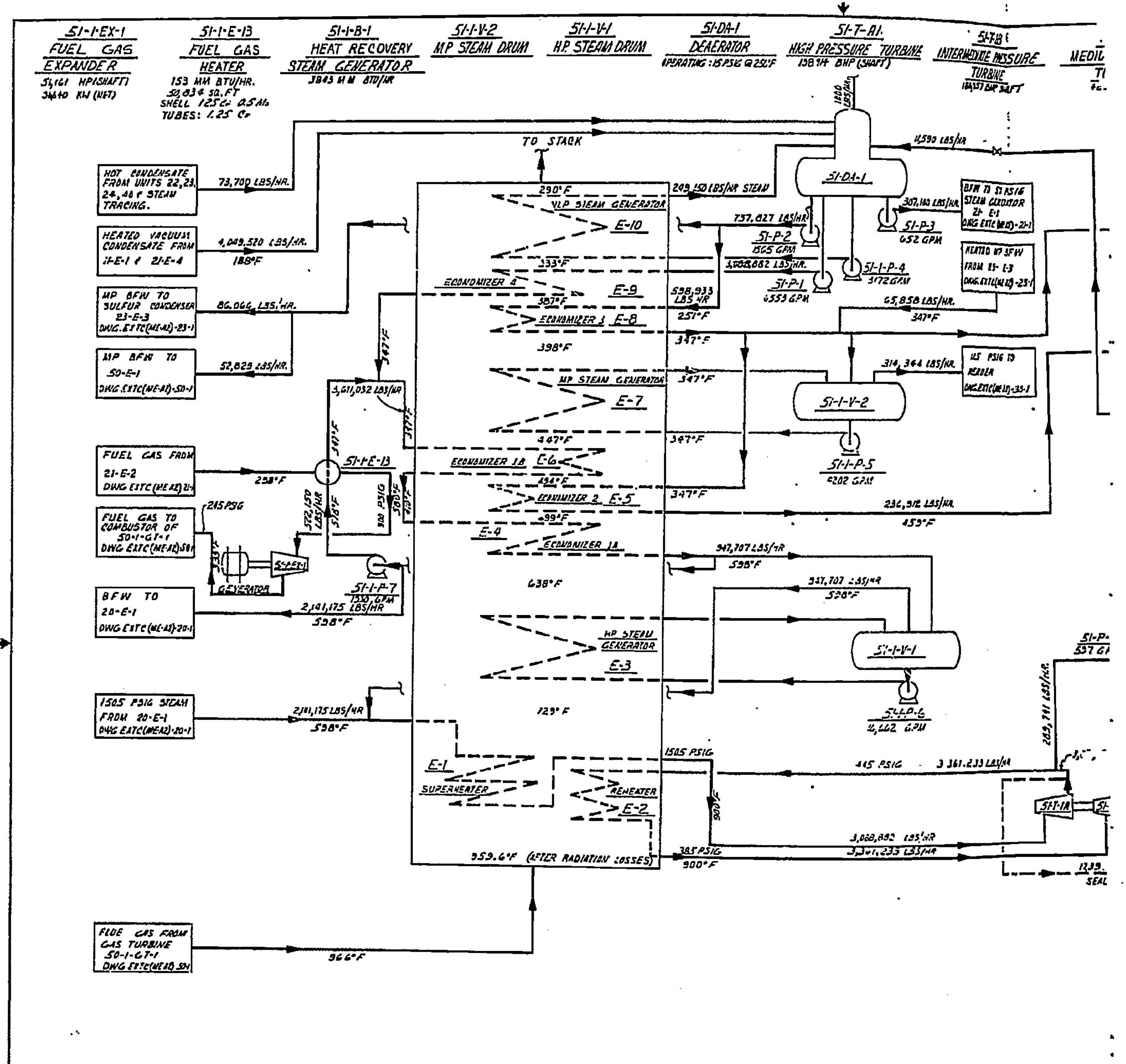
HP and MP evaporators are supplied with steam drums 51-1-V-1 and 51-1-V-2, respectively; and BFW circulation pumps 51-1-P-6 and 51-1-P-5. BFW is pumped through the evaporators, at feed to steam mass ratio of 6:1.

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

Equipment Notes

The HRSG heat transfer sections contain compact horizontal banks of finned tubes connected to headers located outside the gas path. Electrical heating elements are provided to maintain standby readiness during freezing conditions. The equipment supplied herein is standard today and is readily available.

Steam turbines and generators are commercially available. The expander is a prototype machine, because of the high inlet temperature and pressure. The machine will have a barrel-type casing and use isocarbon seal or equivalent requiring buffer gas. A reduction gear will be required.



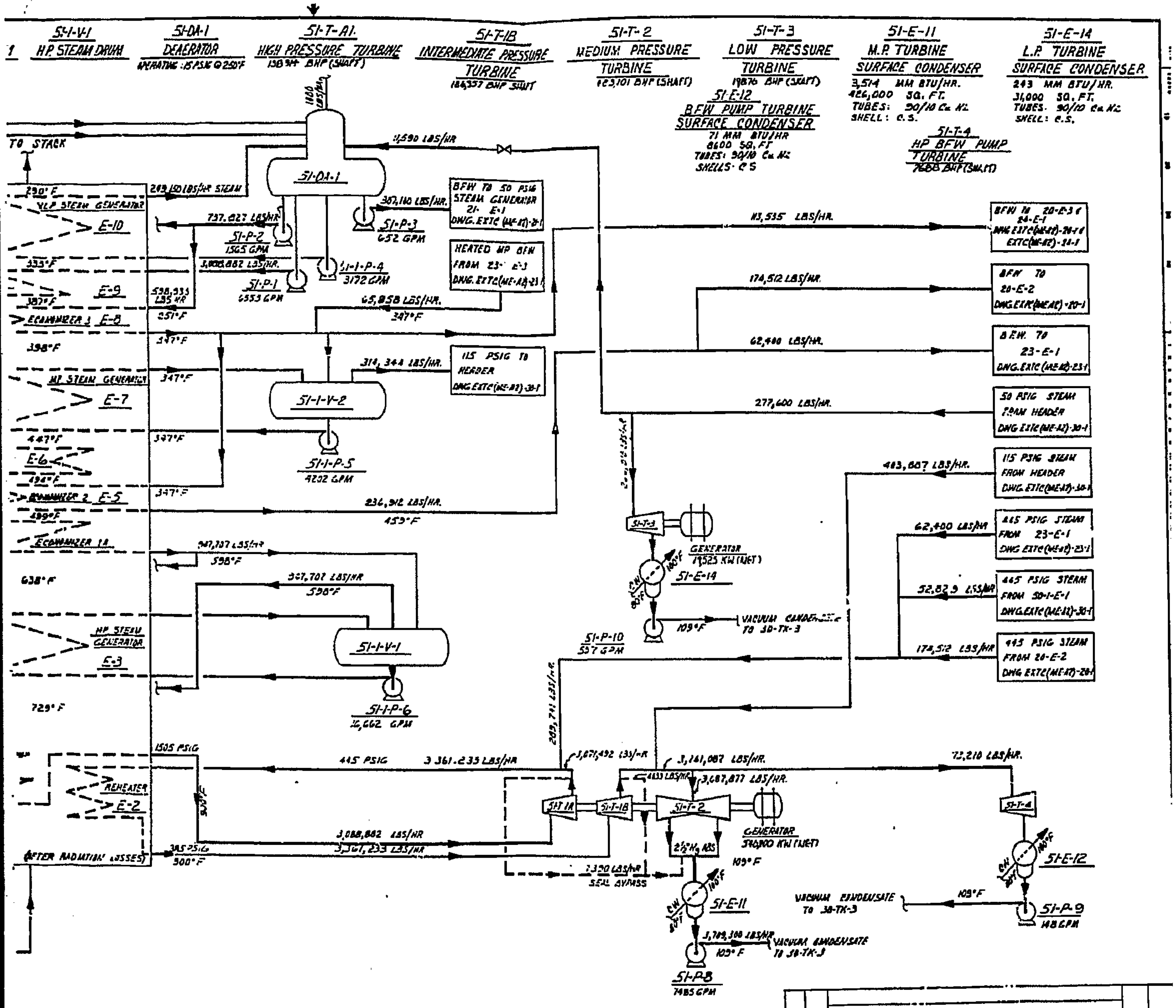
NOTES:

1- THIS FLOW DIAGRAM S
TOTAL OF FOURTEEN PARALLEL
EXCEPTION OF 51-T-1 &2,
ASSOCIATED EQUIPMENT
SINGLE TRAIN.

2- ALL FLOW RATES, DUTIES,
SPECIFICATIONS ARE
PLANT BASIS AT FULL
AS NOTED.

3- THE GENERATOR TERMINAL
IS BASED ON A 98% EFFICIE

4- A TEMPERATURE DROP OF
RADIATION HEAT LO
EXHAUST & HRSG.



NOTES:

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

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

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

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

DATE	REVISION DESCRIPTION	DRAWN	CHECK	APP

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

V. FRIAS
D'BELMITO
RAMCHAMP
SULISER
EPRI

PROCESS FLOW DIAGRAM
HEAT RECOVERY & STEAM GENERATION
CASE A2- BASE CASE (NO METHANOL, 500 PSIG SIM
PAID AND CRG
EPRI

SCALE NONE
448334-EXTC. (ME-A2) -51-1
03

The following descriptions of cooling water systems and support facilities are applicable to both cases.

COOLING WATER SYSTEMS

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

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

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

The cooling-water systems are sized specifically for each case.

GENERAL FACILITIES

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

- Plant and instrument air
- Potable and utility water
- Fuel system
- Nitrogen system
- Effluent water treating

- Flare system
- Fire water system
- Buildings
- Railroad loading
- Electrical system

Plant and Instrument Air

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

Potable and Utility Water

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

Fuel System

Fuel oil is used as startup fuel for the combustion turbines. The fuel system consists of three tanks, two motor-driven pumps, and suction heaters (for viscosity reduction). The fuel oil storage capacity is sufficient to support full operation of all gas turbines for one week: 67,000 bbl tanks for Case A2 and 45,000 bbl tanks for Case B2. The multiple tank system allows the use of more than one type of fuel oil and facilitates tank cleaning on a rotation basis.

Nitrogen System

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

Effluent Water Treating

The water streams treated are:

- Process condensate blowdown from gasification and ash handling
- Storm water
- Utility wastewater
- Cooling tower blowdown

A process condensate blowdown rate of 107 gpm from the gasification and ash handling section, together with a smaller tail gas treating unit blowdown, was selected as the sizing basis for the process condensate treating unit. Process condensate composition data are not adequate to finalize the flow which must be treated in a full-scale commercial GCC plant. The process itself is designed to remove formates, sulfides, ultrafine ash particles, and ammonia from the water using the following steps:

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

The effluent water from the process condensate treating unit is suitable for disposal in a navigable body of water. By-products are a precipitate cake and a biotreater sludge both of which may be combined with the ash cake for disposal. The ammonia stripped from the water is routed to the sulfur plant (Unit 23) furnace. The quantity of ammonia has been judged insufficient to consider its recovery for sale

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

The cooling water system includes one process and four utility mechanical draft cooling towers. Blowdown from the utility cooling towers is softened by cold lime-soda treating before being added to the process cooling water system.

Blowdown from the process cooling tower is combined with the treated process condensate prior to disposal.

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

Flare System

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

Fire Water System

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

Buildings

Following is a list of the buildings included in the capital estimate:

Building	Total Area Square Feet
Substations (5 provided)	7,460
Control Houses	12,250
Operators Shelters	2,000
Administration	30,000
Laboratory	7,500
Cafeteria	7,500

Change House and Guard House	4,800
Fire House	2,400
First Aid	3,200
Maintenance	45,000
Warehouse	16,500

Railroad Unloading and Loading

Provisions are made for unloading coal and for loading liquid sulfur, ash, and sludge, as well as the methanol produced in Case B2. Most of the coal unloading facilities are included in the coal preparation section. The necessary railroad equipment is contained in the general facilities. Our capital estimate for the unloading and loading scope encompasses 13,500 lineal feet of track, three switches, one bumper, and six road crossings. The design of the rail spurs and sidings are in accordance with the standards of a typical local railroad company. We have assumed that railroad company main tracks lie immediately adjacent to the plant site. Two heated 3750 bbl liquid sulfur storage tanks with loading pumps are provided for loading liquid sulfur. Ash from the process area and sludge from the effluent water treating unit are loaded into the railroad cars for transport to a disposal site.

Methanol storage tanks with loading pumps are provided for loading methanol for Case B2. The cost of this loading system is part of Unit 25, the methanol plant.

Electrical System

The following items of the electrical system are outside the scope of the capital estimates:

- The high-voltage switchyard including step-up transformers and switchgear for power export from the steam and gas turbine generators
- Connections from generator terminals to the switchyard
- The high-voltage electric power distribution and associated control and protective system from the switchyard to the high-voltage terminals of the captive transformers for large motor drivers in the oxidant feed unit

Included in the capital estimates are:

- The captive transformers

- The high-voltage electric power distribution system from the switchyard to the high-voltage terminals of the various electrical substations within the plant
- The electrical substations including stepdown transformers supplying power to various consumers at the required voltages