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Economics of the Texaco Gasification Process for Fuel-Gas Production

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Research Project 239

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NOTICE

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ABSTRACT

This study consists of an economic evaluation of oxygen-blown, Texaco-based gasification of 11,000 tons/day of Illinois No. 6 coal for the production of clean, intermediate-Btu fuel gas.

As the high temperature gas coolers required in this type of plant represent the highest risk developmental equipment components, two base-case gas cooling configurations were investigated. The first, involving <u>saturated</u> high-pressure steam generation, is representative of designs currently operating at the 150 ton/day Ruhrchemie plant and being designed for the 1000 ton/day Cool Water Demonstration plant. The second base case configuration involved the higher risk and undemonstrated design which incorporates steam <u>superheating</u> capability in the gas cooling section. Results of the evaluation indicated no economic incentives to develop superheating capability in the Texaco high temperature gas coolers.

The plant design employing only <u>saturated</u> high-pressure steam generating capability in the hot gas coolers produced 6664 x 10^6 Btu/hr of clean fuel gas (94.6% sulfur removal) and 142.4 MW of export power. Assuming mature technology, a plant startup date of 1990, a 10 percent annual inflation, a minimum after-tax return on equity of 20 percent/year for nonregulated company ownership, and a 1980 dollar by-product power credit of 50 mills/kWh, the estimated first year fuel gas costs would be:

Fuel Gas Plant Ownership	Current Dollars (1990)	Constant 1980 Dollars
Investor Owned Utility	$11.63/10^6 Btu	$4.27/10^6 Btu
Nonregulated Company	$14.48/10^6 Btu	$5.32/10^6 Btu

These results indicate that clean fuel gas produced from coal in mature Texaco gasification plants has the potential to be competitive with petroleum derived fuels.

Included in the evaluation were substudies which assessed the impacts of the use of a gas recycle, a change in the extent of sulfur removal, certain economies of scale, and changes in steam cycle conditions. The impacts of these changes on the estimated costs of fuel gas were minor.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This final report, Economics of the Texaco Gasification Process for Fuel Gas Production, presents a detailed engineering and economic evaluation of Texaco-based gasification of Illinois No. 6 coal for the production of clean, intermediate-Btu fuel gas. Previous evaluations, conducted by Fluor Engineers and Constructors, Inc., under RP 239-2, have indicated that Texaco-based coal gasification-combined-cycle power plants employing currently available (2000°F) gas turbines have the potential to be cost competitive with conventional coal-fired steam plants designed to comply with the 1979 Federal New Source Performance Standards.

Texaco's coal gasification technology is currently in the final stages of commercial development. The 150 ton/day Ruhrchemie plant in Oberhausen, West Germany has been operating for over three years. Construction of a 1000 ton/day Texaco gasification facility at Tennessee Eastman's Kingsport plant for the production of methanol and other chemicals is well underway. Finally, construction of the 1000 ton/day Texaco gasification-combined-cycle demonstration plant at Southern California Edison Company's Cool Water facility started at the end of 1981. EPRI is a major participant in this latter project scheduled for plant startup in mid-1984.

This study, therefore, was designed to investigate the potential economics of producing clean intermediate-Btu fuel gas from large, Texaco-based, coal gasification plants.

PROJECT OBJECTIVES

The specific objectives of this engineering study were to:

- Determine the cost of producing clean, intermediate-Btu fuel gas from Texaco-based gasification systems, using the most current cost and performance information available.

- Evaluate the potential economic incentives for the development of steam superheating capability in the high temperature gas coolers of Texaco-based fuel gas plants.
- Determine the impacts of various process design modifications on the efficiency and cost of fuel gas production.

PROJECT RESULTS

It is important to realize at the outset that this study of Texaco gasification differs fundamentally from all previous EPRI studies of Texaco-based systems. Cost estimates for the gasification and gas cooling sections of the plant have been updated to reflect current information from Ruhrchemie's 150 ton/day plant as well as certain design information from the Cool Water plant. Prior Fluor evaluations for EPRI of Texaco-based systems (see EPRI Report Numbers AF-642, AF-753, AF-916, AF-1288, AP-1543, AP-1624, AP-1725 and AP-2212) presented gasification section costs based only on data from the 15 ton/day Montebello pilot plant and did not have the benefit of the more recent experience. Therefore, cost estimates appearing in this report for the Texaco gasification section will differ from data published in previous EPRI reports.

To satisfy the first objective of the study, a fuel gas plant processing 11,000 tons/day of Illinois No. 6 coal was designed. This plant raised high-pressure saturated steam in the hot gas coolers (radiant and convective) configured in a manner similar to those in the Ruhrchemie plant and the Cool Water design. This plant which removes 94.6 percent of the sulfur in the coal, produces 6664 x 10^6 Btu/hr of clean fuel gas and 142.4 MW of export power at an overall thermal efficiency of 77.1 percent. Capital and operating costs for this plant starting up in 1990 are presented below:

	Investor Owned Utility Ownership		Nonregulated Company Ownership	
	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars
Total Capital Requirement, $\$10^6$	2,233	903	2,229	902
First Year (1990) Fuel Gas Cost, $\$/10^6$ Btu	11.63	4.27	14.48	5.32
Levelized Fuel Gas Cost, $\$/10^6$ Btu	19.11	3.27	24.51	5.32

(Financial criteria used to generate the above estimates can be found in the Summary, Table S-2.)

The above results indicate that fuel gas produced by a utility owned Texaco-based gasification plant has the potential to be lower in cost than equivalent crude oil derived products. For nonregulated company ownership the current economic potential is not as clear. The $\$5.32/10^6$ Btu production cost estimate is based on an after-tax annual return on equity of 20 percent (assuming 10 percent/year general inflation). This level of return is considered marginal for a high risk venture such as the production of a new product from an as yet unproven technology. If the after-tax return on equity requirement is increased to 30 percent, the nonregulated company fuel gas production cost would be $\$8.80/10^6$ Btu in constant 1980 dollars. Such a fuel gas price could only achieve parity with fuel oil in the mid-1990's if fuel oil were to experience an average annual real price growth (above general inflation) of 3 percent.

It is important to realize that the capital and operating cost estimates presented in this report are representative of those to be anticipated for mature technology plants. It is therefore to be expected that costs experienced for the first few Texaco-based fuel gas plants would be significantly higher than those presented here. It must also be realized that until the Texaco technology has been proven at full commercial scale in the Cool Water Demonstration Project, the process still poses significant risk.

In order to assess the potential economic incentives for developing superheating capability in the hot gas coolers, a second plant was designed with both superheating and reheating of steam in these coolers. This eliminated the requirement to burn product gas for superheating and reheating steam, thereby increasing the net fuel gas production by 15 percent. On the other hand, the export power generated decreased by 63 percent. For a utility owned plant, if the export power is credited at 41 mills/kWh or more, no incentive was found to develop superheating capability in the hot gas coolers. This finding is extremely useful as it can save substantial development costs for high risk equipment that offers no economic incentives for development.

Finally, many sensitivity studies to both design and financial parameters were conducted. Major conclusions from these sensitivity studies are:

- If the sulfur removal requirement is increased from 94.6 percent to essentially complete removal (i.e., 1 ppm sulfur in the product gas), the cost of fuel gas would increase by less than 4 percent.

- At a by-product electricity credit of 50 mills/kWh, the cost of fuel gas is insensitive to steam cycle conditions. However, at a by-product power credit of 100 mills/kWh, an efficient reheat steam cycle could reduce fuel gas costs by as much as 20 percent.
- If the capital investment required for the fuel gas plant were to increase by 35 percent, the levelized fuel gas production cost would increase by 13 percent.

In conclusion, it must be pointed out that neither the new ACRS tax rules nor recently promulgated energy tax credits were applied in any of the financial analyses presented in this report. Application of these favorable tax rules would tend to somewhat reduce the fuel gas production cost estimates presented.

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FLOW DIAGRAMS

SUMMARY

Previous EPRI studies have identified integrated coal gasification-combined-cycle power generation as an attractive option for new base load power plants. Coal gasification for power generation, however, is not limited to such new integrated facilities, but can be considered as a means of producing an ultraclean intermediate-Btu gas (IBG) from coal for replacing oil or natural gas in existing fired equipment. There is a need, therefore, to develop the economics of stand-alone clean fuel gas from coal plants.

EPRI has identified the Texaco coal gasification process as one of the more interesting second generation gasification schemes for future application to utility power production. Recently EPRI became a major participant in the Cool Water Demonstration Project through which the Texaco process will be demonstrated on a commercial scale in an integrated gasification-combined-cycle (GCC) power plant at Southern California Edison's Cool Water Station.

Prior Fluor evaluations performed for EPRI assessed the costs of Texaco-based gasification systems based on performance information from the 15 ton/day Montebello pilot plant. Consequently, an enormous scale-up was required in order to estimate the cost of mature technology, commercial-scale gasifiers. More recent data from Ruhrchemie's 150 ton/day Texaco gasification plant as well as detailed designs for the Cool Water plant allow better estimates of commercial-scale systems to be made. <u>Therefore, cost estimates appearing in this report for the Texaco gasification section will differ from data published in prior EPRI reports</u>.

The primary objectives of this study were to evaluate the following:

- Using the most current cost and performance information available, determine the cost of producing clean, intermediate-Btu fuel gas from Texaco-based gasification plants.

- Evaluate the potential economic incentives for the development of superheating capability in the high temperature gas coolers of Texaco-based fuel gas plants.

- Determine the impacts of various process design modifications on the efficiency and cost of fuel gas production.

DISCUSSION OF BASE CASE RESULTS

The oxygen-blown Texaco coal gasification process has significant steam raising capability in the raw gas cooling section. Depending on the design choices made in this section of the plant, large quantities of by-product power can be exported by a Texaco-based fuel gas plant processing 11,000 ton/day of Illinois No. 6 coal. The magnitude and economics of by-product power generation rely in part upon the design choices adopted for the superheating and reheating of high-pressure steam. Current practice in the 150 ton/day Ruhrchemie plant and the Cool Water design is to produce only saturated, high-pressure steam in the raw gas coolers because this mode of operation minimizes metal temperatures in these high temperature heat transfer devices. Such a design has very little impact on a GCC plant because saturated steam produced in the raw gas coolers can be superheated and reheated in the combined cycle heat recovery steam generator. In a fuel gas plant, however, this option does not exist. Therefore, superheating and reheating in this type of configuration would have to be effected by burning some fraction of the clean fuel gas product in an external fired heater. Such an internal consumption would reduce the net production of clean fuel gas.

However, superheating and reheating in the raw gas cooling section increases equipment costs and risks for this section of the system. Therefore, a need exists for the evaluation of the overall impact on fuel gas cost of superheating and reheating design options in Texaco-based plants. Two base case designs have been developed: one (Case EXT-SS) which generates high-pressure <u>saturated</u> steam in the raw gas coolers, and the other (Case EXT-SH) which generates and reheats high-pressure <u>superheated</u> steam in the raw gas cooling section. <u>This second case incorporates highly developmental heat transfer equipment for superheating and reheating steam</u> and represents a significant extension of the technology both currently being employed by Ruhrchemie and now being designed for Cool Water.

The Performance Summary, Table S-1, shows that both plants produce a gas with a higher heating value of 282.3 Btu/SCF (dry). However, the saturated steam case produces 6664 x 10^6 Btu/hr of clean fuel gas whereas the superheated steam case produces 7634 x 10^6 Btu/hr of clean fuel gas. The net by-product power generated in the saturated and superheated steam cases is 142.40 MW and 53.30 MW

Table S-1

PERFORMANCE SUMMARY - TEXACO GASIFICATION FUEL GAS PLANTS - BASE CASES(1)

Steam Cycle, psig/°F/°F	1450/900/900	1450/1000/1000
Steam Generated in Gas Coolers	Saturated	Superheated
Gas Temperature Entering Heat Recovery, °F	2400	2400
Sulfur Removal, %	94.6	94.6
Oxidant Plant Compressor Drivers	Motors	Motors
Nominal Capacity of Gasifier, tons/day	1375	1375
CASE DESIGNATION	EXT-SS	EXT-SH
GASIFICATION		
Coal Feed Rate, lb/hr (dry)	806,666	806,666
Oxygen*/Coal Ratio, lb/lb m.f.	0.8921	0.8921
Oxidant Temperature, °F	300	300
Slurry Feed Solids Content, weight %	66.5	66.5
Gasification Section Avg Pressure, psig	600	600
Raw Gasifier Effluent Temperature, °F	2,400	2,400
Raw Gasifier Effluent HHV (dry basis), Btu/SCF**	275.8	275.8
Cold Gas Efficiency (raw gas HHV/coal feed HHV x 100), %	74.64	74.64
POWER SYSTEM		
Temperature of Fuel Gas to Gas Expander, °F	600	600
Gas Expander Exit Temperature, °F	195	195
Condenser Pressure, Inches Hg abs	2.5	2.5
Fired Heater Stack Temperature, °F	400	--
Gas Expander Power#, MW	61.75	61.75
Steam Turbine Power#, MW	258.97	166.85
Oxygen Plant Power##, MW	1.81	1.81
Power Consumed, MW	180.14	177.11
Net System Power, MW	142.40	53.30
OVERALL SYSTEM		
General Facilities Water Consumption, GPM	150	150
Land, acres	190	175
Ash Disposal Rate, Dry ST/D	1023	1023
Sulfur By-Product, ST/D	354	354
Process and Deaerator Makeup Water, GPM	467	450
Cooling Tower Makeup Water, GPM	6,370	3,212
Cooling Water Circulation Rate§§, 10^3 GPM	227	167
Cooling Tower Heat Rejection§§, % of coal HHV	22.12	16.26
Air Cooler Heat Rejections, % of coal HHV	2.10	3.04
Clean Fuel Gas Efficiency (exported clean gas HHV x 100/coal feed HHV), %	64.67	74.11
Energy Recovery Efficiency (exported power + exported clean gas HHV)/coal feed HHV x 100, % ¶	77.11	78.77
Clean Fuel Gas HHV (dry basis), Btu/SCF	282.3	282.3
Net Clean Fuel Gas Product, 10^6 SCFD	566.63	649.25
10^3 SCF/ton DAF coal	65.03	74.52
10^6 Btu/hr	6664	7637

(1) This table is identical to Table 3-1, page 3-2
* Dry basis, 100 percent oxygen
** Excluding the HHV of H_2S, COS, and NH_3
\# At generator terminals
\#\# From power recovery expander 11-ME-1
§§ Includes process and power plant portions
¶ Export power credited at 9000 Btu/kWh

respectively. More high-pressure saturated steam is produced in the raw gas cooling section of Case EXT-SS than high-pressure superheated steam is generated in Case EXT-SH. As a consequence, the saturated steam case generates more electric power. However, due to the internal consumption of the fuel gas for superheating and reheating the steam in Case EXT-SS, less fuel is available for export. The energy recovery efficiencies are 77.11 percent and 78.77 percent for the saturated and superheated steam cases respectively. These results indicate that there is very little performance incentive for developing superheating capability in the raw gas coolers.

Based on the financial criteria in Table S-2, an economic analysis of the cost of fuel gas was performed. A summary of production costs and of selling price estimates for Texaco-based fuel gas is shown in Table S-3. This table presents fuel gas production cost and selling price estimates for both investor owned utility and nonregulated company ownership. It is important to realize that the fuel gas cost estimates presented in this report do not include consideration of the new ACRS tax rules. Nor do they include any additional tax credits beyond the 10 percent shown in Table S-2. Application of these latest tax incentives will tend to reduce the fuel gas costs shown in this report.

From Table S-3, it can be seen that the configuration generating saturated steam in the raw gas coolers (Case EXT-SS), if owned by an investor owned utility, could produce clean fuel gas at a first year cost of $\$4.27/10^6$ Btu in constant, mid-1980 dollars ($\$11.63/10^6$ Btu in 1990 dollars). The design producing superheated steam in the raw gas coolers (Case EXT-SH), also owned by a utility company, would produce clean fuel gas at approximately the same cost. The reason that fuel gas from the saturated steam design is cost competitive with that from the superheated steam case is that the saturated case generates substantially more steam. At a by-product electricity credit rate of 50 mills/kWh used in these economic evaluations, the results of Table S-3 indicate that there would be no advantage to employing the superheated steam option over using the reduced risk saturated steam design in the gas coolers. The electricity credit would have to drop below 41 mills/kWh before the superheating option would become marginally attractive.

This conclusion concerning the lack of any economic incentive to develop superheating capability in the raw gas coolers applies equally when the fuel gas plant is owned by a nonregulated company. For this case the fuel gas product could be

Table S-2

FINANCIAL CRITERIA FOR REVENUE REQUIREMENT CALCULATIONS (1)

Plant Location	•	Southern Illinois
Post-1980 General Inflation Rate	•	10%/year
Year of Plant Startup	•	1990
Design and Construction Period	•	4 years
Project Book Life	•	30 years for an Investor Owned Utility
	•	20 years for a Nonregulated Company
Project Tax Life	•	13 years for Synfuels Plants
Tax Depreciation Method	•	Sum-of-the-Year-Digits
Net Plant Salvage Value	•	10% of PFI
Delivered Coal Cost (Mid-1980$)	•	$1.30 /$10^6$ Btu
Real Coal Price Escalation (Above General Inflation)	•	1%/year
Property Tax Rate	•	2%/year of Escalated PFI
Insurance Rate	•	1%/year of Escalated PFI
Federal Income Tax Rate	•	46%
State Income Tax Rate	•	6%
Investment Tax Credit	•	10% of Escalated PFI. Normalized over period of commercial operation for utility ownership. Credited during construction period for nonregulated company ownership.
Project Financing		
Investor Owned Utility		
Common Equity	•	35% at 16%/year after tax return
Preferred Stock	•	15% at 12.75%/year dividend
Debt	•	50% at 12.25%/year interest
Nonregulated Company		
Common Equity	•	100% at 20%/year after tax return
Preferred Stock	•	Zero %
Debt	•	Zero %
Capacity Factor	•	90%
By-Product Electricity Credit	•	50 mills/kWh in mid-1980$. The cost of electricity is allowed to escalate at the general inflation rate.

(1) This table is identical to Table S-1, page S-2

Table S-3

FUEL GAS PRODUCTION COSTS AND SELLING PRICE ESTIMATES (1)

	Investor Owned Utility Texaco-Based Fuel Gas Plant				Nonregulated Company Texaco-Based Fuel Gas Plant			
Case Designation	EXT-SS		EXT-SH		EXT-SS		EXT-SH	
Steam Generated in Gas Coolers	Saturated		Superheated		Saturated		Superheated	
Net Fuel Gas Production, 10^6 Btu/hr*	6664		7637		6664		7637	
Net By-Product Power, MW*	142.40		53.30		142.40		53.30	
By-Products Credited	Electricity		Electricity		Electricity		Electricity	
	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars
TOTAL CAPITAL REQUIREMENT For Startup in 1990, $/FOEB**/Day	68,503	27,700	66,178	26,760	68,404	27,660	66,080	26,720
TOTAL CAPITAL REQUIREMENT ($1,000)	2,233,078	902,970	2,203,802	891,132	2,229,420	901,491	2,200,144	889,653
Cost of Fuel Gas#, $/$10^6$ Btu								
First Year (1990)	11.63	4.27	11.56	4.25	14.48	5.32	14.01	5.15
Tenth Year (1999)	20.67	3.22	21.47	3.35	34.14	5.32	33.03	5.15
Twentieth Year (2009)	43.50	2.61	47.03	2.83	88.54	5.32	85.68	5.15
Levelized##	19.11	3.27	19.81	3.39	24.51	5.32	23.74	5.15

(1) This table is similar to Table 5-2, page 5-4.

* Production at 100 percent of design capacity.

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

\# End-of-year cost.

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

sold at a first year price of $\$5.32/10^6$ Btu in constant, mid-1980 dollars ($\$14.48/10^6$ Btu in 1990 dollars).

The conclusion that there is no apparent incentive to develop superheating capability in the raw gas coolers is extremely important as this knowledge can save substantial development costs for high risk equipment that offers no economic incentives for further development.

It is most important to realize that the capital and operating cost estimates presented in this report are representative of those to be anticipated for mature technology plants, that is, the fourth or fifth commercial scale plant to be built. It is therefore to be expected that costs experienced for the first few Texaco-based fuel gas plants could be significantly higher than those presented here. It must also be realized that until the Texaco technology has been proven at full commercial scale in the Cool Water Demonstration Project, the process will still pose significant risk.

The estimated fuel gas costs presented require assumptions to be made concerning both design and financial factors. Therefore, the sensitivities of fuel gas costs to various changes in many of these parameters have been estimated and are presented in Table S-4. The major conclusion to be derived from the sensitivities presented in this table is that even if the capital cost estimates presented in this report increase by 35 percent, the first year cost of fuel gas produced by a utility owned plant would remain attractive at $\$5.12/10^6$ Btu in constant, mid-1980 dollars ($\$13.95/10^6$ Btu in 1990 dollars).

COMPARISONS WITH PREVIOUS EPRI STUDIES

In 1975 and 1976, Fluor conducted evaluations of fuel gas production from a number of current and advanced coal gasification systems (reported in EPRI report numbers AF-244 and AF-782). These earlier results generally are not comparable with the results presented in this report for the following reasons:

- Most of the technologies evaluated were at an early stage of development. Subsequent evaluations have generally shown that less optimistic performance and capital cost estimates accompany greater definition of such technologies.
- Different financial criteria were employed in the earlier studies.

Table S-4

LEVELIZED AND FIRST YEAR (1990) FUEL GAS COST SENSITIVITY TO DESIGN AND FINANCIAL FACTORS
CONSTANT MID-1980 DOLLARS
INVESTOR OWNED UTILITY OWNERSHIP

	CASE EXT-SS				CASE EXT-SH			
	Levelized Cost of Fuel Gas		First Year (1990) Cost of Fuel Gas		Levelized Cost of Fuel Gas		First Year (1990) Cost of Fuel Gas	
	Levelized $/10^6 Btu	Percentage Change From Base	First Year $/10^6 Btu	Percentage Change From Base	Levelized $/10^6 Btu	Percentage Change From Base	First Year $/10^6 Btu	Percentage Change From Base
Base Case Results	3.27	Base	4.27	Base	3.39	Base	4.25	Base
70% Capacity Factor Operation	3.80	+16.2	5.16	+20.8	3.84	+13.3	5.02	+18.1
3% Escalation in Real Coal Price	4.55	+39.1	4.73	+10.8	4.50	+32.7	4.65	+9.4
35% Increase in Plant Facilities Investment	3.70	+13.1	5.12	+19.9	3.76	+10.9	4.98	+17.2
5% Annual Inflation Rate*	3.28	+0.3	3.20	-25.1	3.41	+0.6	3.34	-21.4
Clean Fuel Gas Efficiency Decrease by 10%	3.74	+14.4	4.86	+13.8	3.80	+12.1	4.76	+12.0
Clean Fuel Gas Efficiency Increase by 10%	2.88	-11.9	3.79	-11.2	3.05	-10.0	3.83	-9.9

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

Annual return on common equity	10.73%
Annual preferred stock dividends	7.63%
Annual interest on debt	7.15%
Annual coal price escalation	6.05%

The only commercial technology evaluated in these earlier studies was the dry ash Lurgi process. Therefore, to provide a bridge between the earlier studies and this work, EPRI updated the cost estimates for an oxygen-blown Lurgi fuel gas plant to 1980 dollars and evaluated the fuel gas production costs on the same financial basis as was used in this report. The results are shown in Table S-5.

Extreme care must be taken in making any comparisons between the estimates presented in Table S-5 for the following reasons:

- Preliminary data on the performance of caking bituminous coal (Illinois No. 6) in the Lurgi gasifier were used as the basis for the early Lurgi study.
- The Lurgi plant was designed in 1975 and therefore does not reflect the benefit of extensive data developed from the SASOL II experience.
- The design basis for the Lurgi plant was somewhat different than that specified for these later Texaco studies, and this has resulted in an inconsistent basis for comparison.

DISCUSSION OF DESIGN SENSITIVITY STUDIES

The gas cooling section of a Texaco fuel gas plant is developmental and costly. Decreases in the capital requirement and the cost of fuel gas could possibly be realized through design changes in the gas cooling configuration. Table S-6 summarizes the economic results from the base cases and substudy cases for investor owned utility production of fuel gas. The design changes considered include:

- The use of a cold gas recycle from the particulate scrubber to quench the hot gasifier effluent instead of the use of a radiant gas cooler.
- Variations in the degree of sulfur removal.
- The use of larger (2200 ton/day instead of 1375 ton/day) capacity gasifiers.
- Variations in steam cycle temperature and pressure.

The major conclusions to be derived from these system design substudies are summarized below:

- The use of a cold gas recycle producing a 1500°F feed gas to the convective heat exchangers would increase the cost of product fuel gas only marginally. This important result must be treated with caution. It is most encouraging to the extent that it indicates a possible alternative to the current design concept of radiant/convection gas cooling that will not significantly impact the cost of gas. However, it must be

Table S-5

FUEL GAS COST ESTIMATES FROM
OXYGEN-BLOWN LURGI-BASED PLANTS
EMPLOYING ILLINOIS NO. 6 COAL
INVESTOR OWNED UTILITY OWNERSHIP

	Texaco-Based Plant (This Study)		Oxygen-Blown Lurgi Fuel Gas Plant			
Case Designation	EXT-SS		MX		MX	
Steam Generated in Gas Coolers	Saturated		N/A		N/A	
Net Fuel Gas Production, 10^6 Btu/hr*	6664		5495		5495	
Net By-Product Power, MW*	142.40		63.70		63.70	
By-Products Credited**	Electricity		Electricity Ammonia Hydrocarbons		Electricity Ammonia	
	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars	Current Dollars	Mid-1980 Dollars
Total Capital Requirement For Startup in 1990, \$/FOEB#/Day	68,503	27,700	76,639	30,990	88,164	35,650
Total Capital Requirement (\$1,000)	2,233,078	902,970	2,194,887	887,527	2,194,887	887,527
Cost of Fuel Gas##, $\$/10^6$ Btu						
First Year (1990)	11.63	4.27	14.36	5.28	15.66	5.76
Tenth Year (1999)	20.67	3.22	25.80	4.02	28.86	4.50
Twentieth Year (2009)	43.50	2.61	54.88	3.30	62.81	3.78
Levelized§	19.11	3.27	23.85	4.08	26.64	4.56

* Production at 100 percent of design capacity.
** By-product ammonia credited at \$120.00/short ton. Liquid hydrocarbons, when credited, were valued at \$3.00/10^6 Btu (1980 dollars).
# Barrels of distillate fuel oil (5.85×10^6 Btu/BBL) with higher heating value equivalent to fuel gas produced. Electricity credited at 9000 Btu/kWh. Hydrocarbons credited where noted.
§ A levelized price is one which, if held constant, will yield the same return on common equity as the varying year-by-year values.
End-of-year cost.

remembered that this recycle gas quench concept has not been demonstrated at any scale. If the assumption that 1500°F gas is acceptable in the convection gas coolers is not found to be technically feasible, this conclusion might not be valid.

- Sulfur Removal

  When the sulfur removal specification is increased from 83.6 percent to 94.6 percent, the cost of fuel gas increases insignificantly (1.2 percent for the high-pressure, superheated steam case).

  When sulfur removal is increased from 94.6 percent to essentially complete sulfur removal (corresponding to one ppm total sulfur in the product gas on a mole basis), the cost of fuel gas produced by an investor owned utility increases less than 4 percent.

- Since only relatively moderate costs are associated with deep sulfur removal, Texaco-based fuel gas plants appear to be capable of producing ultraclean fuel economically even in nonattainment regions.

- The development and use of a larger capacity (2200 ton/day) Texaco gasifier has the potential of reducing the cost of fuel gas by 5 percent. Some of the savings associated with this economy of scale in Cases EXT-SS and EXT-SH will be offset by the impact on overall plant availability of the slightly reduced fraction of spare operating trains in these designs.

- The fuel gas cost is relatively insensitive to steam cycle conditions at a 1980 dollar electricity credit of 50 mills/kWh. However, at higher by-product electricity credits the more efficient steam cycles result in significant reductions in fuel gas costs (i.e., at an electricity credit of 100 mills/kWh, changing from a 900°F non-reheat cycle to a 1000°F/1000°F reheat steam cycle will reduce the cost of fuel gas by up to 20 percent).

Table S-6

PRODUCTION COST ESTIMATES (†)
FOR TEXACO-BASED FUEL GAS
(INVESTOR OWNED UTILITY, MID-19?0 DOLLARS)

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

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

(†) This table is identical to Table 6-1, page 6-3.
*Case EXT-SS or EXT-SH with 2200 ton/day gasifiers.
¶One ppm total sulfur in the product gas on a mole basis.
#Production at design capacity.
##This value was derived by dividing Total Capital Requirement (in $1000) by the Fuel Oil Equivalent Barrel output per day. Using a conversion factor of 5.85×10^6 Btu/FOEB for fuel gas. Similarly, electricity production was converted to a FOEB/day equivalent assuming an energy value of 9,000 Btu/kWh.
δElectricity is credited at 50 mills/kWh in Mid-1980 Dollars and is allowed to escalate at the general inflation rate.

Section 1

INTRODUCTION

PURPOSE OF THE STUDY

The oxygen-blown Texaco coal gasification process shows promise for utilization in gasification-based clean fuel gas plants. This coal gasification technology is derived from Texaco Development Corporation's established commercial process for partial oxidation of heavy petroleum fractions.

Previous Fluor evaluations of the oxygen-blown Texaco coal gasification process for EPRI have focused on the use of this technology in a gasification/combined-cycle (GCC) power plant. EPRI's interest in this process includes continued funding of pilot plant studies at Texaco's Montebello facility and work with coals in Germany. More recently, EPRI has executed a contract to become a major participant in the Cool Water Project along with Texaco and Southern California Edison (SCE) and others. The goal of this project is the design, construction, and successful operation of an integrated, Texaco-based, 100 MW GCC demonstration plant at SCE's Cool Water Station near Barstow, California. This demonstration plant will be the first operating GCC system in the U.S. to use a commercial-scale Texaco gasifier.

Previous Fluor evaluations (for EPRI) of coal gasification for clean fuel gas production concentrated on gasification systems that were in a very early stage of development with the exception of the dry ash Lurgi technology (see EPRI reports AF-244 and AF-782). Recently, much interest has been shown in the use of Texaco's coal gasification technology for clean fuel gas production. Therefore, the overall objective of this study was to assess the cost of fuel gas produced in a Texaco-based gasification plant.

It is important to realize that Fluor has been assessing the costs of Texaco-based gasification systems under contract to EPRI for the past three years. All cost estimates generated in past studies (see EPRI reports AF-642, AF-753, AF-916, AF-1288, AP-1543, AP-1624, AP-1725 and AP-2212) were based on limited data for the Texaco portion of the system as well as a massive scale up from the 15 ton/day Montebello pilot plant to an estimated mature commercial capacity of 2200 ton/day gasifier. Within the last year, information released from Ruhrchemie's 150 ton/day Texaco gasification plant in Oberhausen, West Germany, have indicated that the initial cost projections for the Texaco gasification technology were low. Therefore, another major objective of this project was to update cost estimates for the Texaco coal gasification process to reflect recent data from both the Oberhausen plant as well as the Cool Water design effort.

The oxygen-blown Texaco coal gasification process has significant steam raising capability in the raw gas cooling section. This allows for the development of Texaco-based clean fuel gas plants which, depending on the design choices made, can export large quantities of by-product electric power. The relative quantity and economics of by-product power generation will depend in part upon the design choices adopted for the superheating and reheating of the high-pressure steam. Superheating/reheating in a gas fired heater reduces net production of the clean fuel gas since a fraction of the gross production is used internally. On the other hand, superheating/reheating in the raw gas cooling section increases equipment costs and risks for this section of the plant. Therefore, this engineering and economic evaluation of Texaco-based clean fuel gas plant designs contains substudies of options for steam superheating/reheating for by-product power generation.

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

In a Texaco-based GCC system, the cost of the acid gas removal, sulfur recovery and tail gas treating units is a relatively small fraction (approximately four percent) of the total plant investment. These same units constitute a much larger fraction (six to nine percent) of the total plant investment for a clean fuel gas plant. The impact of increased sulfur removal standards on fuel gas plant economics was therefore also studied.

In summary, the two principal objectives of this study were to:

- determine the cost of producing intermediate Btu gas in a Texaco-based gasification plant using high temperature gas cooling equipment producing saturated steam.
- evaluate the potential economic incentives for developing superheating capability in high temperature gas coolers.

Secondary objectives were to assess the impacts of certain process design modifications on the overall system efficiency and on the cost of fuel gas.

DESCRIPTION OF THE BASE CASES

The two base cases designed to achieve the principal objectives of this study were Base Case EXT-SS for high-pressure saturated steam generation, and Base Case EXT-SH for high-pressure superheated steam generation.

These cases employ Chicago summer design conditions and consist of an oxygen-blown Texaco gasification system. The gasification system operates at 600 psig using 98 mole percent oxygen as the oxidant. This oxidant is produced in an air separation plant and supplied to the gasifier at 720 psig and 300°F as in the AP-1624 report.

BASE CASE EXT-SS (High-Pressure Saturated Steam Generation). Gas exiting from the gasifiers (at 2400°F) is used to produce saturated, high-pressure steam at 1505 psig in the first energy recovery unit of the raw gas cooling system. All subsequent superheating and reheating of this steam is done in a heater fired with part of the clean fuel gas produced. The heat transfer service for each of the remaining exchangers in the raw gas cooling system has been chosen to maximize the overall thermal efficiency and minimize internal consumption of fuel gas.

The steam cycle consists of 1450 psig, 900°F superheated steam and a 900°F reheat temperature. Fuel gas expanders recover power by expanding the clean fuel gas to 50 psia.

BASE CASE EXT-SH (High-Pressure Superheated Steam Generation). This base case differs from Base Case EXT-SS primarily in the design of the raw gas cooling system and the steam cycle. The energy recovery unit in the raw gas cooling system produces superheated, high-pressure steam and also performs reheating of steam exhausted from the HP section of the power turbine. The heat transfer service for each of the remaining exchangers in the raw gas cooling system has been chosen to maximize the overall thermal efficiency of the entire plant. There is no internal consumption of the product fuel gas.

The steam cycle consists of 1450 psig, 1000°F superheated steam and a 1000°F reheat temperature.

Substudy Cases

Other process designs were evaluated and will be discussed in the section titled "Design Sensitivity Studies."

Included in the substudies is a scale-up of both base case designs with the use of 2200 ton/day gasifiers in place of the 1375 ton/day gasifiers employed in the base cases and all other substudies. The smaller gasifier is representitive of the scale planned for demonstration plants like Cool Water and the larger gasifier represents an assessment of the capacity of future mature systems when the technology has been well established.

TECHNICAL CRITERIA

Plant designs are based on technical criteria established by the Electric Power Research Institute (EPRI). These criteria include water and coal analyses, site location, and general plant requirements.

Fluor developed the plant designs for all cases. Some of the information needed to design the gasification systems was provided to Fluor from previous studies performed by the Texaco Development Corporation.

The analysis of the Illinois No. 6 coal which was used in all the study cases is 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 is not valid, then each of the cases presented here would require additional coal handling equipment. This change would affect the overall plant investment estimates, but would not alter the comparisons between cases.

The site for each of the plants is the Chicago area; Table 1-2 shows pertinent conditions for the site. Raw water makeup to the plant is assumed to be Chicago city water. The Chicago Department of Public Works provided an analysis of finished water from the South District filtration plant, Table 1-3. These data have been used in recent EPRI coal gasification reports and were extracted from EPRI Report AF-244.

The only net plant products are clean fuel gas, electricity, and recovered sulfur. Total gaseous sulfur emissions (including that present in product fuel gas) from the base case plant designs are restricted to 0.32 lb SO_2 equivalent per 10^6 Btu (HHV) of coal fed to the gasifiers.

Electric power is assumed to be available to each plant, for startup and emergency situations. Each plant is a grass roots installation, and fuel oil storage is provided for fired heater startup in saturated steam cases where a fired heater is used. This allows steam production to gradually bring the plant on-line. In addition to the major onsite units, the plant includes the following facilities in the cost estimate for each case:

- Cooling water systems (process plant and power block)
- Plant and instrument air
- Potable and utility water
- Fuel system

The process equipment used in each plant design consists primarily of commercially available units. Advanced designs were incorporated for the following items of equipment:

- The gasifier high-temperature heat recovery equipment designs used in the raw gas cooling systems are all extensions of the current state of the art for such equipment. Considerable development and testing work will be required before these designs reach commercial status.
- The 2200 ton/day gasifiers represent a potential scale-up of the commercial size, 1000 ton/day, gasifier which will be demonstrated by the Cool Water Project.

The present estimates represent those for mature technologies.

Table 1-1

COAL ANALYSIS

Type	Illinois No. 6
PROXIMATE ANALYSIS (Wt %)	
Moisture	12.0
Ash	8.8
Fixed Carbon	47.8
Volatile Matter	31.4
	100.0
ULTIMATE ANALYSIS - DAF COAL (Wt %)	
Carbon	77.26
Hydrogen	5.92
Oxygen	11.14
Nitrogen	1.39
Sulfur	4.29
Other	-
	100.00
HEATING VALUE - AS RECEIVED	
Higher Heating Value (HHV) (Btu/lb)	11,241
Net Heating Value (LHV) (Btu/lb)	10,758

Table 1-2

DESIGN SITE CONDITIONS

LOCATION	Chicago, Illinois
ELEVATION	600 feet
	Summer Design Cases
AMBIENT PRESSURE, psia	14.4
AMBIENT TEMPERATURES, °F	
Dry Bulb	88
Wet Bulb	75
WINTER DRY BULB, °F	0

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)	--
Hardness as $CaCO_3$ equivalents	
Total	168
Noncarbonate	30

Color	1 unit
pH	7.9
Turbidity	0
Specific Conductance @ 25°C	275 microhoms

Section 2

PLANT DESCRIPTIONS OF BASE CASES

GENERAL - BASE CASES

Two grass roots plants for fuel gas production based on oxygen-blown Texaco gasifiers with nominal capacities of 1375 short tons per day coal each are shown schematically on block flow diagrams EXT-SS-1-1 and EXT-SH-1-1. These plants consume 11,000 short tons per day of Illinois No. 6 coal, fed to the gasifiers in a water slurry containing 66.5 weight percent solids. The differences between the two cases are primarily in sections of Unit 20, High-Temperature Gas Cooling and Scrubbing; Unit 21, Low-Temperature Gas Cooling; and Unit 51, Power Generation, which includes a steam turbogenerator, and also a fired heater in Case EXT-SS.

The first block flow diagram (EXT-SS-1-1) represents the plant flow scheme in which the hot crude gas containing molten slag from the gasifier is used in a radiant/convection configuration, without recycle gas cooling, as a source of high-level heat for the generation of saturated steam (SS) at 1505 psig. All subsequent superheating and reheating of the steam take place in the fired heater.

Block flow diagram EXT-SH-1-1 represents the plant flow scheme in which hot crude gas is used, without recycle gas cooling, as a source of heat for the generation of high-pressure superheated steam (SH) at 1450 psig, 1000°F, as well as for the subsequent reheating of steam to 1000°F. This scheme does not require a fired heater.

In each case, the main plant consists of coal handling, grinding and slurry charging, oxidant feed, gasification, gas cooling, acid gas removal, and power generation systems. Coal receiving, storage, and conveying are done in a single train to minimize space and operating labor requirements. Coal grinding requires five parallel operating trains and one spare train. The oxidant feed unit has five parallel operating trains. There are eight parallel operating and two spare gasification/high temperature heat recovery trains. A two-train ash handling carbon recovery system (without spare) serves all of the gasification units. The gas cooling and acid gas removal units each consist of two operating parallel trains, while the power generation system has two parallel gas expanders, a single high-pressure steam turbogenerator, and a single fired heater in the saturated steam case only.

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

Table 2-1 shows the number of operating and spare trains for major sections of each plant.

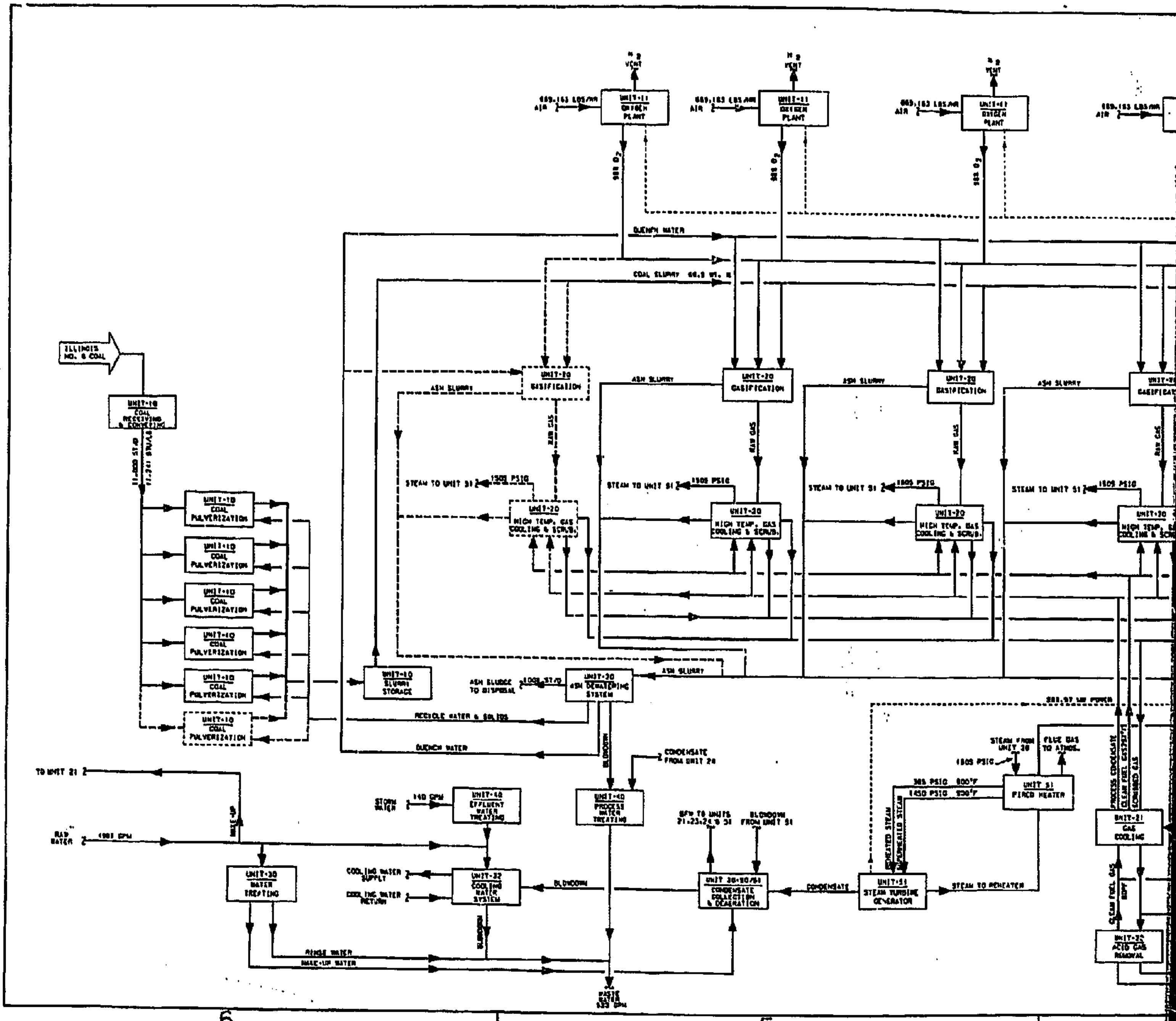
Table 2-1

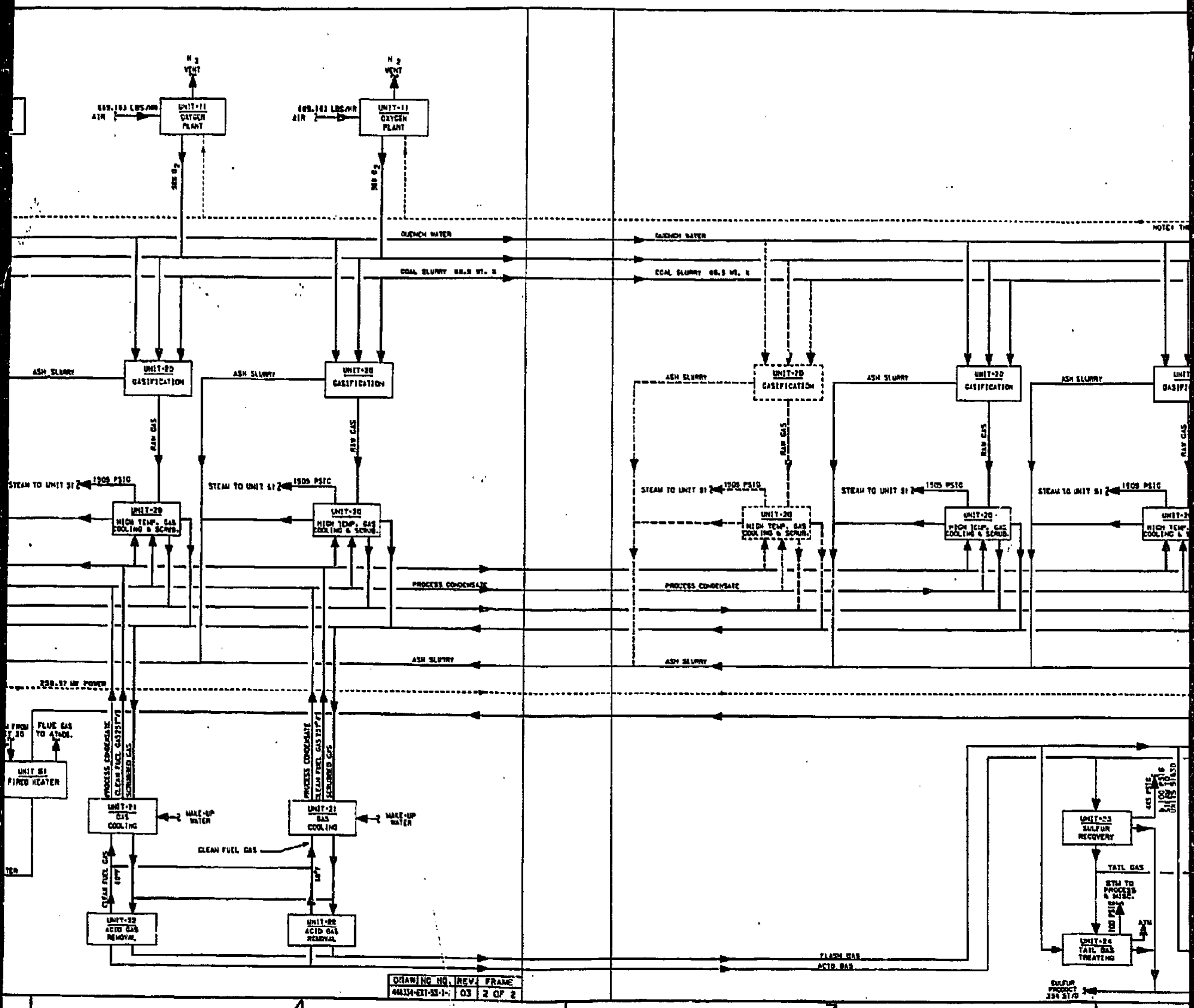
TRAINS OF EQUIPMENT IN MAJOR PLANT SECTIONS
BASE CASES EXT-SS AND EXT-SH

Unit No.	Name	Operating	Spare
10	Coal Handling	1	0
10	Coal Grinding	5	1
10	Slurry Charging	8	2
11	Oxidant Feed	5	0
20	Gasification	8	2
20	High-Temperature Gas Cooling and Gas Scrubbing	8	2
20	Ash Handling and Carbon Recovery	2	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, Boiler Feedwater, and Condensate System		
	• Condensate Collection and Deaeration	1	0
	• Water Treating	1	0
32	Cooling Water System	1*	0
40	Process Condensate Treating	1	0
40	Effluent Water Treating	1	0
50	Gas Expander/Generator	2	0
51	Fired Superheater/Reheater	**	0
51	Steam Turbine/Generator	1	0

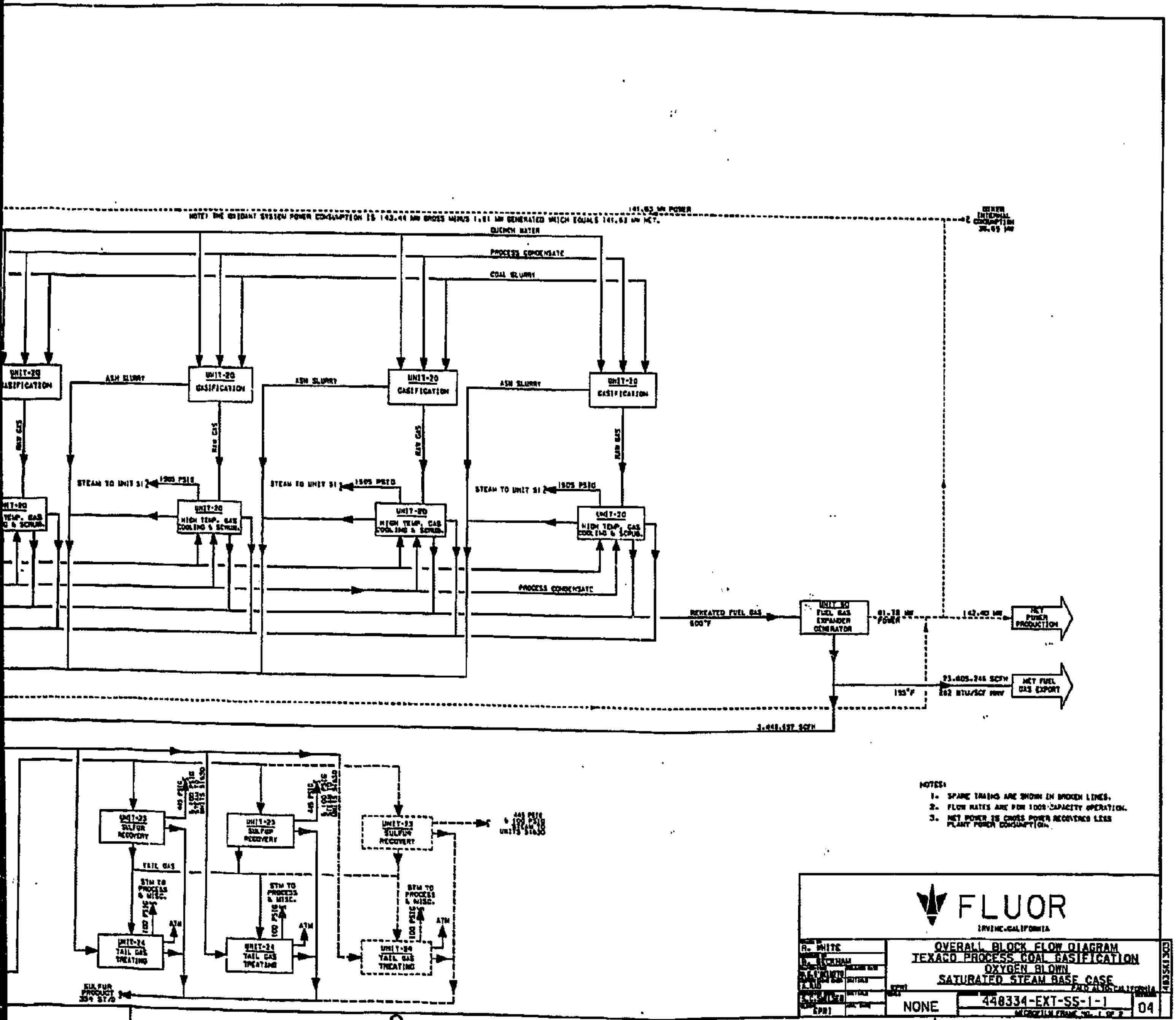
* The cooling tower dedicated to the process plant sections is separate from the towers dedicated to the steam turbogenerator condenser

** One in Case EXT-SS and none in Case EXT-SH

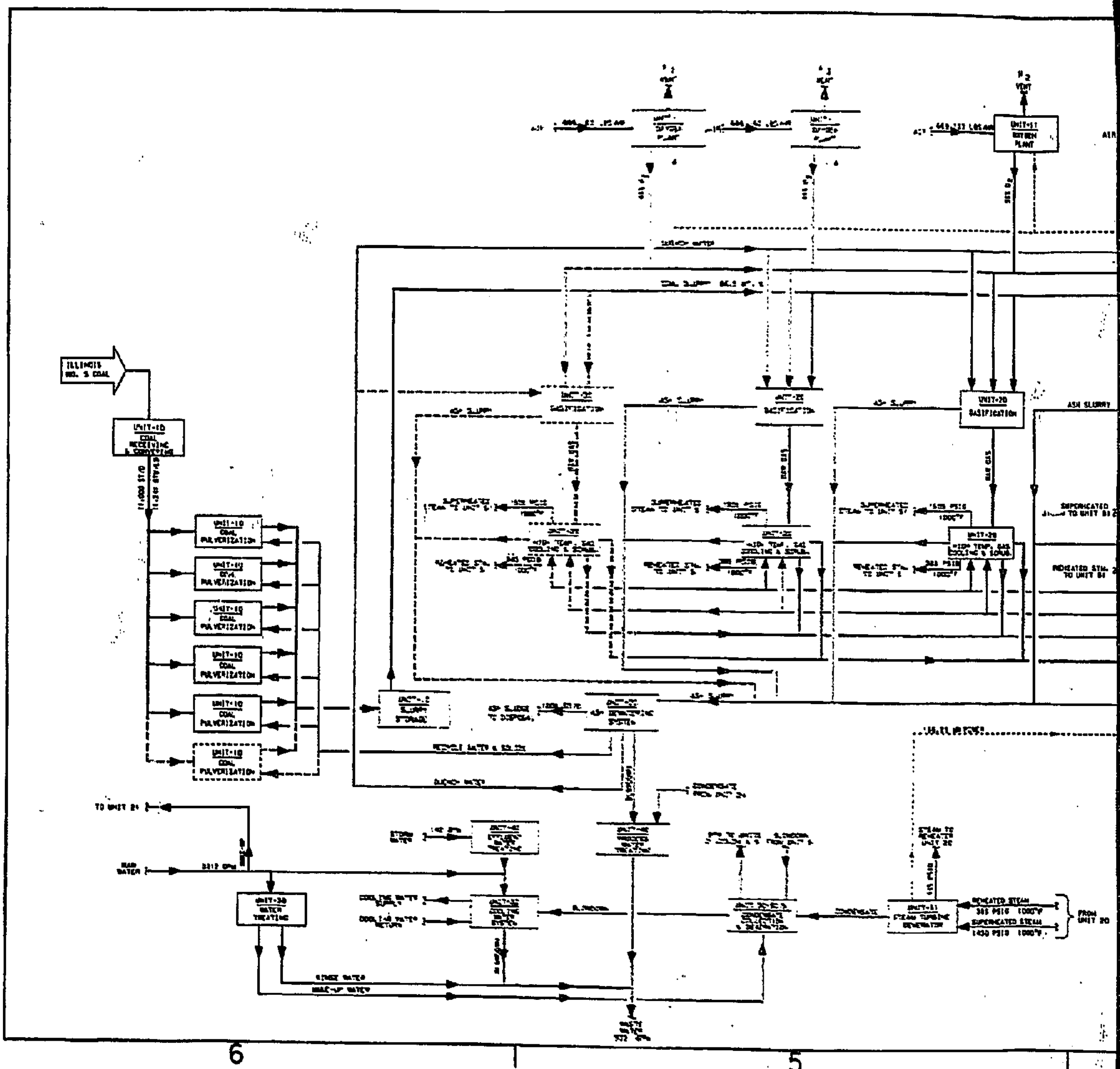

ILLINOIS NO. 6 COAL
UNIT-10 COAL RECEIVING & CONVEYING
UNIT-10 COAL PULVERIZATION
UNIT-10 SLURRY STORAGE
UNIT-11 OXYGEN PLANT
AIR
VENT
QUENCH WATER
COAL SLURRY
UNIT-20 GASIFICATION
ASH SLURRY
RAW GAS
STEAM TO UNIT 51
1505 PSIG
UNIT-20 HIGH TEMP. GAS COOLING & SCRUB.
UNIT-20 ASH DEWATERING SYSTEM
ASH SLUDGE TO DISPOSAL
RECYCLE WATER & SOLIDS
QUENCH WATER
BLOWDOWN
CONDENSATE FROM UNIT 24
TO UNIT 21
MAKE-UP
RAW WATER
STORM WATER
140 GPM
UNIT-40 EFFLUENT WATER TREATING
UNIT-40 PROCESS WATER TREATING
UNIT-30 WATER TREATING
COOLING WATER SUPPLY
COOLING WATER RETURN
UNIT-22 COOLING WATER SYSTEM
RINSE WATER
MAKE-UP WATER
WASTE WATER
BFW TO UNITS 21,23,24 & 51
BLOWDOWN FROM UNIT 51
CONDENSATE
UNIT 20/50/51 CONDENSATE COLLECTION & DEAERATION
UNIT-51 STEAM TURBINE GENERATOR
STEAM TO REHEATER
STEAM FROM UNIT 20
FLUE GAS TO ATMOS.
1505 PSIG
UNIT 51 FIRED HEATER
REHEATED STEAM
SUPERHEATED STEAM
PROCESS CONDENSATE
CLEAN FUEL GAS
SCRUBBED GAS
UNIT-21 GAS COOLING
CLEAN FUEL GAS
UNIT-23 ACID GAS REMOVAL
6
5



VENT
AIR
UNIT-11
OXYGEN PLANT
QUENCH WATER
COAL SLURRY
ASH SLURRY
UNIT-20
GASIFICATION
RAW GAS
STEAM TO UNIT 51
1505 PSIG
UNIT-20
HIGH TEMP. GAS COOLING & SCRUB.
PROCESS CONDENSATE
FLUE GAS TO ATMOS.
UNIT 51
FIRED HEATER
PROCESS CONDENSATE
CLEAN FUEL GAS
SCRUBBED GAS
UNIT-21
GAS COOLING
MAKE-UP WATER
CLEAN FUEL GAS
UNIT-22
ACID GAS REMOVAL
FLASH GAS
ACID GAS
UNIT-23
SULFUR RECOVERY
TAIL GAS
STM TO PROCESS & MISC.
UNIT-24
TAIL GAS TREATING
SULFUR PRODUCT
NOTE
DRAWING NO.
REV.
FRAME
03
2 OF 2
4
3

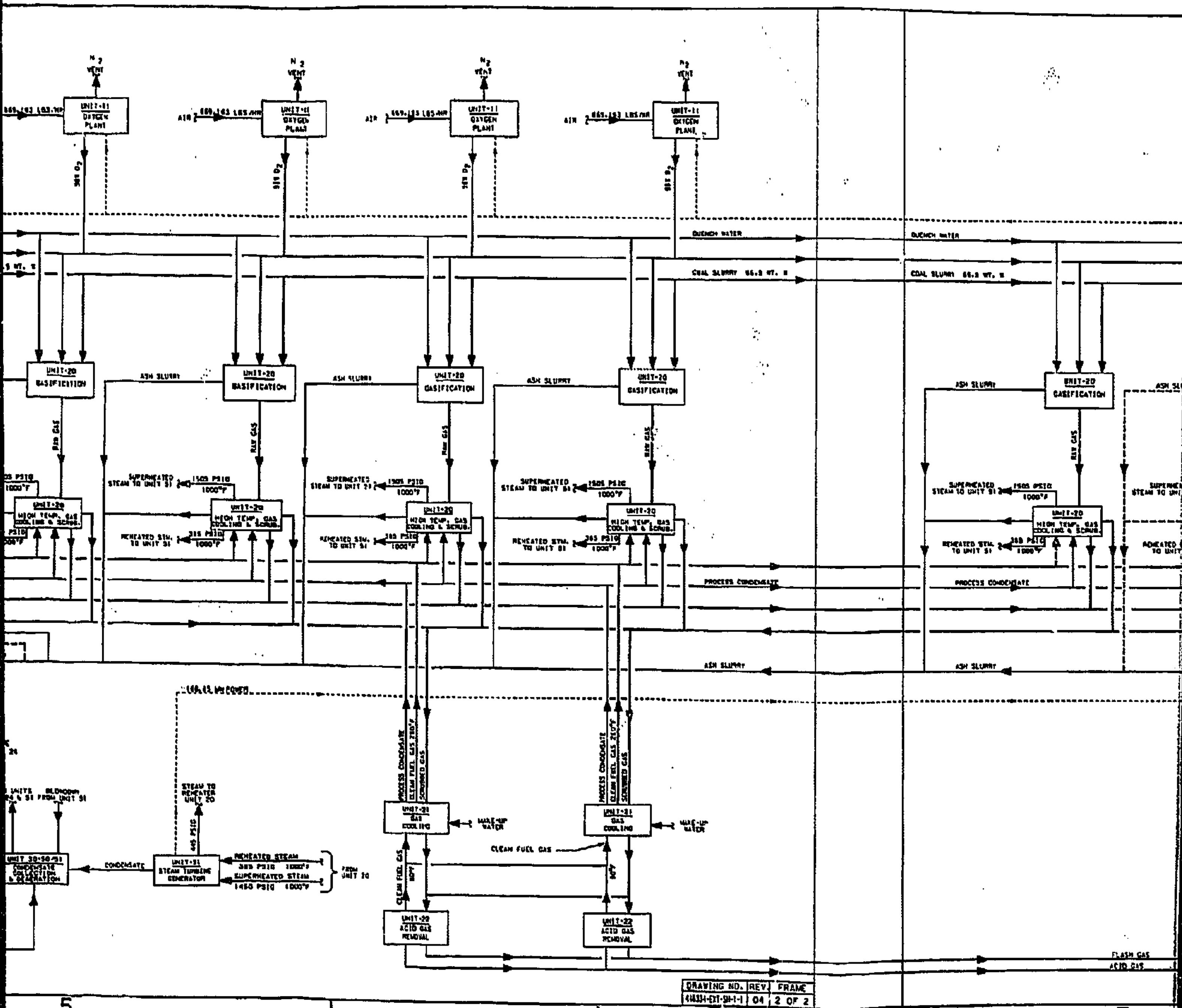

141.83 MW POWER
QUENCH WATER
PROCESS CONDENSATE
COAL SLURRY
UNIT-20
GASIFICATION
ASH SLURRY
RAW GAS
STEAM TO UNIT 51
1905 PSIG
UNIT-20
HIGH TEMP. GAS
COOLING & SCRUB.
PROCESS CONDENSATE
REHEATED FUEL GAS
600°F
UNIT 50
FUEL GAS
EXPANDER
GENERATOR
POWER
NET
POWER
PRODUCTION
NET FUEL
GAS EXPORT
195°F
UNIT-23
SULFUR
RECOVERY
TAIL GAS
STM TO
PROCESS
& MISC.
ATM
UNIT-24
TAIL GAS
TREATING
SULFUR
PRODUCT
NOTES:
1. SPARE TRAINS ARE SHOWN IN BROKEN LINES.
2. FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3. NET POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
FLUOR
IRVINE, CALIFORNIA
R. WHITE
OVERALL BLOCK FLOW DIAGRAM
TEXACO PROCESS COAL GASIFICATION
OXYGEN BLOWN
SATURATED STEAM BASE CASE
EPRI
NONE
448334-EXT-SS-1-1
04

2


ILLINOIS NO. 6 COAL
UNIT-10 COAL RECEIVING & CONVEYING
UNIT-10 COAL PULVERIZATION
UNIT-20 GASIFICATION
ASH SLURRY
SUPERHEATED STEAM TO UNIT 51
REHEATED STM. TO UNIT 51
UNIT-51 STEAM TURBINE GENERATOR
REHEATED STEAM
SUPERHEATED STEAM
FROM UNIT 20
UNIT-30 WATER TREATING
RAW WATER
TO UNIT 24
MAKE-UP WATER
QUENCH WATER
COAL SLURRY
CONDENSATE
BLOWDOWN
WASTE WATER

6

5

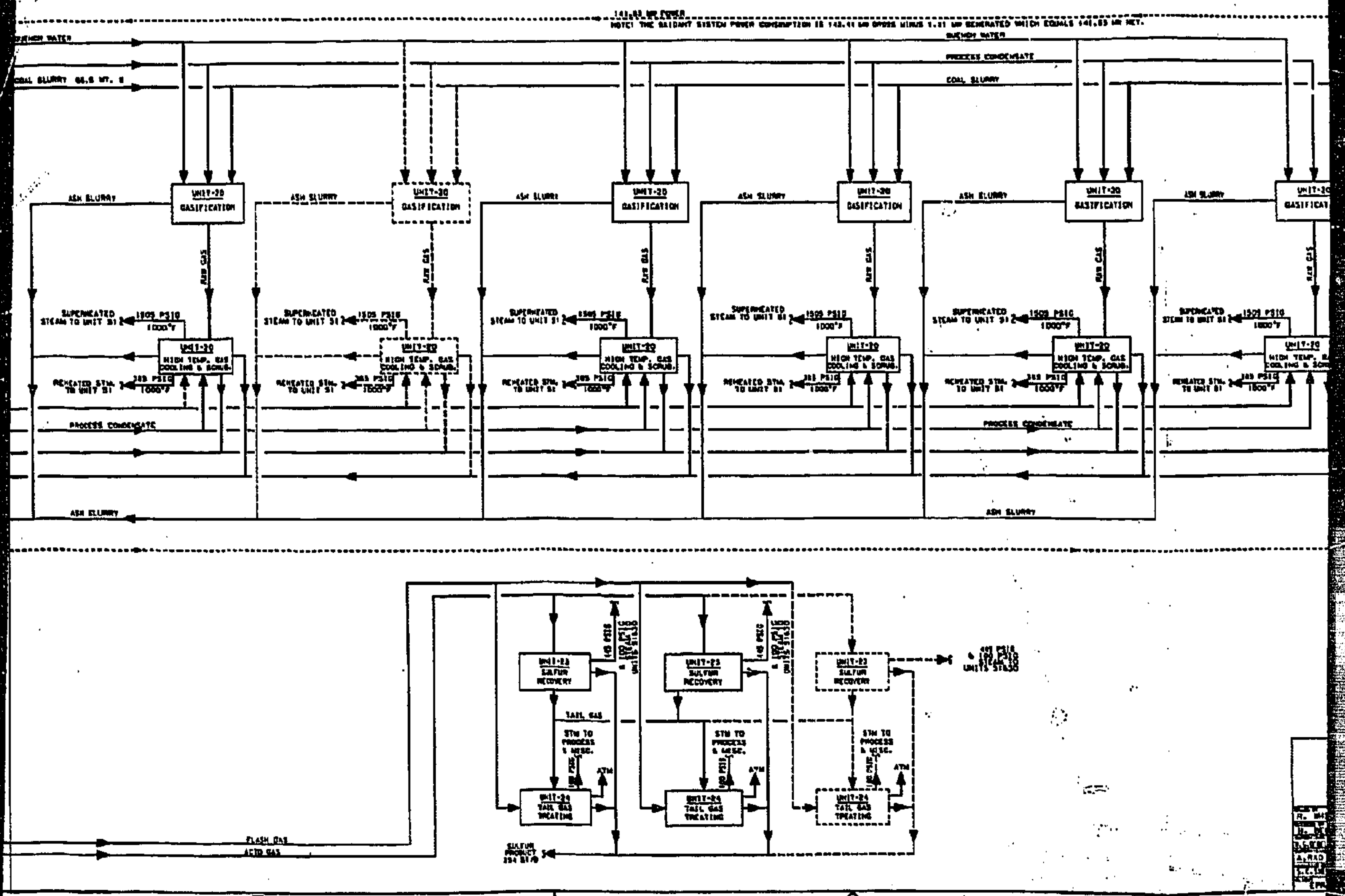


N2 VENT
UNIT-11 OXYGEN PLANT
AIR
669,183 LBS/HR
UNIT-20 GASIFICATION
ASH SLURRY
RAW GAS
QUENCH WATER
COAL SLURRY 66.8 WT. %
SUPERHEATED STEAM TO UNIT 51
1505 PSIG 1000°F
UNIT-20 HIGH TEMP. GAS COOLING & SCRUB.
REHEATED STM. TO UNIT 51
385 PSIG 1000°F
PROCESS CONDENSATE
PROCESS CONDENSATE
CLEAN FUEL GAS 250°F
SCRUBBED GAS
UNIT-21 GAS COOLING
MAKE-UP WATER
CLEAN FUEL GAS
UNIT-22 ACID GAS REMOVAL
FLASH GAS
ACID GAS
STEAM TO REHEATER UNIT 20
UNIT-51 STEAM TURBINE GENERATOR
REHEATED STEAM 385 PSIG 1000°F
SUPERHEATED STEAM 1450 PSIG 1000°F
FROM UNIT 20
CONDENSATE
UNIT 30-50-51 CONDENSATE COLLECTION & DEAERATION
BLOWDOWN FROM UNIT 51
DRAWING NO.
REV.
FRAME
04
2 OF 2

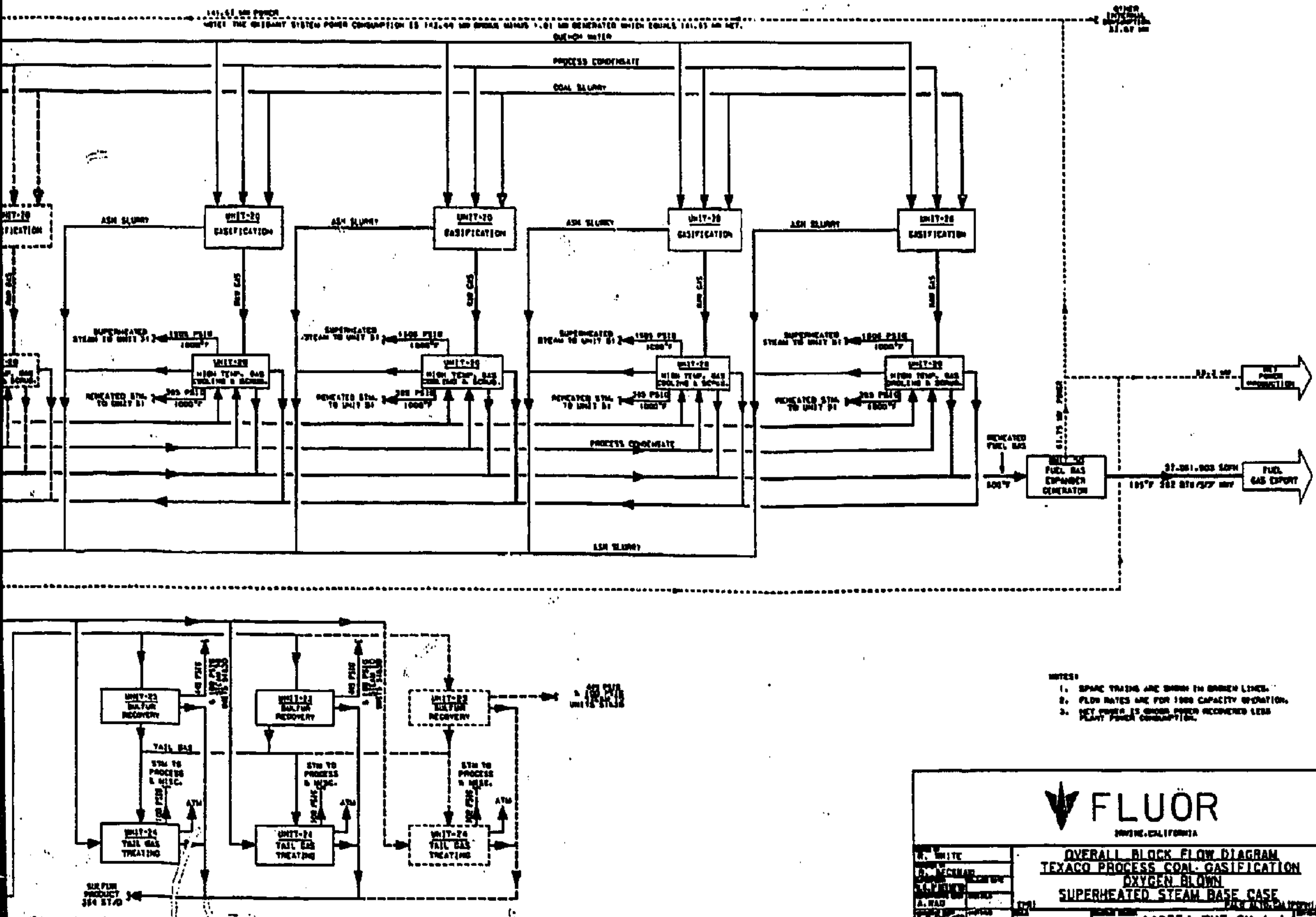
5

4

3



QUENCH WATER
PROCESS CONDENSATE
COAL SLURRY
UNIT-20
GASIFICATION
ASH SLURRY
RAW GAS
SUPERHEATED STEAM TO UNIT 51
1000°F
UNIT-20
HIGH TEMP. GAS COOLING & SCRUB.
REHEATED STM. TO UNIT 51
UNIT-23
SULFUR RECOVERY
TAIL GAS
STM TO PROCESS & MISC.
ATM
UNIT-24
TAIL GAS TREATING
SULFUR PRODUCT
FLASH GAS
ACID GAS
3
2



NOTES:
1. SPARE TRAINS ARE SHOWN IN BROKEN LINES.
2. FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3. NET POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.

FLUOR
IRVINE, CALIFORNIA

OVERALL BLOCK FLOW DIAGRAM
TEXACO PROCESS COAL GASIFICATION
OXYGEN BLOWN
SUPERHEATED STEAM BASE CASE

SCALE: NONE

448334-EXT-SH-1-1

04

COAL HANDLING AND GRINDING/PULVERIZATION, AND SLURRY PREPARATION

Process Flow Diagram EXT-SS-10-1 depicts the arrangement of equipment, which incorporates one train of coal unloading, stacking, reclamation, and conveying, followed by five operating and one spare trains of grinding.

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, which provides about four minutes of storage based on the capacity of the stacking system at 4125 tons per hour. Four vibrating feeders, 10-FE-1A-D, withdraw coal from the hopper and place it on receiving conveyor 10-CV-1. 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 supplies sampling system 10-SA-1. From 10-CV-2, storage conveyor 10-CV-3 transports the coal to tripper 10-TR-1, which supplies double boom stacker 10-ME-1. The stacker travels on tracks and forms up to 3-1/2 day (38,500 tons) live storage piles on either side. Total live storage is limited to seven days to reduce the possibility of spontaneous ignition. The unloading and stacking system is designed to handle a three day supply in eight hours.

Space for a reserve dead pile of up to 60 days storage is provided adjacent to the rail unloading station. The normal dead pile size is assumed to be 23 days. Total capital requirement presented in this report 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.

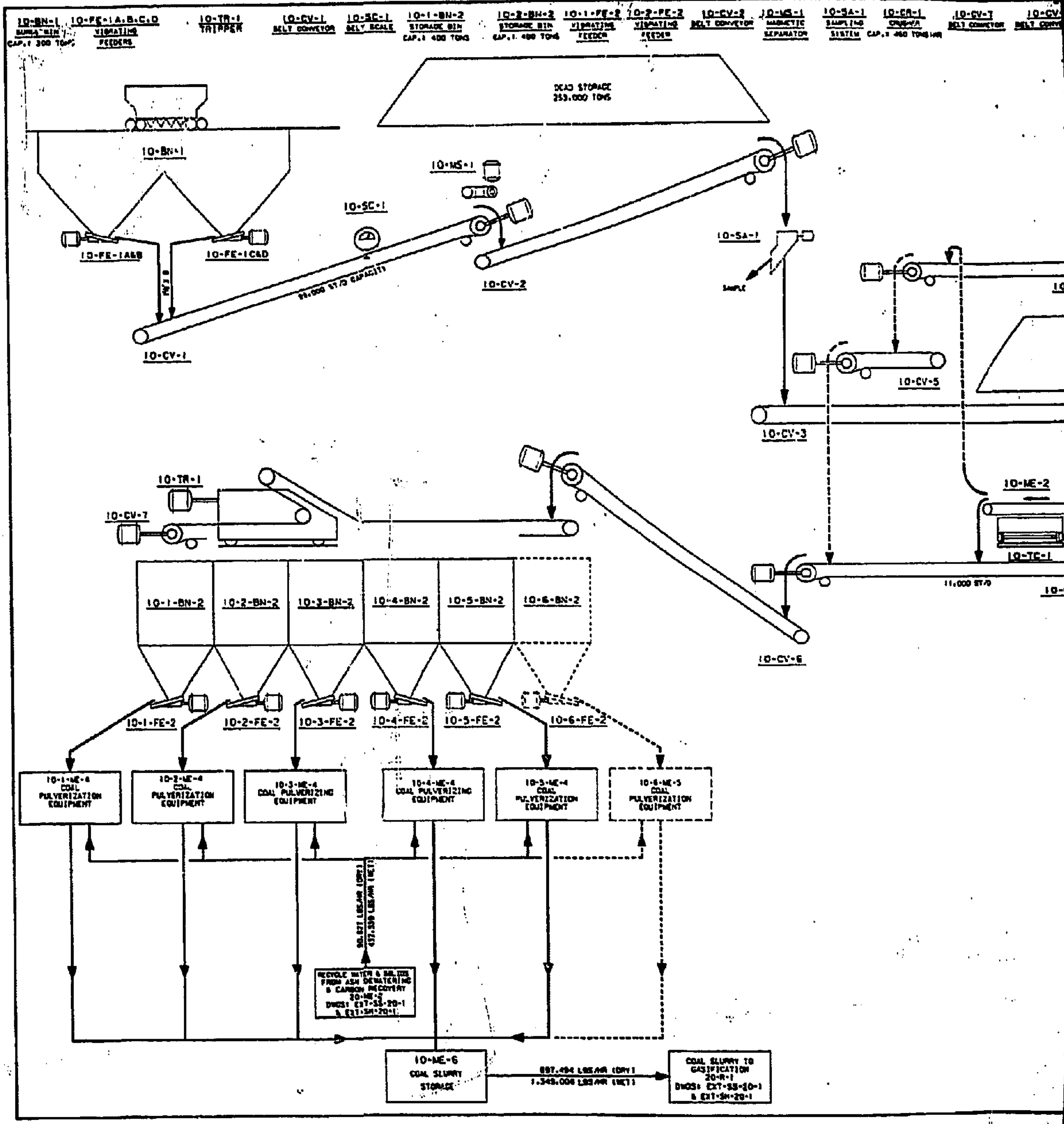
Coal is reclaimed from the storage piles by a bridge-type bucket wheel reclaimer 10-ME-2, rated at 460 tons per hour. This machine is moved between live storage piles as necessary by transfer car 10-TC-1. The wheel moves across the face of the pile, making an angle of repose cut across the many layers 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 conveyor 10-CV-6, which is located near 10-CV-4B. Coal conveyor 10-CV-7 delivers the coal to storage bins 10-BN-2, which provide a total of about 4-1/2 hours of downstream throughput. Vibrating feeders 10-FE-2 supply the grinding mills, which grind the coal wet with recycle process water and solids.

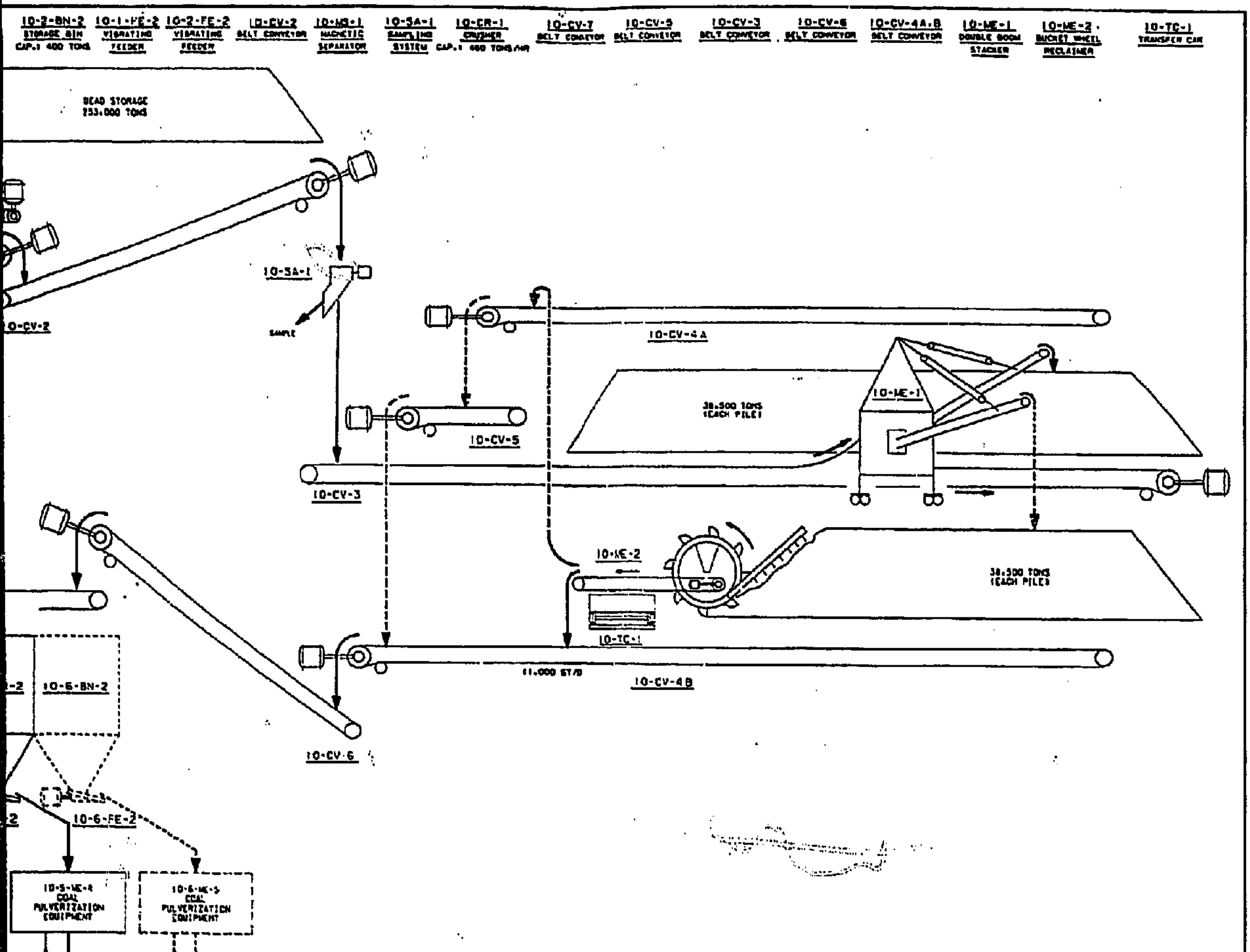
The coal slurry is transferred to holding tanks from the mill discharge tanks and finally stored in two tanks of 12 hours capacity. The 66.5 percent solids slurry is then pumped by eight parallel charge pumps to the eight operating gasifiers of nominal 1375 ton/day coal capacity.

All unloading and conveying systems are 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.

Equipment Notes

All the equipment is commercially available.


10-BN-1
CAP.: 300 TONS
10-FE-1A,B,C,D
VIBRATING FEEDERS
10-TR-1
TRIPPER
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
DEAD STORAGE
253,000 TONS
10-BN-1
10-FE-1A&B
10-FE-1C&D
10-SC-1
10-MS-1
10-CV-1
10-CV-2
10-SA-1
SAMPLE
10-CV-5
10-CV-3
10-TR-1
10-CV-7
10-ME-2
10-TC-1
11,000 ST/D
10-CV-6
10-1-BN-2
10-2-BN-2
10-3-BN-2
10-4-BN-2
10-5-BN-2
10-6-BN-2
10-1-FE-2
10-2-FE-2
10-3-FE-2
10-4-FE-2
10-5-FE-2
10-6-FE-2
10-1-ME-4 COAL PULVERIZATION EQUIPMENT
10-2-ME-4 COAL PULVERIZATION EQUIPMENT
10-3-ME-4 COAL PULVERIZING EQUIPMENT
10-4-ME-4 COAL PULVERIZING EQUIPMENT
10-5-ME-4 COAL PULVERIZATION EQUIPMENT
10-6-ME-5 COAL PULVERIZATION EQUIPMENT
RECYCLE WATER & SOLIDS FROM ASH DEWATERING & CARBON RECOVERY 20-ME-2
DWGS: EXT-SS-20-1 & EXT-SM-20-1
10-ME-6
COAL SLURRY STORAGE
897,494 LBS/HR (DRY)
1,349,006 LBS/HR (WET)
COAL SLURRY TO GASIFICATION 20-R-1
DWGS: EXT-SS-20-1 & EXT-SM-20-1



NOTE:
1. ALL FLOW RATES AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY UNLESS OTHERWISE NOTED.
2. SPARE TRAINS ARE SHOWN IN BROKEN LINES.

FLUOR
IRVINE, CALIFORNIA

P.A. IMBROWE
B. ASH
A. RAO
E.C. SHELSER
EPRI

PROCESS FLOW DIAGRAM
COAL PREPARATION
TEXACO PROCESS-OXYGEN BLOWN, ALL CASES

EPRI PALO ALTO, CALIFORNIA

NONE | 448334-EXT-SS-10-1 | 04

4835K120M

OXIDANT FEED

Process Flow Diagram EXT-SS-11-1 shows the oxidant feed system design. The system has five parallel operating trains, each containing an air compression system, an air separation unit, and an oxygen compression system.

Atmospheric air at 14.4 psia 88°F is compressed to 95 psia in a two-stage axial-centrifugal machine 11-1-C-1. The first-stage and second-stage heats of compression are rejected to cooling water. The water condensed from the feed air in 11-1-E-1 is withdrawn from the bottom of the shell, while the water that condensed in 11-1-E-2 is collected in knockout drum 11-1-V-1. This collected condensate is used as makeup water for the power plant cooling tower system.

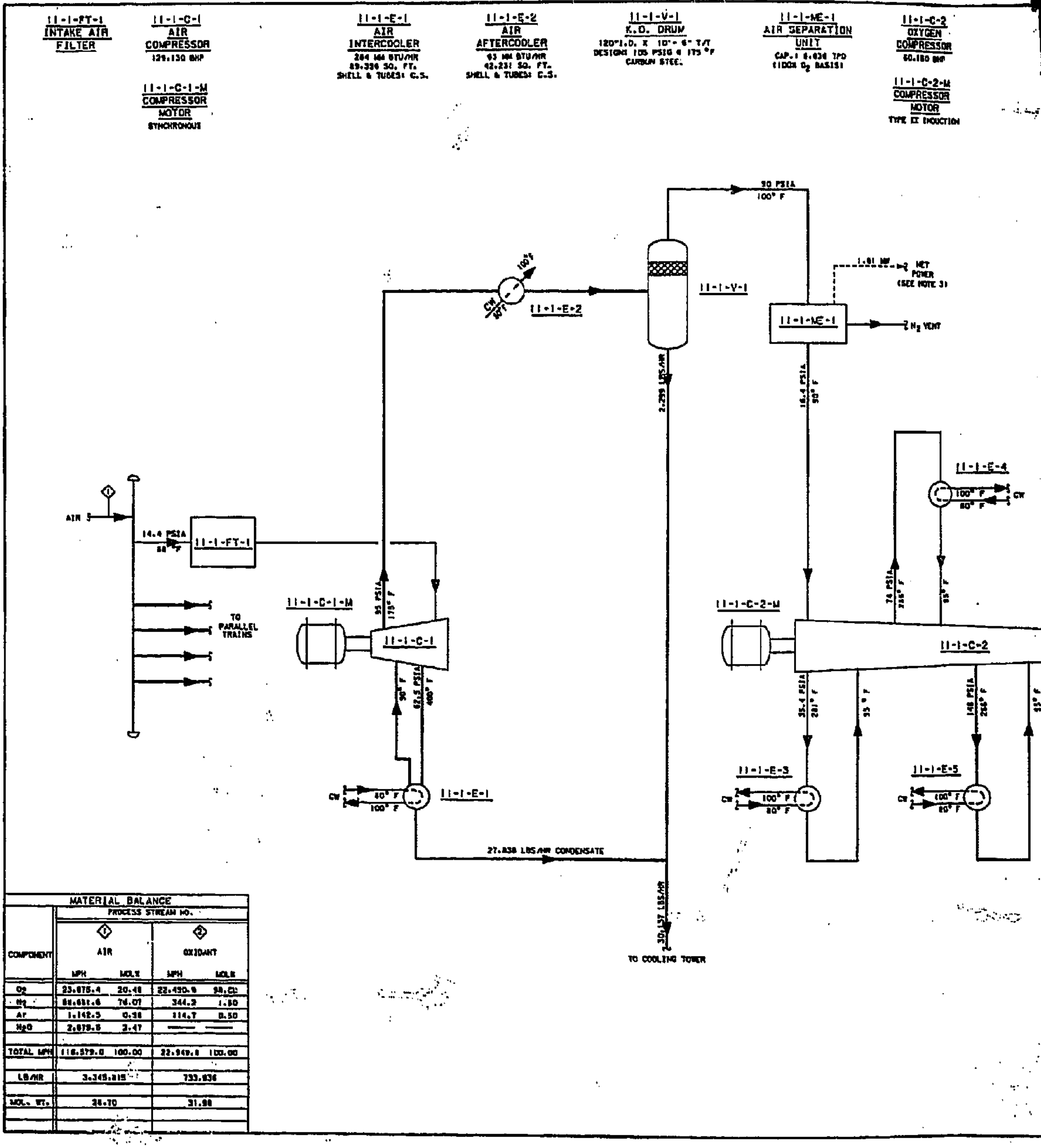
The compressed air at 90 psia 100°F is processed in a cryogenic air separation unit 11-1-ME-1, to produce 98 mole percent oxygen at a rate of 1727 tons of oxygen (100 percent basis) per train per day. The air separation unit operating parameters are typical of those for reversing exchanger plant design, which uses turboexpanders for refrigeration. These turboexpanders produce 1.81 MW of power which is available for plant export.

The 98 mole percent oxygen is discharged from the air separation unit at 16.4 psia and 90°F. This oxidant stream is compressed to 734 psia, prior to being fed to the gasifiers. Oxygen compression is accomplished in a centrifugal compressor consisting of two cases in series, with a speed-increasing gear between them. A total of four water intercooled stages, two in the first case and two in the second, are used in this design. The final discharge temperature is 300°F, which is judged to be within the design limits of commercial equipment.

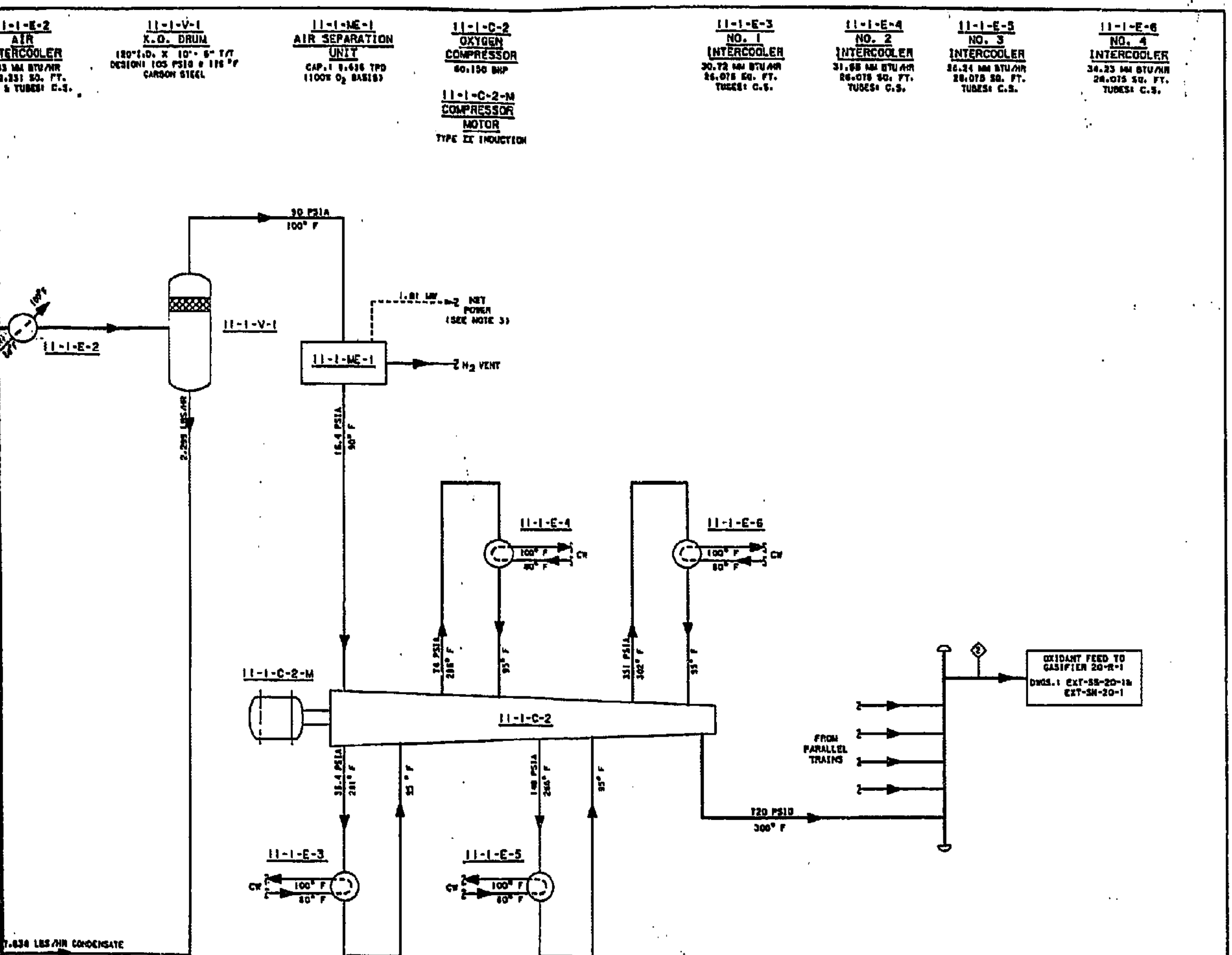
All of the air and oxygen compressors are electric motor driven. The startup of the coal gasification fuel gas plant is greatly simplified by using electric motors rather than steam turbines as drivers in the oxidant feed system . Additionally, the steam distribution and condensate collection systems are simplified by concentrating the higher pressure steam usages in the power generation section of the plant.

Equipment Notes

The compressors and cryogenic air separation plant are commercially available units. The use of water-cooled oxygen compressor intercoolers to obtain a 95°F interstage temperature lowers the required compression horsepower. Previous oxidant feed system designs in EPRI studies used air-cooled exchangers for this service. Minimizing power demands 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. For example, lowering the product oxygen concentration to 95 mole percent may lower the total oxidant feed system power demand by an additional 5 MW.



MATERIAL BALANCE				
	PROCESS STREAM NO.			
COMPONENT	1 AIR		2 OXIDANT	
	MPH	MOL%	MPH	MOL%
O_2	23,875.4	20.48	22,490.9	98.00
N_2	88,681.6	76.07	344.2	1.50
Ar	1,142.5	0.98	114.7	0.50
H_2O	2,879.5	2.47	——	——
TOTAL MPH	116,579.0	100.00	22,949.8	100.00
LB/HR	3,345,815		733,836	
MOL. WT.	28.70		31.98	



NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE OF FIVE PARALLEL TRAINS.
2. ALL FLOW RATES AND EQUIPMENT SPECIFICATIONS ARE ON A TOTAL PLANT BASIS AT FULL CAPACITY EXCEPT AS NOTED.
3. [illegible] THE NET POWER DERIVED FROM THE POWER RECOVERY EXPANDERS.

FLUOR
IRVINE, CALIFORNIA

P.A. IMBRONE
M.C. D'BELMITO
S.C. SMELSER
EPRI

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
TEXACO PROCESS-OXYGEN BLOWN,
ALL CASES EXCEPT SH3

EPRI
PALO ALTO, CALIFORNIA

SCALE: NONE
448334-EXT-SS-11-1
03

483SK1203

GASIFICATION

Process Flow Diagrams EXT-SS-20-1 and EXT-SH-20-1 show the gasification, raw gas cooling, and particulate removal steps for the base case oxygen-blown Texaco fuel gas plants which employ the 1375 ton/day coal gasifiers and the following two alternate energy recovery schemes:

- Saturated steam (1505 psig) generation by recovery of high-level sensible heat from the gasifier effluent gas (Case EXT-SS)
- Superheated steam (1450 psig 1000°F) generation by recovery of high-level sensible heat from the gasifier effluent gas (Case EXT-SH)

Eight operating and two spare gasification trains are provided, along with two trains of ash handling/carbon recovery equipment. The 20-ME-2 and 20-1-ME-4 "boxes" on the flow diagrams represent proprietary sections of the Texaco coal gasification process that contain many equipment items.

The coal slurry and oxygen combine at the gasifier burners, which are oriented downward from the top head of the gasifier. The burners contain cooling coils through which tempered water is circulated. The gasifier 20-1-R-1 operates at a pressure of 600 psig and temperatures in the range of 2300°F to 2600°F. These temperatures are sufficiently above the ash flow point to ensure free flowing molten slag. A portion of the coal feed burns, providing heat for the endothermic gasification reactions. The coal's hydrogen and carbon therefore react to form CO, CO_2, H_2, and very little CH_4, while the sulfur is converted to H_2S and COS. Nitrogen in the coal is converted to free nitrogen (N_2) and a small quantity of ammonia. Fluor has assumed that ammonia entering with recycled slurry water is effectively eliminated by dissociation and combustion reactions in the gasifier.

Energy Recovery

EXT-SS. In the saturated steam base case, hot crude gas with molten ash at 2400°F enters the radiant waste heat boiler 20-1-E-1 where high-pressure saturated steam is generated by recovery of high-level sensible heat. This waste heat boiler is of vertical downflow design with tubes around the walls of the vessel. The second heat recovery unit 20-1-E-2 is a vertical convective boiler unit with water tubes. Due to the uncertain nature of these designs, a process contingency of 20 percent is applied to the estimated total cost of the radiant boiler and a process contingency of 25 percent to the estimated total convective boiler cost.

Raw gas leaving the convective high-pressure saturated steam generator is further cooled by heat exchange to reheat the clean fuel gas and to heat boiler feed water in exchangers 20-1-E-3, 20-1-E-4, and 20-1-E-5. In the first exchanger 20-1-E-3, clean fuel gas that has been reheated to 400°F in exchanger 20-1-E-5 is further reheated to 600°F, prior to being sent to the gas expanders. Boiler feedwater at 349°F flows from the fired heater to be heated to 598°F in exchanger 20-1-E-4.

EXT-SH. Hot crude gas with molten ash at 2400°F is used for energy recovery in this superheated steam base case also. The configuration of the energy recovery equipment for high-pressure superheated steam generation (1450 psig 1000°F) is as follows:

- A vertical radiant boiler (20-1-E-1),
- Followed by a convective reheater with steam in the shell (20-1-E-2A), and
- A convective superheater with steam in the shell (20-1-E-2B).

A 20 percent process contingency is applied to the total cost of the radiant boiler, and a 40 percent process contingency is applied to the total cost of the convective steam superheater and reheater due to the highly uncertain nature of these designs.

Raw gas leaving the superheated steam generation and reheating equipment is cooled further in two heat exchangers that provide for clean fuel gas reheating and boiler feedwater heating in exchangers 20-1-E-3 and 20-1-E-4. Clean fuel gas that has been reheated to 280°F in downstream exchanger (21-1-E-2) is further reheated to 600°F in exchanger 20-1-E-4. Boiler feedwater heated up to 290°F in downstream exchanger (21-1-E-1) is heated to 598°F in exchanger 20-1-E-3.

Particulate Removal

The cooled particulate-bearing raw gas enters Gas Scrubbing Unit 20-1-ME-4, where contact with recycled process condensate results in virtually complete removal of solids. This solids-free raw gas flows to Unit 21, Gas Cooling.

Ash Handling and Carbon Recovery

For both base cases, regardless of the energy recovery process, most of the coal ash is converted to molten slag which falls into a water quench at the bottom of the radiant boiler vessel. Solids entrained in the exit gas are captured in gas scrubbing unit 20-1-ME-4. Two parallel ash dewatering/carbon recovery systems serve all the operating gasifiers. The resulting ash cake, assumed to contain 30 weight percent water, is transported to landfill disposal by rail cars. Part of the reclaimed process water is recycled to the slag quench and coal slurry areas. A slipstream of 170 gpm reclaimed process water is purged to a proprietary Texaco water treating process, in order to avoid chloride buildup, and for the removal of slag and soot particles, dissolved metals, formates, sulfides, and ammonia. This water treating unit is included in the general facilities section of these fuel gas plants. The remainder of the reclaimed process water along with the carbon rich solids is recycled to coal grinding.

Equipment Notes

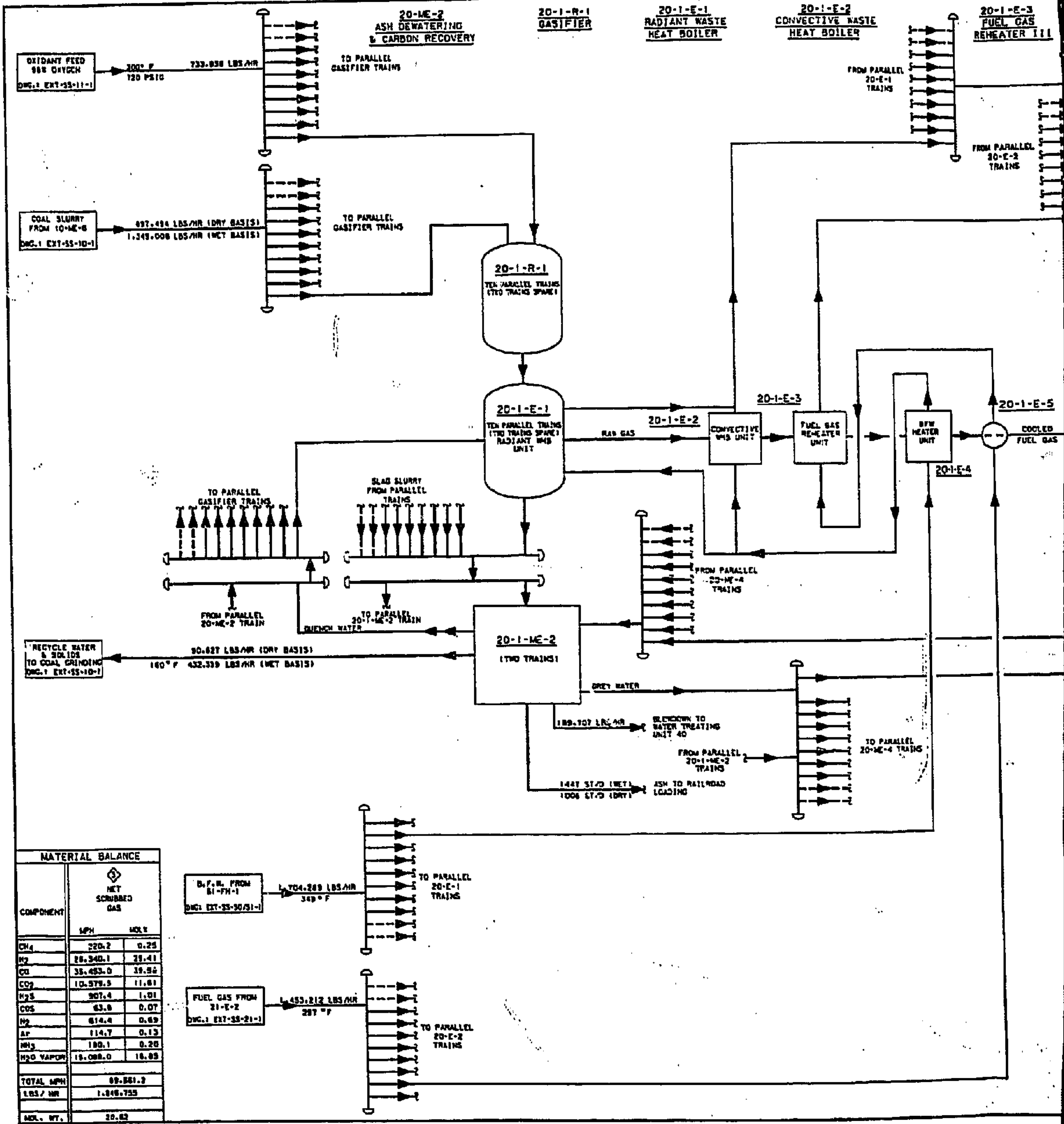
The Texaco gasifier is commercially proven for the gasification of liquid hydrocarbons. Commercial experience with coal gasification is limited. One Texaco coal gasifier has been operating for over three years in Germany at about 560 psig. This gasifier handles only 150 tons/day of coal, a much lower throughput than that of the gasifiers used in this study. The Texaco coal gasification research facility at Montebello, California, is presently testing coals in a gasifier which operates at over 1000 psig.

The slag dewatering system is composed of commercially proven equipment.

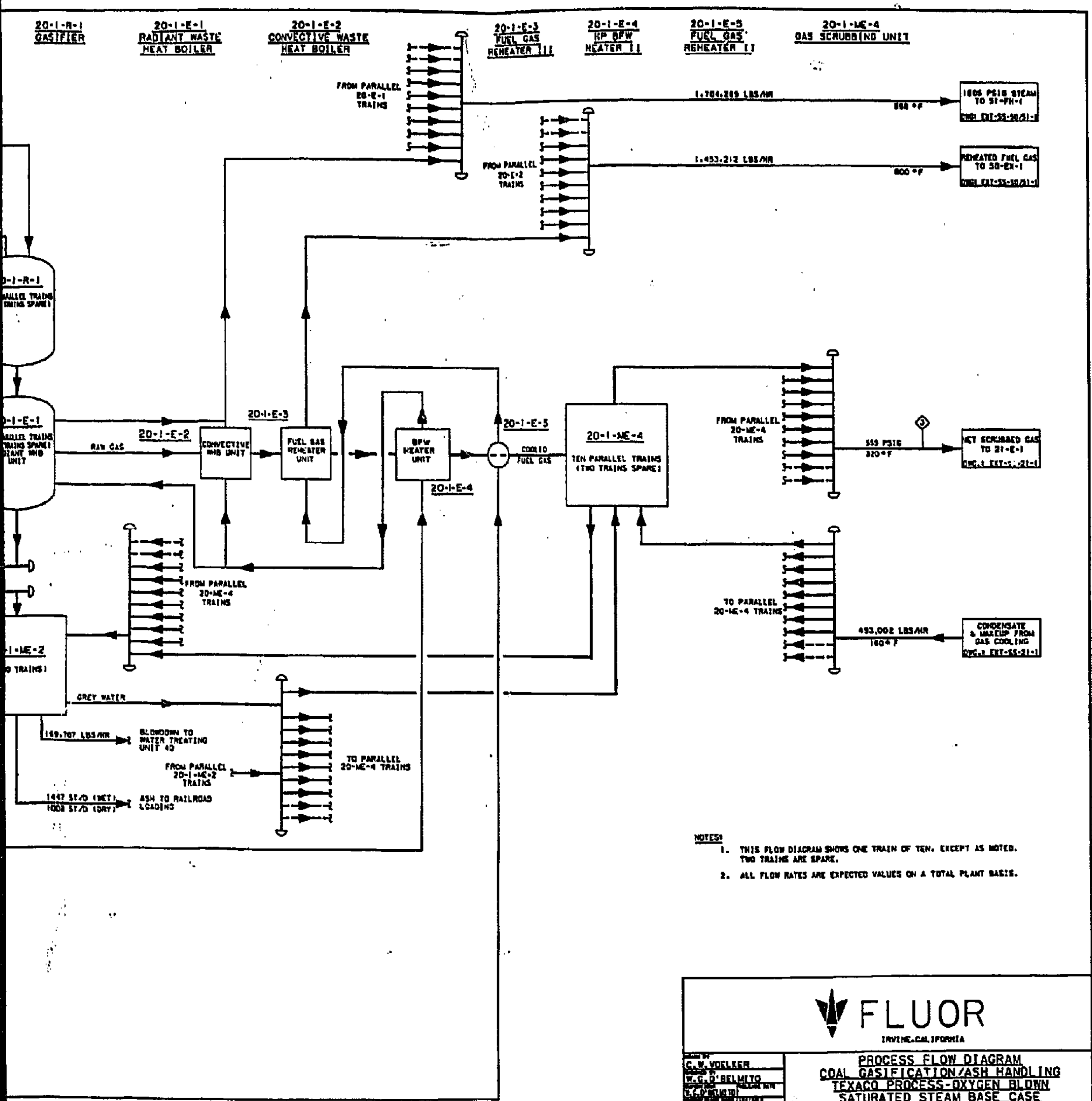
The gas scrubbing unit equipment is commercially available.

The key features in these designs center on the heat transfer equipment used for high-level sensible heat recovery in the two base cases. In Case EXT-SS, 1505 psig saturated steam is generated in unconventional radiant and convective boilers. Such installations have been tested in the 150 ton per day German plant. The superheated steam base case design employs a 1505 psig steam superheater and a 445 psig steam reheater configuration (both superheated and reheated steam temperatures of 1000°F) which is wholly conceptual at this point. A gasification process which operates at temperatures similar to those in the Texaco process has reported superheating 750 psig steam for a very limited time in a pilot plant unit. The designs and cost estimates, adopted in this study, were based on those developed by major waste heat boiler manufacturers.

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 materials costs is beyond the scope of the present work.



MATERIAL BALANCE		
COMPONENT	⟨3⟩ NET SCRUBBED GAS	
	MPH	MOL %
CH_4	220.2	0.25
H_2	26,340.1	29.41
CO	35,453.0	39.56
CO_2	10,579.5	11.81
H_2S	907.4	1.01
COS	63.8	0.07
N_2	614.4	0.69
Ar	114.7	0.13
NH_3	180.1	0.20
H_2O VAPOR	15,088.0	16.85
TOTAL MPH	89,561.2	
LBS/HR	1,846,755	
MOL. WT.	20.62	



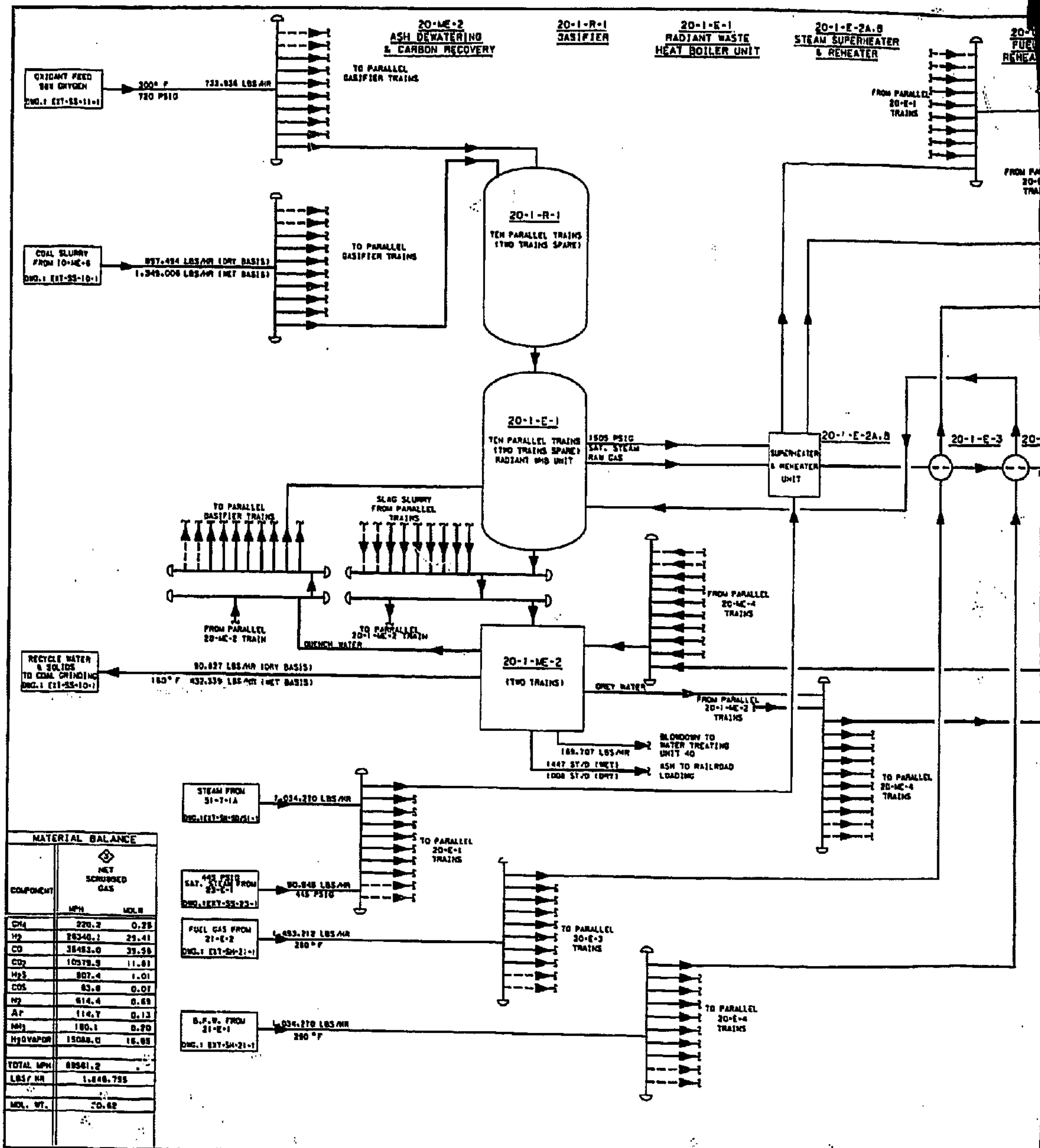
NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF TEN, EXCEPT AS NOTED. TWO TRAINS ARE SPARE.
2. ALL FLOW RATES ARE EXPECTED VALUES ON A TOTAL PLANT BASIS.

FLUOR
IRVINE, CALIFORNIA

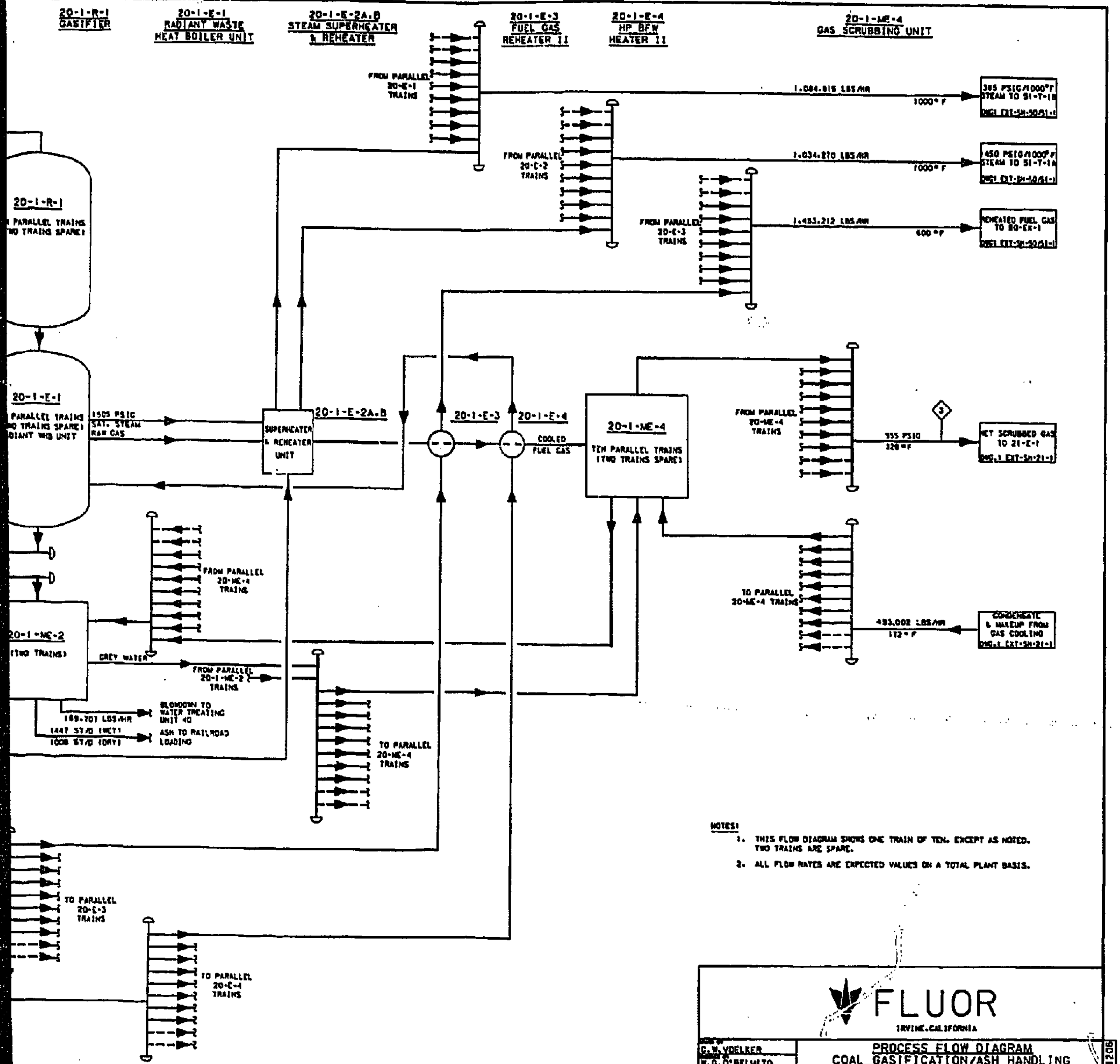
PROCESS FLOW DIAGRAM
COAL GASIFICATION/ASH HANDLING
TEXACO PROCESS-OXYGEN BLOWN
SATURATED STEAM BASE CASE

C.W. VOELKER
W.G. D'BELMITO
A. RAO
C.C. GRISER
EPRI
PALO ALTO, CALIFORNIA
NONE
448334-EXT-SS-20-1
04



MATERIAL BALANCE

COMPONENT	③ NET SCRUBBED GAS MPH	MOL %
CH_4	220.2	0.25
H_2	26340.1	29.41
CO	35453.0	39.59
CO_2	10579.5	11.81
H_2S	907.4	1.01
COS	63.6	0.07
N_2	614.4	0.69
Ar	114.7	0.13
NH_3	180.1	0.20
H_2O VAPOR	15088.0	16.85
TOTAL MPH	89561.2	
LBS/HR	1,846,795	
MOL. WT.	20.62	



NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF TEN, EXCEPT AS NOTED. TWO TRAINS ARE SPARE.
2. ALL FLOW RATES ARE EXPECTED VALUES ON A TOTAL PLANT BASIS.

FLUOR
IRVINE, CALIFORNIA

C.W. VOELKER
W.O. D'BELMITO
W.O. D'BELMITO
A. RAO
S.C. SMELSER
EPRI

PROCESS FLOW DIAGRAM
COAL GASIFICATION/ASH HANDLING
TEXACO PROCESS-OXYGEN BLOWN
SUPERHEATED STEAM BASE CASE

EPRI
PALO ALTO, CALIFORNIA

NONE	448334-EXT-SH-20-1	04

483SK1206

GAS COOLING

Process Flow Diagrams EXT-SS-21-1 and EXT-SH-21-1 show one of two parallel trains in this section for the saturated steam and superheated steam base cases, respectively. No spare train is provided for either case.

Solids-free raw gas from the particulate scrubbing section is cooled to 105°F on the tube side of a series of exchangers. Ammonia is then removed in an ammonia scrubber.

EXT-SS. In the saturated steam case diagram EXT-SS-21-1, saturated solids-free gas at 320°F is cooled to 304°F on the tube side of the boiler feedwater heater I, 21-1-E-1, where high-pressure boiler feedwater from the deaerator is heated from 251°F to 290°F. The raw gas is next cooled in fuel gas reheater I, 21-1-E-2, by reheating clean fuel gas leaving the acid gas removal unit, from 80°F to 262°F. Further cooling of the raw gas down to 140°F is accomplished in a vacuum condensate and makeup heater, 21-1-E-3.

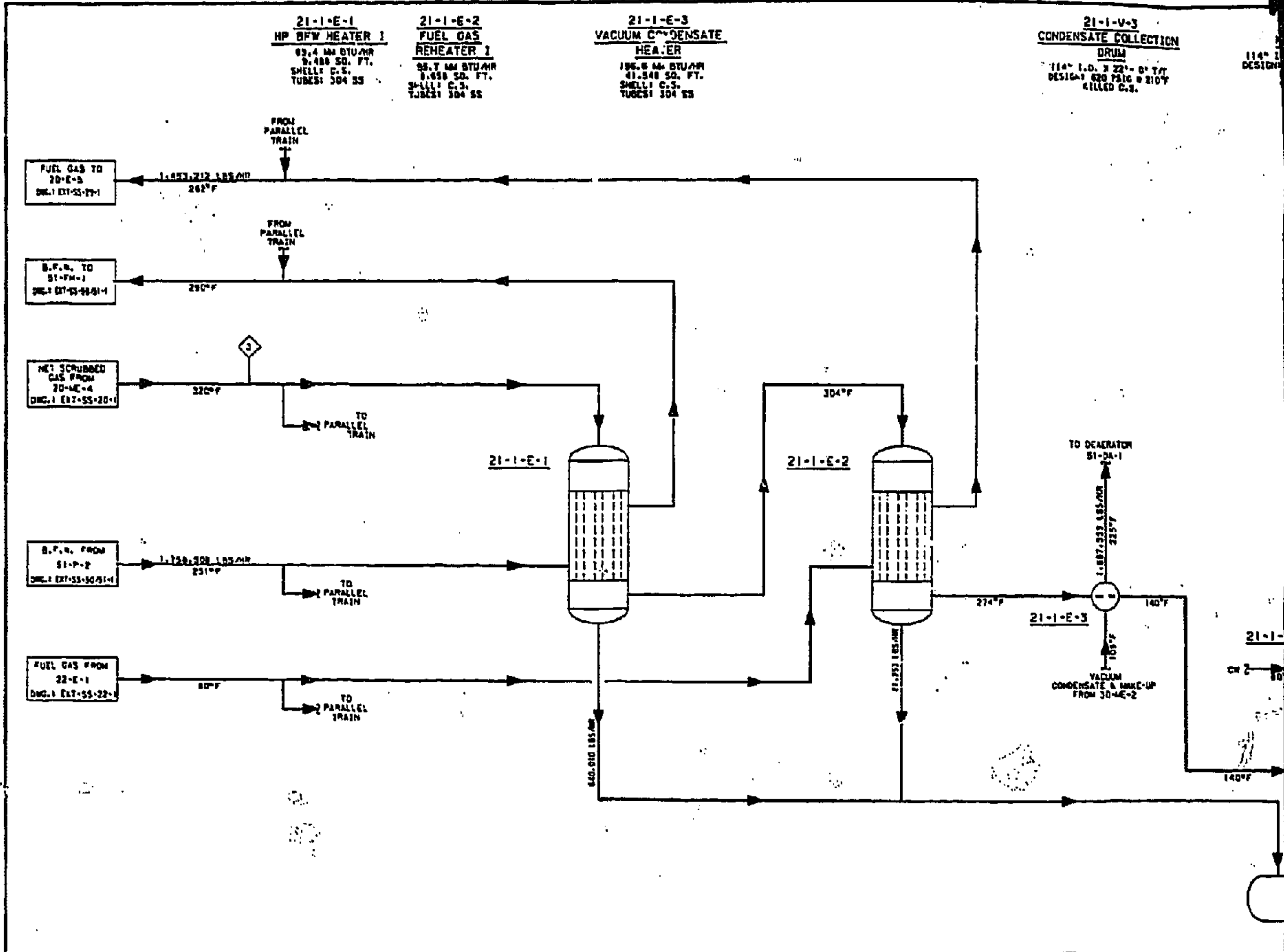
Process condensate from the exchangers is collected in collection vessel 21-1-V-3. This hot condensate along with makeup water flows under pressure to the particulate scrubbing section 20-ME-4.

The overhead gases from the condensate collection vessel are further cooled by cooling water to 105°F in trim cooler 21-1-E-4, before entering ammonia absorber 21-1-V-2, which contains six sieve-type trays. Ammonia is removed down to one ppm by contacting the gas countercurrently with raw water at 70°F. Absorber overhead gas at 100°F flows to the acid gas removal unit for removal of H_2S and COS. The liquid flow, from the bottom of the ammonia scrubber, is combined with some of the hot process condensate from 21-1-V-1 and makeup in collection drum 21-1-V-3 before the total stream enters particulate scrubbing section 20-ME-4.

EXT-SH. In the superheated steam case diagram EXT-SH-21-1, saturated solids-free gas at 320°F is cooled to 310°F on the tube side of high-pressure boiler feedwater heater I, 21-1-E-1, while heating the boiler feedwater from the deaerator from 253°F to 290°F. The raw gas is further cooled to 281°F in fuel gas reheater I, 21-1-E-2, wherein clean fuel gas leaving the acid gas removal unit is heated from 80°F to 280°F. The raw gas along with the condensate leaving the exchanger at 281°F is cooled to 226°F in a vacuum condensate and makeup heater, 21-1-E-3. Next, the raw gas is cooled down to 140°F in an air cooler. An air cooler was used because this low-level thermal energy cannot be effectively utilized in the heat integration scheme. Process condensate from the exchangers and overhead gas from condensate collection vessel are handled in the same manner as in the saturated steam case.

Equipment Notes

All equipment is commercially available.



MATERIAL BALANCE				
	PROCESS STREAM NO.			
	3		4	
COMPONENT	NET SCRUBBED GAS		CRUDE GAS	
	MPH	MOL %	MPH	MOL %
CH_4	220.2	0.23	220.2	0.30
H_2	26,340.1	29.41	26,340.1	35.36
CO	35,453.0	39.58	35,453.0	47.62
CO_2	10,579.5	11.81	10,579.5	14.21
H_2S	901.4	1.01	907.4	1.22
COS	63.8	0.07	63.8	0.09
N_2	614.4	0.65	614.4	0.82
Ar	114.7	0.13	114.7	0.15
NH_3	180.1	0.20	0.1	0.00
H_2O VAPOR	15,088.0	16.83	155.6	0.21
TOTAL MPH	89,581.2		74,449.0	
LBS/HR	1,846,755		1,57?,675	
MOL. WT.	20.62		21.15	

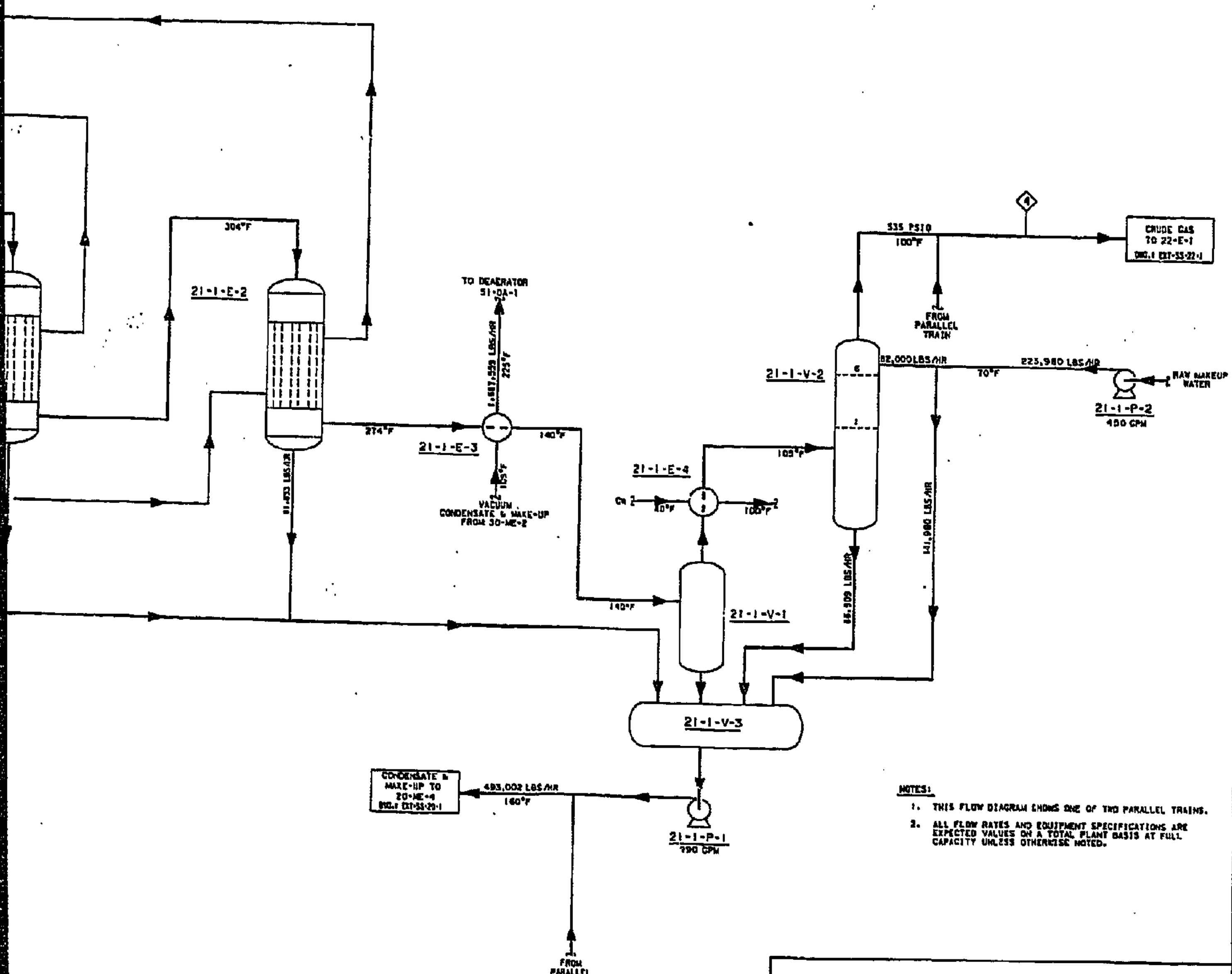
21-1-E-3
VACUUM CONDENSATE HEATER
196.6 MM BTU/HR
41,348 SQ. FT.
SHELL: C.S.
TUBES: 304 SS

21-1-V-3
CONDENSATE COLLECTION DRUM
114" I.D. X 22'- 0" T/T
DESIGN: 620 PSIG @ 310°F
KILLED C.S.

21-1-V-1
K.O. DRUM
114" I.D. X 18'- 0" T/T
DESIGN: 620 PSIG @ 310°F
KILLED C.S.

21-1-E-4
RAW GAS TRIM COOLER
24.4 MM BTU/HR
7,878 SQ. FT.
SHELL: C.S.
TUBES: 304 SS

21-1-V-2
AMMONIA SCRUBBER
108" I.D. X 20'- 0" T/T
DESIGN: 620 PSIG @ 280°F
KILLED C.S. W/410 SS TRAYS



NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE OF TWO PARALLEL TRAINS.
2. ALL FLOW RATES AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY UNLESS OTHERWISE NOTED.

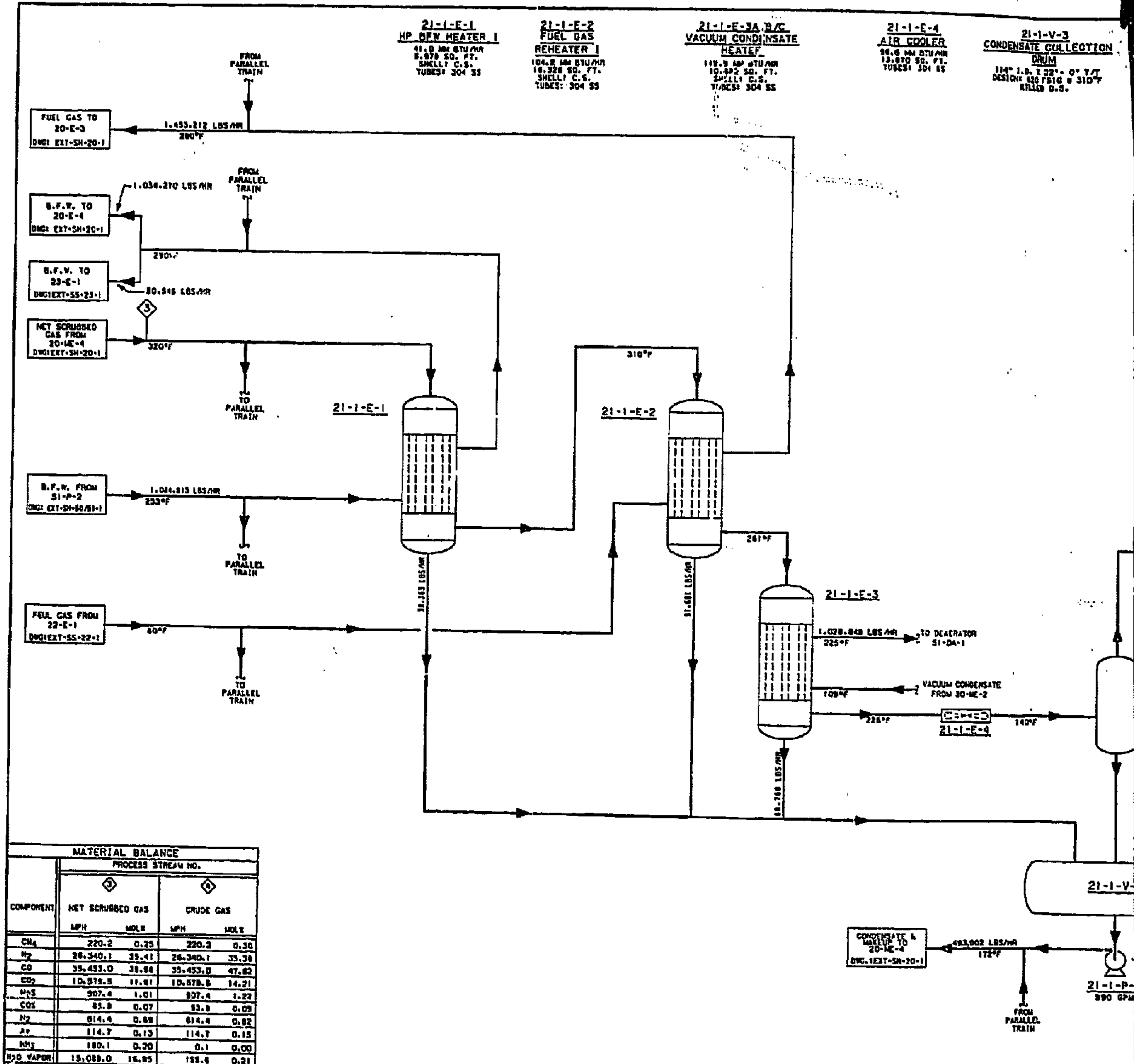
FLUOR
IRVINE, CALIFORNIA

DRAWN BY	P.A. IMBRONE
CHECKED BY	W.O. O'BELMITO
	E.C. O'BELMITO
	A. RAO
	S.C. SHELSER
	EPRI

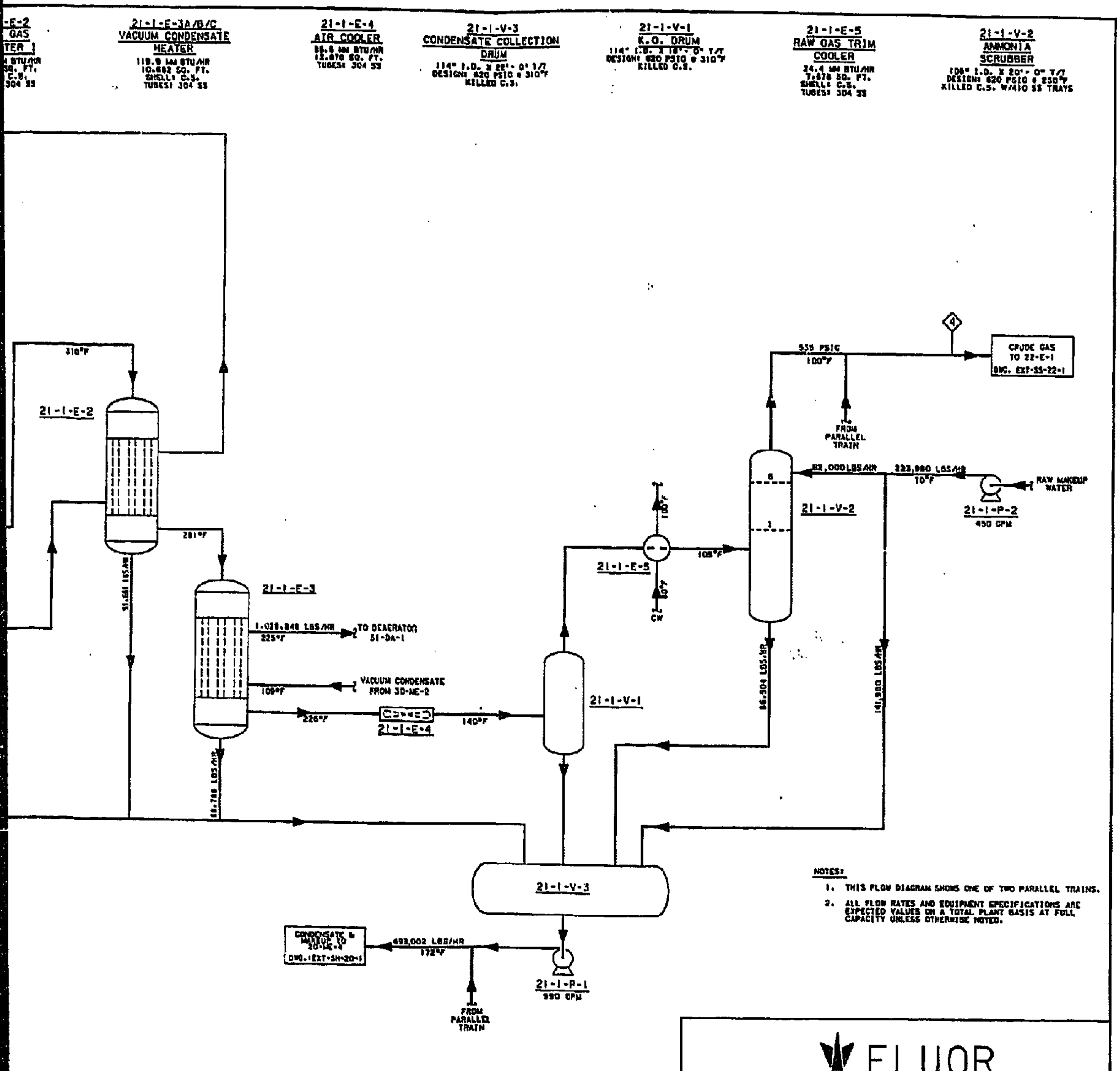
PROCESS FLOW DIAGRAM
GAS COOLING
TEXACO PROCESS-OXYGEN BLOWN,
SATURATED STEAM BASE CASE
EPRI
PALO ALTO, CALIFORNIA

SCALE	DRAWING NUMBER	REV
NONE	448334-EXT-SS-21-1	04

403SK1205



MATERIAL BALANCE				
	PROCESS STREAM NO.			
	3		4	
COMPONENT	NET SCRUBBED GAS		CRUDE GAS	
	MPH	MOL%	MPH	MOL%
CH_4	220.2	0.25	220.2	0.30
H_2	26,340.1	29.41	26,340.1	35.38
CO	35,453.0	39.58	35,453.0	47.62
CO_2	10,579.5	11.81	10,578.6	14.21
H_2S	907.4	1.01	907.4	1.22
COS	63.8	0.07	63.8	0.09
N_2	614.4	0.69	614.4	0.82
Ar	114.7	0.13	114.7	0.15
NH_3	180.1	0.20	0.1	0.00
H_2O VAPOR	15,088.0	16.85	155.6	0.21
TOTAL MPH	89,581.2		74,449.0	
LBS/HR	1,846,755		1,574,678	
MOL. WT.	20.62		21.15	



NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE OF TWO PARALLEL TRAINS.
2. ALL FLOW RATES AND EQUIPMENT SPECIFICATIONS ARE EXPECTED VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY UNLESS OTHERWISE NOTED.

FLUOR
IRVINE, CALIFORNIA

D. CLAVEAU
W. G. D'BELMITO
A. RAO
J. C. SMELSER
EPRI

PROCESS FLOW DIAGRAM
GAS COOLING
TEXACO PROCESS-OXYGEN BLOWN
SUPERHEATED STEAM BASE CASE

EPRI PALO ALTO, CALIFORNIA

NONE | 448334-EXT-SH-21-1 | 04

ACID GAS REMOVAL

Process Flow Diagram EXT-SS-22-1 depicts one of the two parallel acid gas removal trains. This same design is used for both the saturated and superheated steam base cases. No spare train is provided.

The acid gas removal system employs the Selexol® process for selective removal of hydrogen sulfide (H_2S). This process involves the absorption of 99.0 percent of the entering hydrogen sulfide and 34.6 percent of the entering carbonyl sulfide (COS) from the plant.

The 100°F gas from the ammonia scrubber is cooled by heat exchange with treated fuel gas in feed/fuel gas exchanger 22-1-E-1. The cooled gas then flows through acid gas absorber 22-1-V-1, where it contacts Selexol® solvent countercurrently over a packed bed. The treated gas from the absorber flows through knockout drum 22-1-V-5 for recovery of solvent mist, and is warmed to 80°F against incoming feed gas in 22-1-E-1. Further reheating of the clean fuel gas to 600°F occurs in Units 20 and 21.

The rich solvent from the absorber is reduced in pressure through a hydraulic turbine 22-1-HT-1, which supplies about half of the power required by lean solution pump 22-1-P-1. This solvent stream then enters flash drum 22-1-V-2 where 90 percent of the sulfur-free combustible gases disengage from the loaded solvent. This flash gas is used as a reducing gas in the tail gas treating unit. However, approximately 98 percent of the H_2S and COS are retained in the loaded solvent because of their selective absorption.

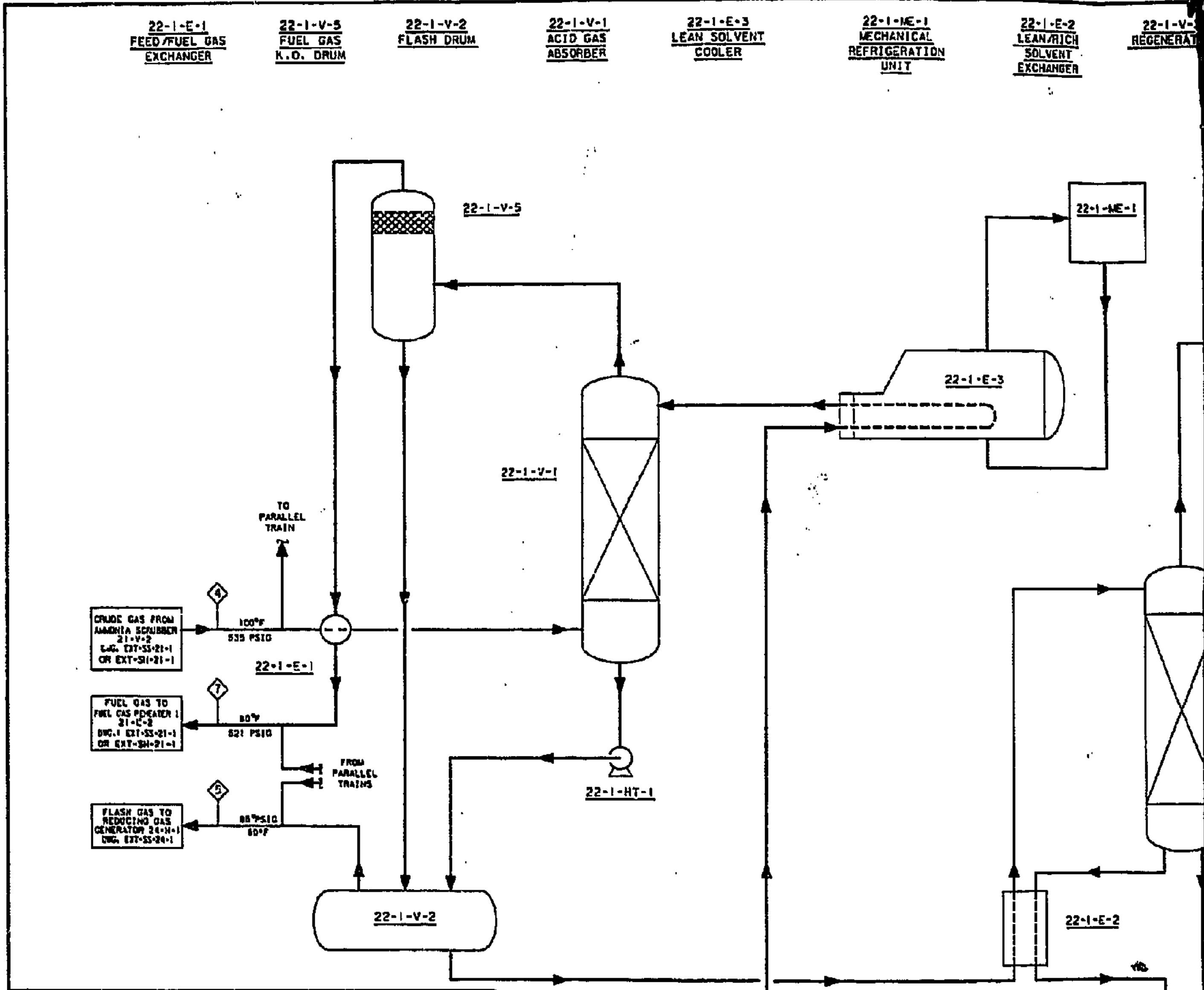
The loaded or rich solvent from the flash drum is heated by exchange with regenerated lean solvent in plate exchanger 22-1-E-2 and flows to the top of regenerator 22-1-V-3. Absorbed H_2S, COS, CO_2, H_2O, and minor amounts of other components are stripped from the solution by application of heat, supplied by condensing 100 psig steam in the regenerator reboiler 22-1-E-4. The regenerated solvent is cooled in lean/rich solvent exchanger 22-1-E-2, then is pumped back to absorber 22-1-V-1 through lean solvent cooler 22-1-E-3. Solvent cooling in 22-1-E-3 is provided by 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-5. The condensate resulting from this cooling step is separated in knockout drum 22-1-V-4 and is refluxed to the regenerator. The acid gas ultimately sent to sulfur recovery contains about 38 volume percent H_2S. Temperature in the overhead receiver, expected to be 120°F, will be adjusted to maintain the unit water balance.

Refrigeration System

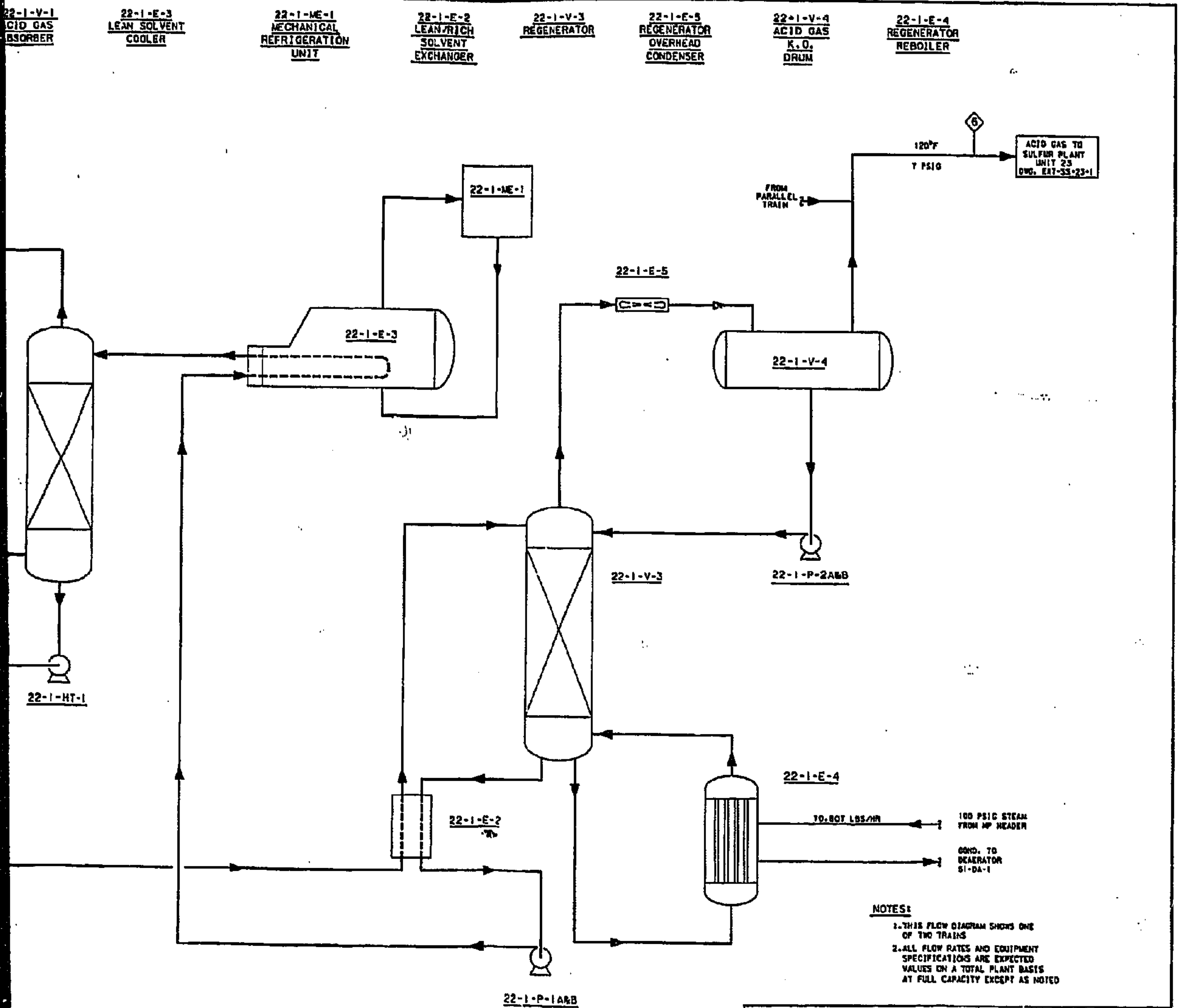
The refrigeration system employed is a typical packaged fluorocarbon unit. The compressor, receiver, and condensing equipment are fabricated on skids and installed near lean solvent cooler 22-1-E-3.

Equipment Notes

The majority of equipment in this section is carbon steel. This equipment has been used in similar service for several years. Plate-type exchangers for the lean/rich solvent exchanger service are less costly than conventional shell-and-tube exchangers for this service.



MATERIAL BALANCE								
	PROCESS STREAM NUMBER							
COMPONENT	4 CRUDE GAS		5 FLASH GAS		6 ACID GAS		7 FUEL GAS	
	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	220.2	0.30	1.7	0.38	0.6	0.02	217.9	0.30
H_2	26,340.1	35.38	46.3	10.03	2.6	0.10	26,292.2	36.87
CO	35,453.0	47.62	129.6	28.71	16.3	0.60	35,307.1	49.53
CO_2	10,579.8	14.21	254.1	58.28	1,619.3	60.05	8,706.1	12.21
H_2S	907.4	1.22	17.1	3.78	880.9	32.67	9.4	0.01
COS	63.8	0.09	1.5	0.33	20.6	0.76	41.7	0.06
N_2	614.4	0.82	1.4	0.31	0.1	0.00	612.9	0.86
Ar	114.7	0.15	0.2	0.04	0.0	0.00	114.5	0.16
NH_3	0.1	0.00	0.0	0.00	0.1	0.00	0.0	0.00
H_2O	155.6	0.21	0.6	0.13	155.3	5.80	0.0	0.00
TOTAL MPH	74,449.0	100.00	451.6	100.00	2,695.8	100.00	71,301.6	100.00
TOTAL LB/HR	1,574,678		15,663		103,816		1,453,212	
MOL WT	21.15		34.68		38.24		20.38	



NOTES:

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

FLUOR
IRVINE, CALIFORNIA

C.R. BETTS
W.G. D'BELMITO
W.G. D'BELMITO
A. RAO
S.C. SWELSEN
EPRI

PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM TEXACO PROCESS-OXYGEN BLOWN, ALL CASES EXCEPT SS5, SH4 & SH5

EPRI PALO ALTO, CALIFORNIA

NONE | 448334-EXT-SS-22-1 | 04

4635K1208

SULFUR PLANT

Process Flow Diagram EXT-SS-23-1 describes the basic sulfur plant design used for both the base cases. The entire sulfur plant system consists of two 50 percent parallel operating trains and one 50 percent spare train. Total sulfur recovery in this system is 27,039 lb/hr (324.5 ST/D) or 87 percent of the sulfur fed in the coal. Another 2,451 lb/hr (29.4 ST/D) of sulfur is recovered in Beavon/Stretford Unit 24. This additional recovery boosts the total recovered sulfur to approximately 95 percent of that contained in the coal feed.

The sulfur plant is a two-stage, acid gas bypass-type Claus unit. About one-third of the 120°F acid gas from the Selexol® unit is burned in a sulfur furnace, 23-1-H-1, thereby converting H_2S to H_2O and SO_2. Air for the combustion in the furnace is supplied by blower 23-1-BL-1. Heat from the combustion products is recovered by generating 455 psig steam in waste heat boiler 23-1-E-1. The 900°F exhaust gas from the sulfur furnace is mixed with the flow of acid gas which by-passes the furnace and the resultant 597°F gas mixture is fed to 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 in the mixture which is slightly greater than 2:1 to force the converter reaction toward completion.

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

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

This exothermic reaction is catalyzed by a bed of Kaiser S-501 alumina catalyst contained within the converter and produces a 181°F gas temperature rise. Since the converter reaction is limited by thermodynamic equilibrium, complete conversion of the H_2S and SO_2 to elemental sulfur is not achieved.

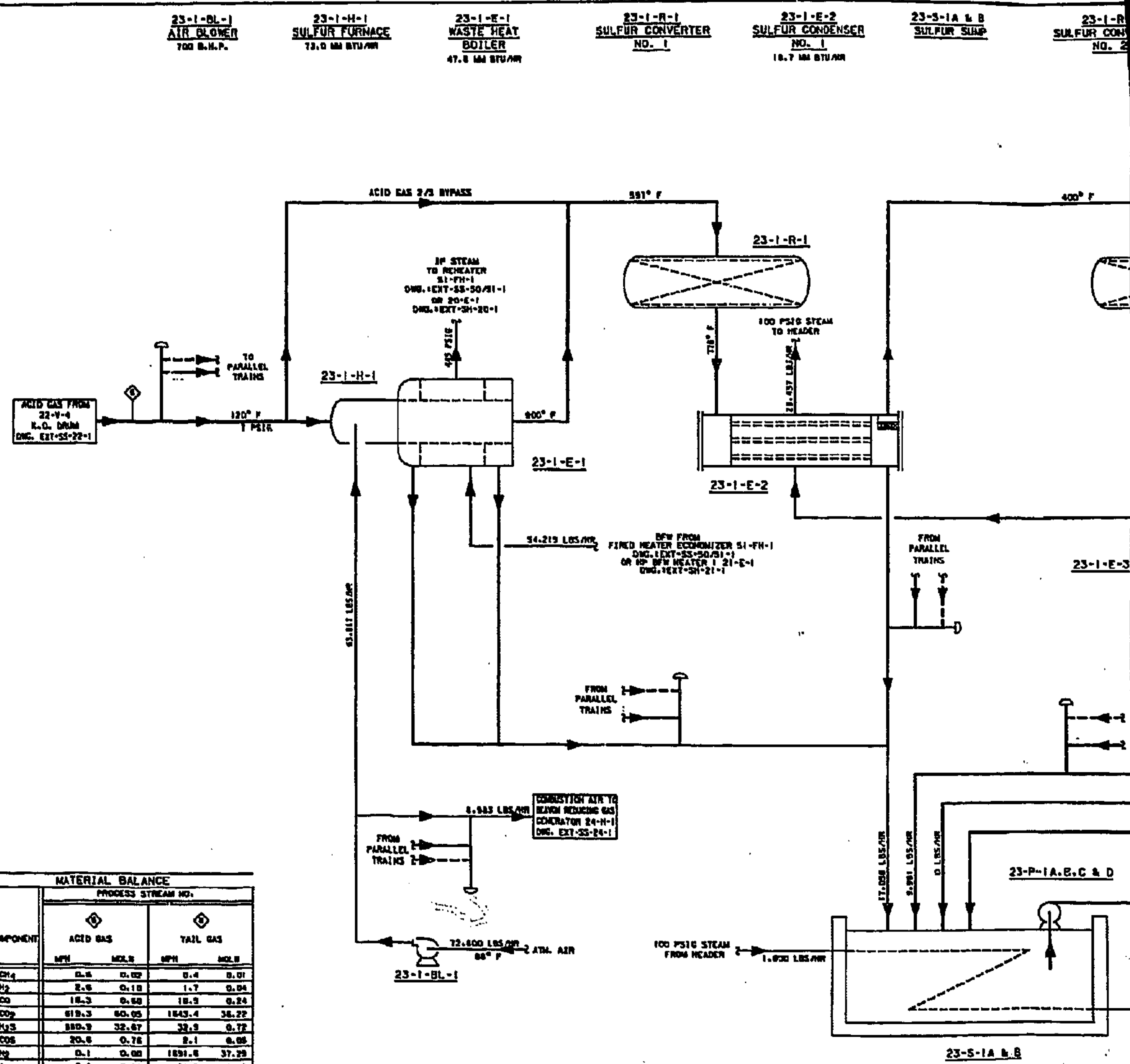
The gaseous sulfur produced in the first converter is condensed and recovered by cooling the effluent gas to 400°F in 23-1-E-2. Steam at 100 psig is generated by this cooling process. The sulfur (17,058 lb/hr) condenses in the tubes and flows by gravity to one of two concrete sumps, 23-S-1A&B. Sulfur, a solid at ambient temperatures, is kept molten by condensing 100 psig steam in pipe coils that cover the bottom of the sumps.

The 400°F gases from 23-1-E-2 react further in sulfur converter No. 2, 23-1-R-2, and produce a 91°F gas temperature rise. Sulfur (9,981 lb/hr) in the exhaust gas is condensed and cooled to 285°F in 23-1-E-3 by heat transfer to medium-pressure boiler feedwater. The condensed sulfur then flows to one of the sumps.

Tail gas at 285°F, still containing about 1,776 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.

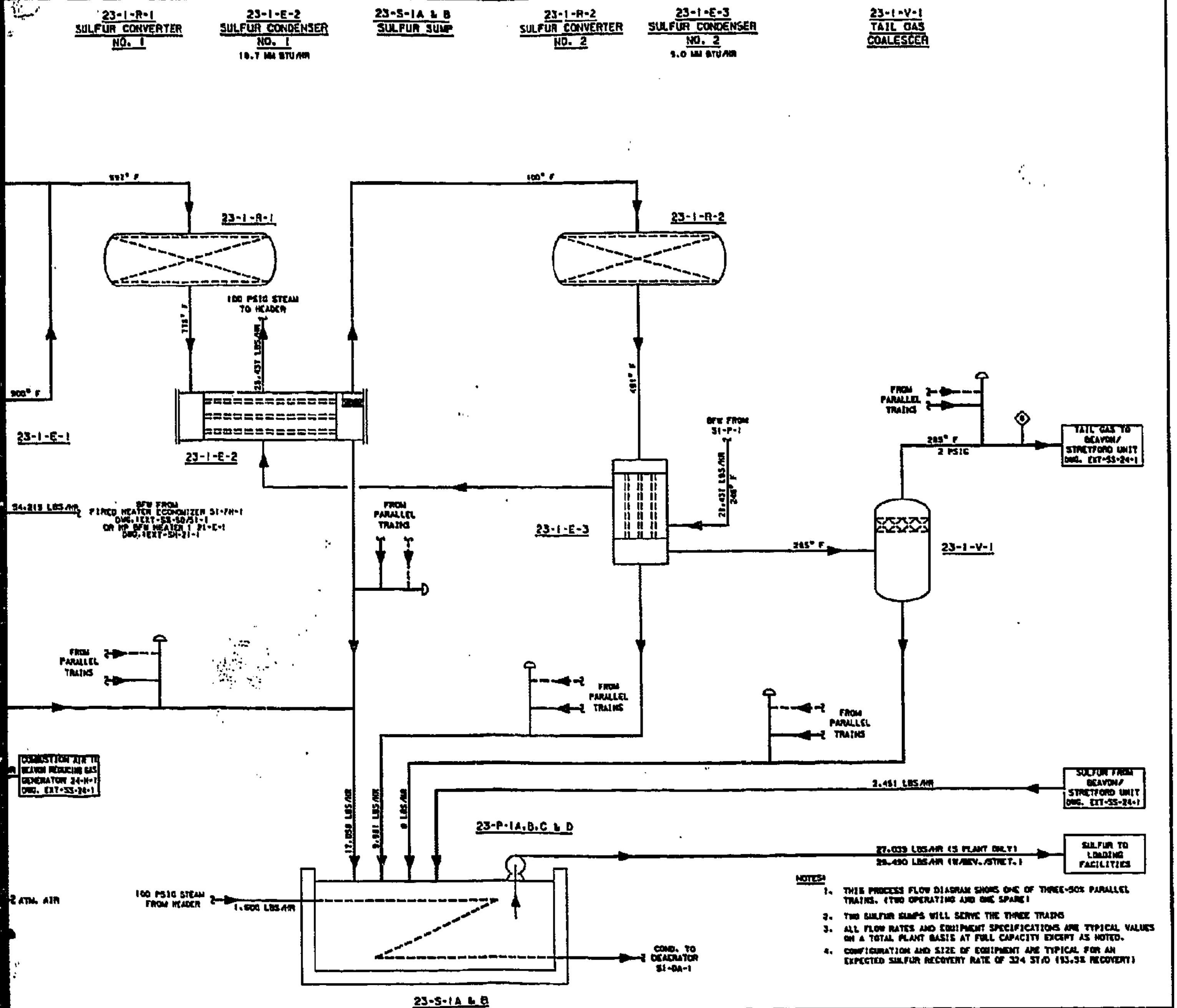
Equipment Notes

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



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NO.			
	ACID GAS		TAIL GAS	
	MPH	MOL %	MPH	MOL %
CH_4	0.6	0.02	0.4	0.01
H_2	2.6	0.10	1.7	0.04
CO	16.3	0.60	18.5	0.24
CO_2	619.3	60.05	1643.4	36.22
H_2S	880.9	32.67	32.9	0.72
COS	20.6	0.76	2.1	0.05
N_2	0.1	0.00	1691.6	37.29
Ar	0.0	0.00	21.8	0.48
NH_3	0.1	0.00	0.1	0.00
H_2O VAPOR	158.3	5.88	1115.4	24.58
SO_2	0.0	0.00	16.6	0.36
S_6	0.0	0.00	0.1	0.00
S_8	0.0	0.00	.5	0.01
VAPOR MPH	2,698.8	100.00	4,537.4	100.00
TOTAL LBS/HR	105,816		143,448	
MOL. WT. MPH	39.24		31.62	



FLUOR
IRVINE, CALIFORNIA

D. CLAVEAU
W.G. O'BELMITO
A. RAO
EPRI

PROCESS FLOW DIAGRAM
SULFUR PLANT (TYPICAL)
TEXACO PROCESS-OXYGEN BLOWN,
ALL CASES EXCEPT SS5, SH4 & SH5

EPRI PALO ALTO, CALIFORNIA

NONE	448334-EXT-SS-23-1	04

483SK1209

TAIL GAS TREATING

Process Flow Diagram EXT-SS-24-1 describes the Beavon/Stretford system design used for both the base cases. As in the sulfur recovery unit, two 50 percent parallel operating trains and a third identical spare train are provided.

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

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

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

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

$$COS + H_2O \rightarrow CO_2 + H_2S \qquad (2\text{-}3)$$

$$S_6 + 6\ H_2 \rightarrow 6\ H_2S \qquad (2\text{-}4)$$

$$S_8 + 8\ H_2 \rightarrow 8\ H_2S \qquad (2\text{-}5)$$

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. Instead, flash gas from the acid gas removal unit supplies the necessary hydrogen and carbon monoxide. The flash gas is partially combusted in reducing gas generator 24-1-H-1, and then mixed with the tail gas stream. The resulting inlet temperature to the Beavon hydrogenation reactor 24-1-V-7 is

650°F. The sulfur conversion reactions listed above, as well as the following "shift" reaction, take place in 24-1-V-7:

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad (2\text{-}6)$$

The effluent from 24-1-V-7 is cooled to 400°F through generation of 100 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:

$$2\ Na_2CO_3 + 2\ H_2S \rightarrow 2\ NaHCO_3 + 2\ NaHS \qquad (2\text{-}7)$$

$$2\ NaHS + 2\ NaHCO_3 + 4\ NaVO_3 \rightarrow 2\ Na_2CO_3 + H_2O + S_2 + Na_2V_4O_9 + 2\ NaOH \qquad (2\text{-}8)$$

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.

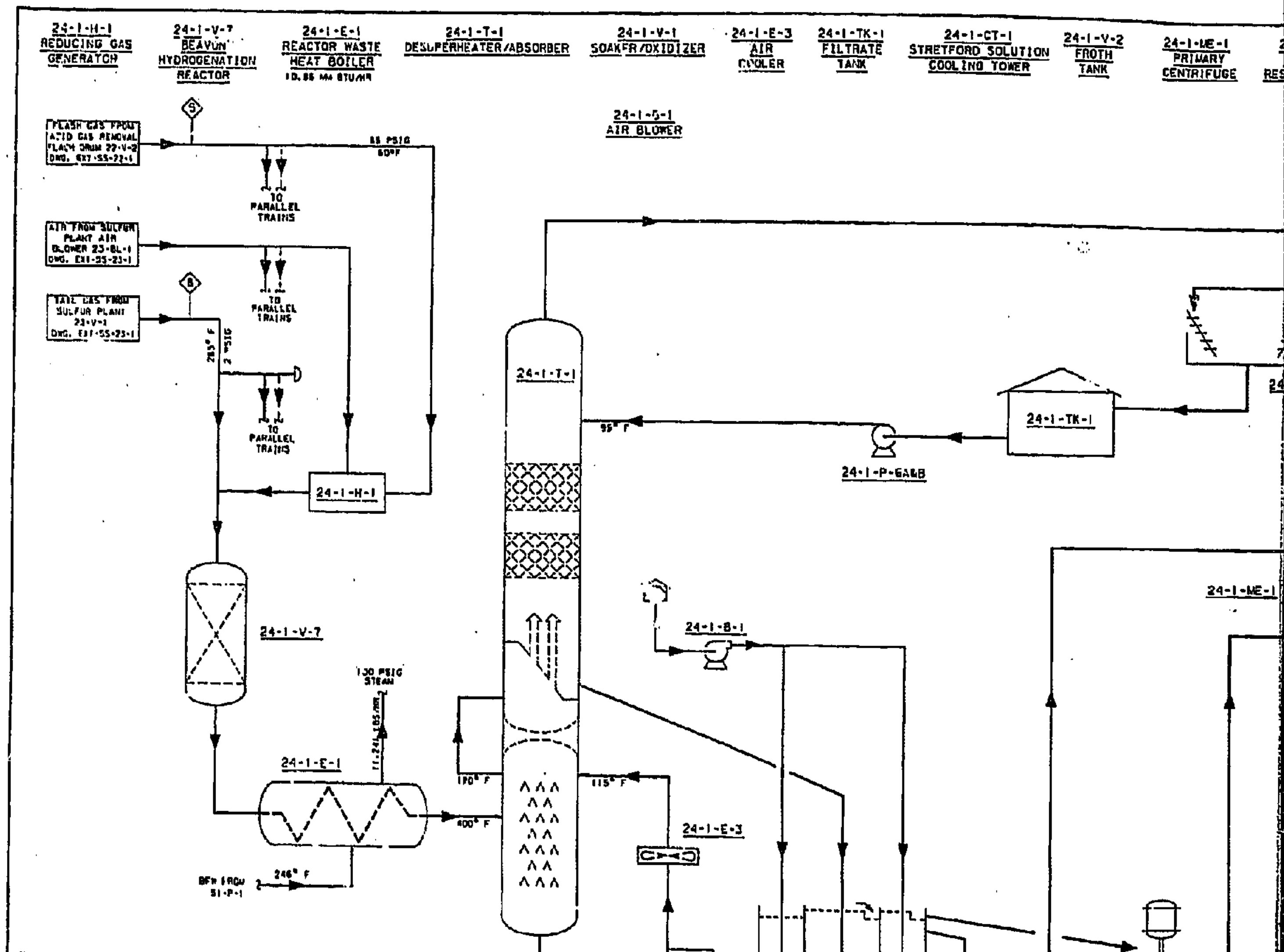
The reacted Stretford solution flows to soaker/oxidizer 24-1-V-1, where the reduced vanadate ($Na_2V_4O_9$) is oxidized to its original form by anthraquinone disulfonic acid (ADA) in the solution. The reduced ADA is subsequently regenerated by air sparged into the tank by blower 24-1-BL-1. The air also 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 through an ejector mixer 24-1-EJ-1, where sulfur is melted by direct injection of 100 psig steam. Molten sulfur (2,451 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.

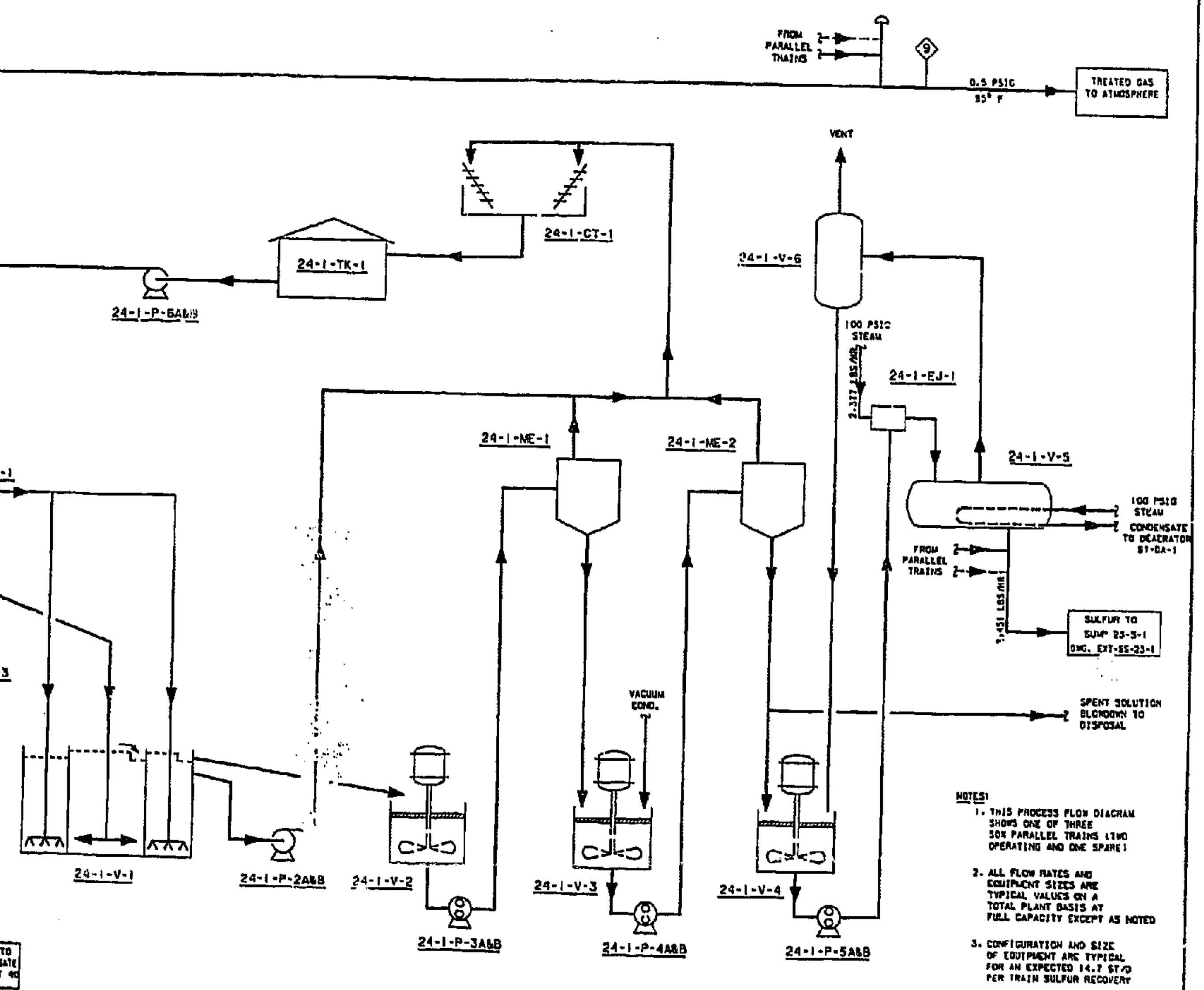
Equipment Notes

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



COMPONENT	MATERIAL BALANCE — PROCESS STREAM NO.					
	⟨5⟩ FLASH GAS		⟨8⟩ TAIL GAS		⟨9⟩ VENT GAS FM. T.G. TREATING UNIT	
	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	1.7	0.38	0.4	0.01	2.1	0.05
H_2	45.5	10.03	1.7	0.04	2.8	0.06
CO	129.6	28.71	10.9	0.24	0.3	0.01
CO_2	254.1	56.28	1,643.4	36.21	2,041.5	46.28
H_2S	17.1	3.79	32.9	0.72	0.3*	0.01
COS	1.5	0.33	2.1	0.05	0.0	0.00
N_2	1.4	0.31	1,691.8	37.28	1,931.1	43.78
Ar	0.2	0.04	21.8	0.48	25.1	0.57
NH_3	0.0	0.00	0.1	0.00	0.0	0.00
H_2O VAPOR	0.6	0.13	1,115.4	24.58	407.4	9.24
SO_2	0.0	0.00	16.5	0.36	0.0	0.00
S_6	0.0	0.00	0.1	0.00	0.0	0.00
S_8	0.0	0.00	0.5	0.01	0.0	0.00
VAPOR MPH	451.6	100.00	4,537.4	100.00	4,410.4	100.00
TOTAL LBS/HR	15,663		[illegible]		152,333.9	
MOL. WT. VAPOR	34.68		31.42		34.50	

24-1-E-3 AIR COOLER	24-1-TK-1 FILTRATE TANK	2[illegible]-1-CT-1 STRET[illegible]D SOLUTION COOLING TOWER	24-1-V-2 FROTH TANK	24-1-ME-1 PRIMARY CENTRIFUGE	24-1-V-3 PRIMARY RESLURRY TANK	24-1-ME-2 SECONDARY CENTRIFUGE	24-1-V-4 SECONDARY RESLURRY TANK	24-1-V-6 FLASH DRUM	24-1-EJ-1 SULFUR MELTER	24-1-V-5 SULFUR SEPARATOR



NOTES:

1. THIS PROCESS FLOW DIAGRAM SHOWS ONE OF THREE 50% PARALLEL TRAINS (TWO OPERATING AND ONE SPARE)
2. ALL FLOW RATES AND EQUIPMENT SIZES ARE TYPICAL VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY EXCEPT AS NOTED
3. CONFIGURATION AND SIZE OF EQUIPMENT ARE TYPICAL FOR AN EXPECTED 14.7 ST/D PER TRAIN SULFUR RECOVERY

FLUOR
IRVINE, CALIFORNIA

D. CLAVEAU
W.G.O'BELMITO
W.G.O'BELMITO
A. RAO
S.C. SMELSER
EPRI

PROCESS FLOW DIAGRAM
BEAVON/STRETFORD UNIT (TYPICAL)
TEXACO PROCESS-OXYGEN BLOWN
ALL CASES EXCEPT SS5, SH4 & SH5

EPRI
PALO ALTO, CALIFORNIA

NONE | 448334-EXT-SS-24-1 | 04

483SK1210

STEAM, BOILER FEEDWATER, AND CONDENSATE

Process Flow Diagrams EXT-SS-30-1 and EXT-SH-30-1 schematically represent the steam, boiler feedwater, and condensate systems for the base case plants. Since the diagrams encompass all parallel trains of equipment within the plant, reference to the train number which normally appears directly after the unit number has been omitted.

Entire generation of steam is accomplished in process plant by sensible and latent heat recovery and operates at three levels for each case:

High-Pressure (HP)	- 1450 psig, 900°F in EXT-SS case and 1000°F in EXT-SH case, at the 51-T-1A turbine inlet
Intermediate-Pressure (IP)	- 385 psig, 900°F in EXT-SS case and 1000°F in EXT-SH case, at the 51-T-1B turbine inlet
Medium-Pressure (MP)	- 100 psig in both cases for process users

EXT-SS

High-pressure steam generation is carried out in the gasifier waste heat boilers 20-E-1 and 20-E-2. Superheating of this high-pressure steam to 900°F occurs in the convection and radiation sections of the fired heater 51-FH-1. All of the superheated high-pressure steam is used to drive back-pressure power turbine 51-T-1A, which exhausts at 445 psig.

Saturated intermediate-pressure steam obtained from the sulfur plant waste heat boilers 23-E-1 is combined with high-pressure turbine exhaust steam. The final steam mixture at 619°F is reheated to 900°F in the convection and radiation sections of the fired heater 51-FH-1. This reheated steam is then sent to back-pressure turbine 51-T-1B, which exhausts at 115 psig.

Part of turbine 51-I-1B exhaust is desuperheated in desuperheater 51-DS-1 and exhausted to the medium-pressure header at 100 psig. Other sources of medium-pressure steam are the three steam generators in the sulfur plant and tail gas treating units. The medium-pressure steam is consumed in sulfur melter 24-EJ-1, the Selexol reboiler 22-E-4, in the deaerator to maintain the deaerator water in a saturated condition at 14 psig, and in other miscellaneous plant equipment. The remainder of steam exhausted from turbine 51-T-1B is used to drive the medium-pressure power turbine 51-T-2 and the high-pressure boiler feedwater pump driver turbine 51-T-3. Each of these two turbines are condensing machines exhausting at 2-1/2 inches Hg absolute (108.7°F). The main surface condenser 51-E-1 accepts cooling water at 80°F and discharges it at 100°F.

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 1505 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 condenser 51-E-1.

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. Regeneration of the polishing unit resin is accomplished in three separate vessels. Polished water at 109°F is then heated to 225°F, in the condensate heaters 21-E-3 before entering deaerator 51-DA-1. Also entering the deaerator are the condensate streams from medium-pressure steam users. The deaerator, providing 10-minute storage, is a horizontal tray unit operating at 14 psig.

Boiler feedwater for steam generation is supplied at two pressures: High-pressure, by the steam-turbine-driven pump 51-P-2A (the spare 51-P-2B is motor driven); and medium-pressure, by the motor-driven 51-P-1A&B. Medium-pressure boiler feedwater pump 51-P-1 provides the relatively small amount of feedwater needed in the steam generators 23-E-2, 23-E-3, 24-E-1 and the desuperheater 51-DS-1.

High-pressure boiler feedwater is first heated in gas cooling boiler feedwater heaters 21-E-1 to 290°F. The boiler feedwater is heated in the convection section of the fired heater 51-FH-1 to 349°F. Part of this water is "let down" and fed to the sulfur plant waste heat boilers 23-E-1 and remainder of this water is further heated to 598°F in raw gas cooling boiler feedwater heaters 20-E-3, before it enters the gasifier waste heat boilers 20-E-1.

EXT-SH

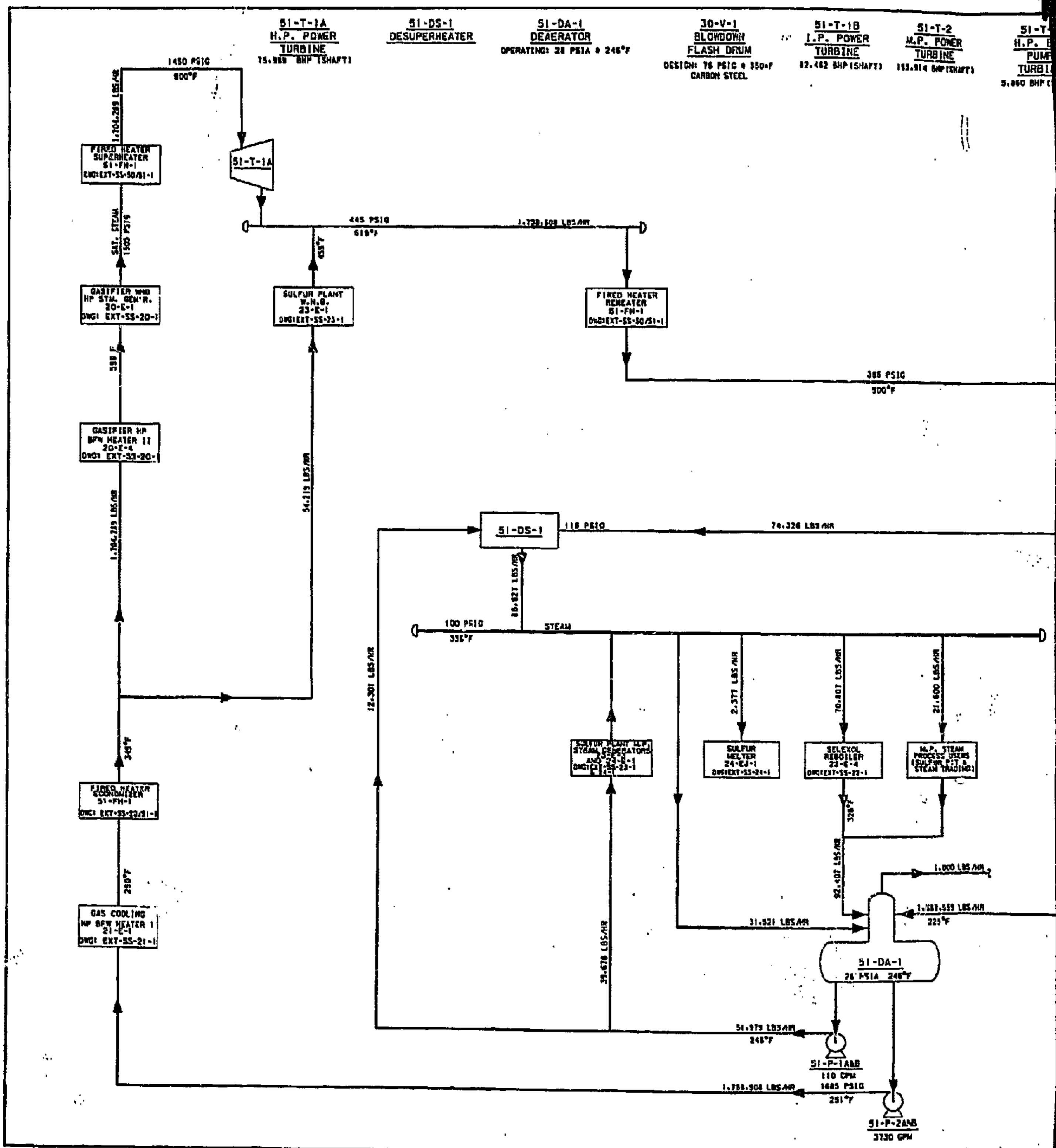
Both high-pressure steam generation and superheating (to 1000°F) are carried out in the gasifier waste heat boiler and superheater 20-E-1A and 20-E-2B. All of the superheated high-pressure steam is used to drive back-pressure power turbine 51-T-1A, which exhausts at 445 psig.

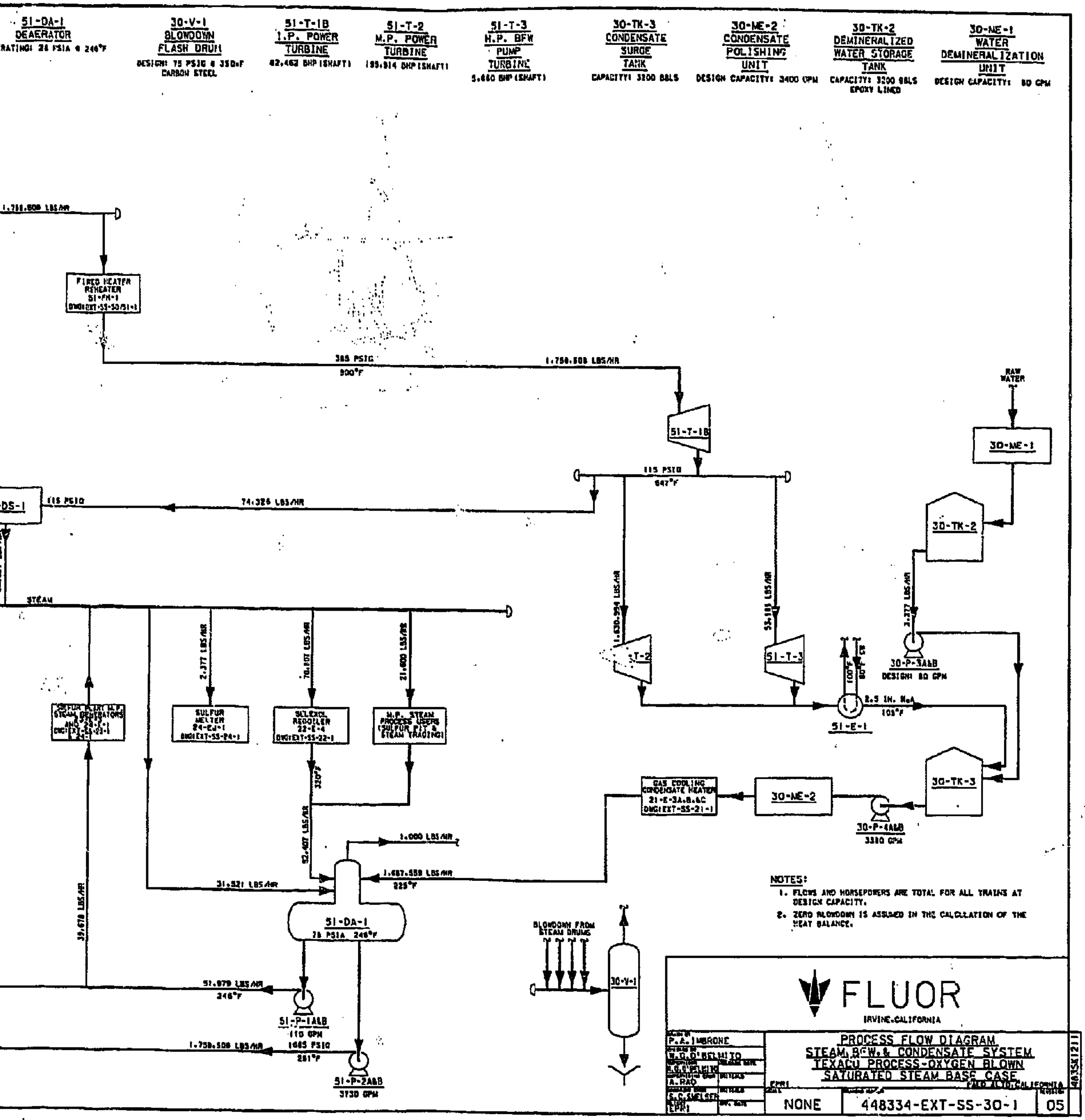
Saturated intermediate-pressure steam obtained from the sulfur plant waste heat boilers 23-1-1 is combined with high-pressure turbine exhaust steam. The final steam mixture at 707°F is reheated in the gasifier waste heat reheater 20-E-2A to 1000°F. The reheated steam is then sent to back-pressure turbine 51-T-1B, which exhausts at 115 psig.

Part of turbine 51-T-1B exhaust is desuperheated and exhausted to the medium-pressure steam header at 100 psig and the remainder is used to drive power turbine 51-T-2 and high-pressure boiler feedwater pump driver turbine 51-T-3, just as in the EXT-SS case. The other sources and users of the medium-pressure steam are also the same as those in the EXT-SS case.

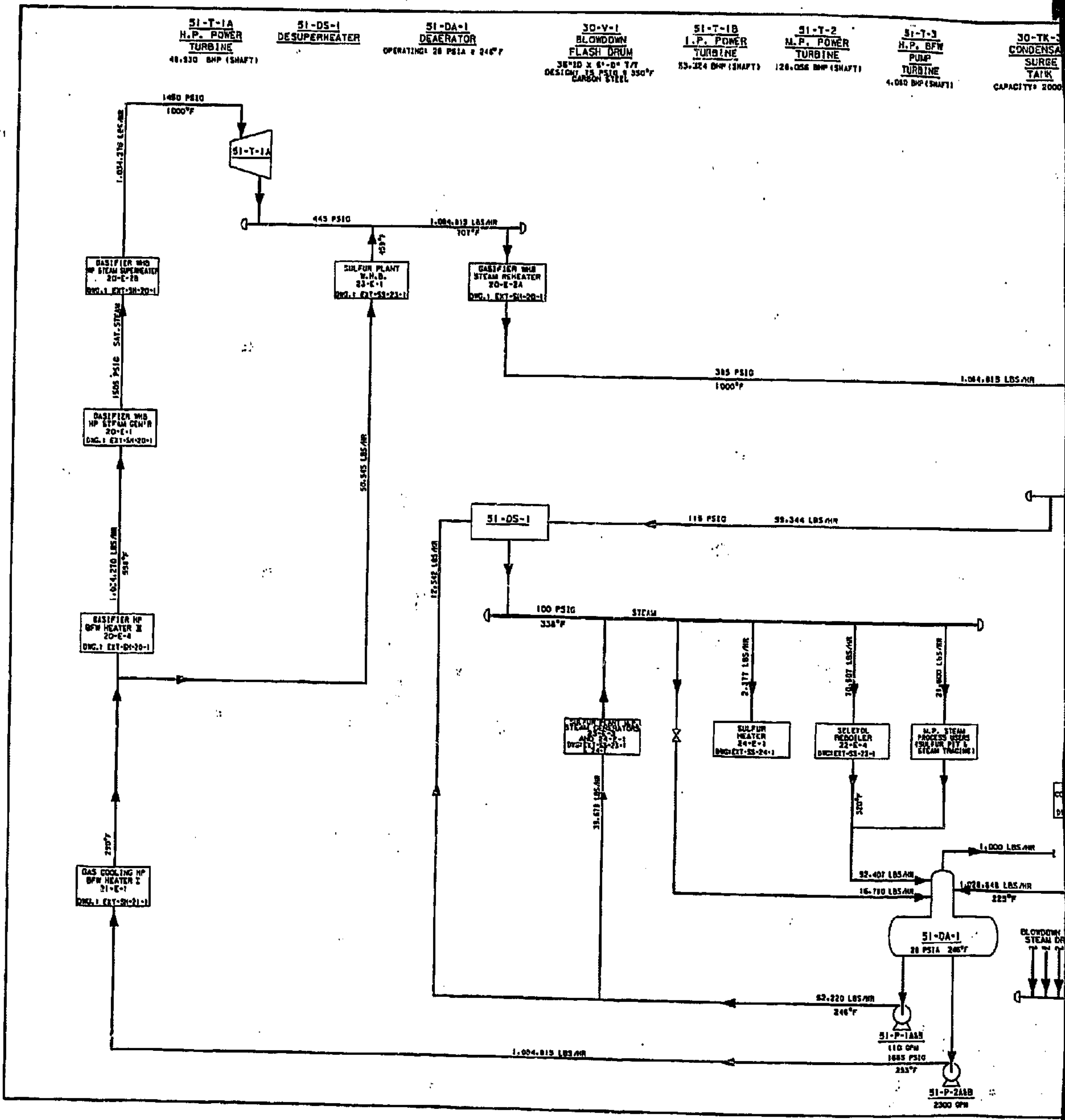
Raw water treatment, condensate polishing and reheating, and boiler feedwater supply are similarly accomplished as in the EXT-SS case.

The high-pressure boiler feedwater after being heated to 290°F in gas cooling boiler feedwater heaters 21-E-1 is split and directly sent to the sulfur plant waste heat boilers 23-E-1, after "letting down" and raw gas cooling boiler feedwater heaters 20-E-3.


51-T-1A
H.P. POWER
TURBINE
51-DS-1
DESUPERHEATER
51-DA-1
DEAERATOR
30-V-1
BLOWDOWN
FLASH DRUM
CARBON STEEL
51-T-1B
I.P. POWER
TURBINE
51-T-2
M.P. POWER
TURBINE
FIRED HEATER SUPERHEATER 51-FH-1
GASIFIER WHB HP STM. GEN'R. 20-E-1
GASIFIER HP BFW HEATER II 20-E-4
FIRED HEATER ECONOMIZER 51-FH-1
GAS COOLING HP BFW HEATER 1 21-E-1
SULFUR PLANT W.H.B. 23-E-1
FIRED HEATER REHEATER 51-FH-1
1450 PSIG
445 PSIG
385 PSIG
100 PSIG
STEAM
SULFUR MELTER
SELEXOL REBOILER
51-P-1A&B
51-P-2A&B
3730 GPM



51-DA-1 DEAERATOR
30-V-1 BLOWDOWN FLASH DRUM
DESIGN: 75 PSIG & 350°F CARBON STEEL
51-T-1B I.P. POWER TURBINE
42,462 BHP (SHAFT)
51-T-2 M.P. POWER TURBINE
51-T-3 H.P. BFW PUMP TURBINE
5,660 BHP (SHAFT)
30-TK-3 CONDENSATE SURGE TANK
CAPACITY: 3200 BBLS
30-ME-2 CONDENSATE POLISHING UNIT
DESIGN CAPACITY: 3400 GPM
30-TK-2 DEMINERALIZED WATER STORAGE TANK
CAPACITY: 3200 BBLS EPOXY LINED
30-ME-1 WATER DEMINERALIZATION UNIT
DESIGN CAPACITY: 80 GPM
FIRED HEATER REHEATER 51-FH-1
385 PSIG
900°F
1,758,508 LBS/HR
RAW WATER
51-T-1B
30-ME-1
115 PSIG
647°F
74,326 LBS/HR
30-TK-2
STEAM
2,377 LBS/HR
SULFUR MELTER 24-EJ-1
SELEXOL REBOILER 22-E-4
M.P. STEAM PROCESS USERS (SULFUR PIT & STEAM TRACING)
30-P-3A&B
DESIGN: 80 GPM
51-E-1
101°F
30-TK-3
GAS COOLING CONDENSATE HEATER 21-E-3A,B,&C
30-ME-2
30-P-4A&B
3380 GPM
1,000 LBS/HR
31,521 LBS/HR
225°F
51-DA-1
25 PSIA 240°F
BLOWDOWN FROM STEAM DRUMS
30-V-1
51-P-1A&B
110 GPM
1685 PSIG
261°F
51-P-2A&B
3730 GPM
NOTES:
1. FLOWS AND HORSEPOWERS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. ZERO BLOWDOWN IS ASSUMED IN THE CALCULATION OF THE HEAT BALANCE.
FLUOR
IRVINE, CALIFORNIA
P.A. IMBRONE
PROCESS FLOW DIAGRAM
STEAM, BFW, & CONDENSATE SYSTEM
TEXACO PROCESS-OXYGEN BLOWN
SATURATED STEAM BASE CASE
EPRI
PALO ALTO, CALIFORNIA
NONE
448334-EXT-SS-30-1
05


51-T-1A
H.P. POWER
TURBINE
51-DS-1
DESUPERHEATER
51-DA-1
DEAERATOR
30-V-1
BLOWDOWN
FLASH DRUM
CARBON STEEL
51-T-1B
I.P. POWER
TURBINE
51-T-2
M.P. POWER
TURBINE
51-T-3
H.P. BFW
PUMP
TURBINE
1450 PSIG
1000°F
51-T-1A
445 PSIG
GASIFIER WHB
HP STEAM SUPERHEATER
20-E-2B
SULFUR PLANT
W.H.B.
23-E-1
GASIFIER WHB
STEAM REHEATER
20-E-2A
SAT. STEAM
385 PSIG
1000°F
GASIFIER WHB
HP STEAM GEN'R
20-E-1
51-DS-1
100 PSIG
STEAM
GASIFIER HP
BFW HEATER II
20-E-4
SULFUR
HEATER
24-E-1
SELEXOL
REBOILER
22-E-4
M.P. STEAM
PROCESS USERS
GAS COOLING HP
BFW HEATER I
21-E-1
51-DA-1
51-P-1A&B
51-P-2A&B

30-V-1
BLOWDOWN
FLASH DRUM
36"ID X 6'-0" T/T
DESIGN: 75 PSIG & 350°F
CARBON STEEL

51-T-1B
I.P. POWER
TURBINE
53,324 BHP (SHAFT)

51-T-2
M.P. POWER
TURBINE
126,056 BHP (SHAFT)

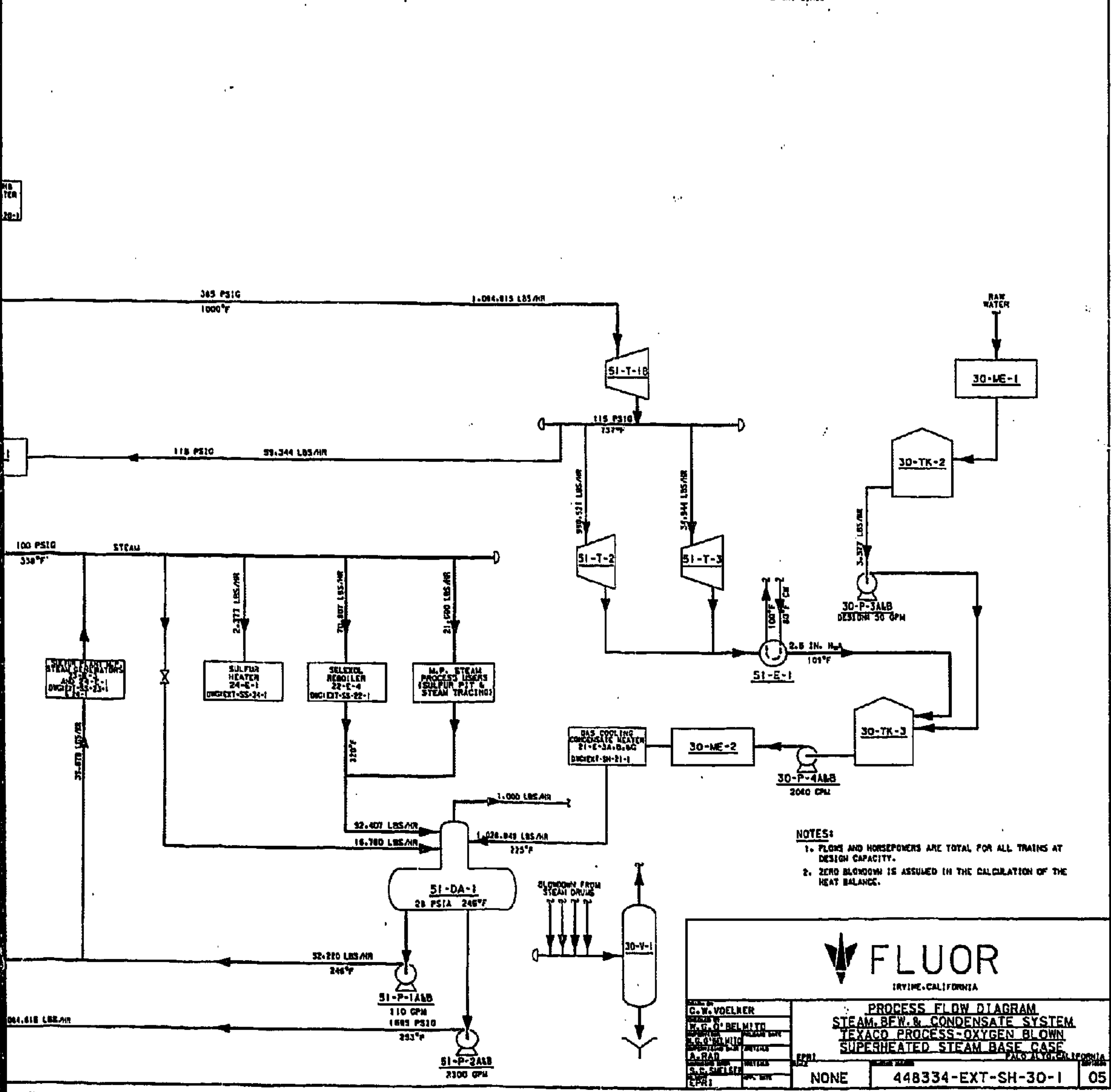
51-T-3
H.P. BFW
PUMP
TURBINE
4,050 BHP (SHAFT)

30-TK-3
CONDENSATE
SURGE
TANK
CAPACITY: 2000 BBLS

30-ME-2
CONDENSATE
POLISHING
UNIT
DESIGN CAPACITY: 2100 GPM

30-TK-2
DEMINERALIZED
WATER STORAGE
TANK
CAPACITY: 2000 BBLS
EPOXY LINED

30-ME-1
WATER
DEMINERALIZATION
UNIT
DESIGN CAPACITY: 50 GPM



NOTES:

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

FLUOR
IRVINE, CALIFORNIA

C.W. VOELKER	PROCESS FLOW DIAGRAM
W.G. O'BELMITO	STEAM, BFW. & CONDENSATE SYSTEM
W.G. O'BELMITO	TEXACO PROCESS-OXYGEN BLOWN
A. RAD	SUPERHEATED STEAM BASE CASE
S.C. SMELSER	EPRI PALO ALTO, CALIFORNIA
EPRI	NONE | 448334-EXT-SH-30-1 | 05

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