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

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Prepared by
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Irvine, California

ELECTRIC POWER RESEARCH INSTITUTE

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

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ABSTRACT

This report presents the results of economic screening studies for several current and advanced coal gasification processes coupled with combined cycle power generation. The objective of these studies was to identify whether significant economic and/or environmental incentives exist for using such systems compared to the current practice of direct coal firing and stack gas cleanup.

The processes studied included the Lurgi dry ash gasifier, the British Gas Corporation Slagger, and three entrained bed processes offered by Combustion Engineering, Foster Wheeler, and Texaco. All these processes were integrated with combined cycle plants based on advanced gas turbine technology (2,400°F Combustion Outlet) estimated by Westinghouse to be available in the 1981-1985 time period.

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

The conclusions reached in the report are that several of the processes considered are potentially attractive and are, or can be, available for commercialization during the next decade. In particular, the entrained bed processes appear to offer substantial environmental as well as economic advantages owing to their simplicity and lack of by-products.

It is concluded that development emphasis should be placed on compression, power generation, and heat transfer equipment, rather than on more gasification processes.

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SUMMARY

This study was performed to identify whether significant economic and/or environmental incentives exist for using various current and advanced gasification processes coupled with gas turbine combined cycle power plants to generate electricity. The case studies were all performed using advanced gas turbine designs based on a 2400°F combustor outlet temperature. These designs were provided by Westinghouse. These turbines are not currently available, but with development, are expected to be available in the 1981 to 1985 time period (Appendix A).

The processes studied were as follows:

a. Lurgi dry ash gasifier (MACW)
b. British Gas Corporation slagging gasifier (MXSC)
c. Foster Wheeler entrained bed gasifier (EXHC & EAHC)
d. Combustion Engineering atmospheric pressure process (EALC)
e. Texaco entrained bed process (EXTC)

The letter codes for each case are explained in Appendix D.

The plant sizes were selected to match current utility practice of building plants in the 1000-1500 megawatt (MW) capacity range. These plants all feed a constant coal rate equivalent to 10,000 tons/day of Illinois No. 6, and produce in the range of 1000 to 1200 megawatts of power (Table S-1). All cases use Illinois No. 6 coal except case MACW which uses a New Mexico coal.

To help set the results in perspective, a comparison case has been included which is based on a 1000 MW coal fired boiler plant with a stack gas scrubber. Capital costs and plant operating characteristics for this case are based on a report by the Bechtel Corporation[1]. Since the cost estimate for this plant was prepared outside Fluor, it may not be entirely consistent with the other Fluor estimates. However, all capital cost allowances (i.e., construction loan interest, start-up costs, working capital) as well as plant operating costs have been generated on an equivalent and consistent basis to those developed for the gasification -

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1. "Coal Fired Power Plant Capital Cost Estimates," EPRI AF-342, Project SOA 76-329, January 1977, prepared by Bechtel Power Corp., San Francisco, California.

combined cycle plants. Nevertheless, caution should be exercised when making comparisons between this case and other cases.

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

The major conclusions of this study are as follows:

- Most of the gasification combined cycle systems studied have the potential for being available for commercialization in the mid to late 1980's. Some of these technologies appear to be significantly more efficient (lower cost of services) than current coal fired boiler technology.

- Some of these processes offer substantial environmental advantages because of simplicity and high temperature operation (EXTC, EAHC, EXHC and EALC). This statement assumes that the two stage entrained systems (EAHC, EXHC and EALC) will not produce any liquid hydrocarbon by-products.

- The cost of the gasification section of these plants is, by itself, rarely a major portion of the total plant cost, suggesting that added developmental funds should go not only to gasifier improvement, but to other areas such as gas turbines, compressors, and high temperature, dirty service heat transfer devices.

- There appears no reason to favor oxygen blown versus air blown gasification for power production. In fact, the reverse may be true.

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

- Use data provided by process developers.
- Produce no net products except electricity, sulfur and ammonia.

- Meet current environmental restrictions for an Illinois plant location (1.2 lb SO_2/MM Btu of coal fired).
- Provide all facilities required to permit stand-alone operation of a grass roots plant.

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

- Mid-1976 dollars with no escalation.
- Thirty-six month construction period for gasification - combined cycle plants.
- Forty-eight month construction period for coal fired boilers with stack gas scrubbers.
- Eight percent construction loan interest, compounded over the plant construction schedule.
- Coal cost of $1.00/MM Btu and $2.00/MM Btu.
- Seventy percent operating load factor.
- Twenty-five year plant life.
- Fifty:fifty debt:equity ratio.
- Eight percent annual bond interest.
- Twelve percent annual return on equity after taxes.

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

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

Some comments as to the present suitability of the various processes for commercialization seem in order. The comments given here on the development status of each of the technologies are not comprehensive and are not intended to detract from any of the technologies studied. The purpose is only to provide a perspective from which the conclusions presented in this report may be viewed.

The Lurgi dry ash gasifier is now, and has been for some time, in commercial use worldwide for gasifying noncaking coals. The gasification step can be considered commercialized. The coupling of this process with electricity generation, especially in a load following mode, has not been well demonstrated as of this writing. A 170 MW plant in Lunen, West Germany has been built and operated. This plant has a different power cycle than that considered for these studies, i.e., a supercharged boiler followed by an expansion turbine. STEAG now reports the plant has proved fully functional[2].

A gasification - combined cycle test facility of 25 MW capacity is under construction at Pekin, Illinois (Commonwealth Research Corp.). This plant is scheduled for completion in 1980. It will use the standard U.S. combined cycle, featuring a gas turbine followed by a steam boiler or HRSG(as in the studies reported here). In the meantime dynamic simulation studies are being carried out by General Electric under EPRI sponsorship (RP-914) to identify control strategies for this type of plant.

The Texaco gasification process has been commercially used with petroleum residues for many years in synthesis gas applications. The process has been demonstrated on a pilot scale with coal by Texaco Development Corp. A demonstration scale plant using coal has been built for synthesis gas production in West Germany for RuhrChemie. This unit is scheduled for a first quarter 1978 startup. Another demonstration scale unit, for ammonia synthesis gas, has recently been announced for TVA. Thus, while the Texaco process has been commercially demonstrated on

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2. Combined Gas/Steam Turbine Power Stations with Coal Pressure Gasification Unit operating to the STEAG-Lurgi System, H. Meyer - Kahrweg, STEAG, Essen, West Germany, 55th Annual GPA Convention, March 22-24, 1976.

petroleum, it has not yet been demonstrated on a large scale with coal, nor has it yet been demonstrated in combination with a combined cycle power plant.

A key feature of this process is a high temperature heat exchanger used to raise steam from hot gasifier effluent. While not an "off the shelf" item, successful designs for such units have been developed by two West German firms, Steinmuller and Siegener. Several of these units have seen extended commercial service. Gasification in general would profit greatly by further developments in this type of equipment.

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

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

The BGC slagging gasifier has been demonstrated in fairly large scale equipment by the British Gas Corporation at the Westfield Development Center. This technology could probably be ready in nearly the same time frame as the Texaco process. It is, however, based on equipment which is somewhat more complex than the Texaco gasifier. Although it has the further disadvantage of requiring separation and handling of liquid by-products and fines, it appears from these studies to offer the potential for extremely efficient and low cost power production when integrated with combined cycle equipment.

The atmospheric pressure entrained, two stage gasifier represented by the Combustion Engineering design (Case EALC) is currently being piloted at the 120 ton/day scale at Windsor, Connecticut. Babcock and Wilcox built and successfully operated a number of atmospheric and higher pressure, two stage, entrained devices in the 1950's and 1960's ranging in capacity from 1/4 ton coal/hr. to 20 tons coal/hr. Only one of these devices - the Alliance unit (2-1/2 tons coal/hr.) - was air blown. Although B & W reported the absence of tars in the crude gas, this unit was generally operated at higher temperatures than the 1700°F supplied by Combustion Engineering for this study. Based on the results of this study, development either in compression equipment for fuel gas, or development of low pressure ratio gas power turbines would be needed to make the process attractive for combined cycle power production. It must be kept in mind, however, that no attempt was made in this study to optimize the power cycle to match constraints imposed by the particular gasification process being considered. It is entirely conceivable, therefore, that interaction of the Combustion Engineering gasifier with a different configured power system would result in a cost of electricity that is competitive with those costs generated for the other cases studied under this contract.

The high pressure, two-stage entrained gasifiers represented by the Foster Wheeler cases (EAHC and EXHC) are in a much earlier stage of development than any of the above system. A proposal has been submitted by Foster Wheeler to the Department of Energy for the construction of a process development unit, but no commitment has yet been made. Small scale laboratory development work for this type of gasifier was conducted by Bituminous Coal Research. This work was, however, limited to an investigation of the devolatilization or upper stage only. The BIGAS pilot plant at Homer City employs a concept similar to that proposed by Foster Wheeler. Major differences between the BIGAS reactor and the Foster Wheeler gasifier are: higher pressure operation, different coal feed system and the fact that the BIGAS reactor will only be oxygen blown. The results of this study, however, indicate a clear incentive to develop the air blown Foster Wheeler gasifier for electric power generation. Therefore, unless development efforts are accelerated, it is difficult to predict commercialization of this concept before 1990.

TABLE S-1

SUMMARY OF OPERATING RESULTS

	LURGI MACW	BGC SLAGGER MXSC	FOSTER WHEELER EAHC	FOSTER WHEELER EXHC	COMBUSTION ENGINEERING EALC	TEXACO EXTC (Slurry Feed)	TEXACO EXTC (Dry Feed)
Gasification and Gas Cleaning System							
Coal Feed Rate, Lb/Hr m.f.	1,014,814	798,333	798,333	798,333	798,333	798,333	798,333
Oxygen or Air/Coal Ratio Lb/Lb m.f.(1)	1.562	0.481	2.857	0.609	4.37	0.858	0.806
Oxidant Temperature, °F	340	214	800	335	437	300	300
Steam/Coal Ratio Lb/Lb m.f.	0.758	0.31	0.150	0.624	0	0	0.610
Slurry Water/Coal Ratio Lb/Lb m.f.	NA	NA	NA	NA	NA	0.503	NA
Gasifier Exit Pressure, psig	340	320	360	360	(5)	600 (4)	600 (4)
Crude Gas Temperature, °F	861	820	1,700	1,700	1,700	2,300-2,600	2,300-2,600
Crude Gas HHV (Dry Basis), Btu/SCF (2)	189.1	379.0	174.1	315.4	113.0	281.1	280.7
Temperature of Fuel Gas to Gas Turbine, °F	425	580	800	800	1,200	781	781
Power System							
Gas Turbine Inlet Temperature, °F	2,400	2,400	2,400	2,400	2,400	2,400	2,400
Pressure Ratio	17:1	17:1	17:1	17:1	17:1	17:1	17:1
Gas Turbine Exhaust Temperature, °F	1,137	1,128	1,127	1,133	1,147	1,140	1,140
Steam Conditions, psig/°F/°F	1450/900/1000	1450/900/1000	1450/900/1000	1450/900/1000	1450/900/1000	1450/900/1000	1450/900/1000
Condensing Pressure, Inches Hg abs.	2.5	2.5	2.5	2.5	2.5	2.5	2.5
Stack Temperature, °F	275	275	275	275	275	272	272
Gas Turbine Power, MW (3)	590	857	751	803	886	745	763
Steam Turbine Power, MW (3)	430	385	504	384	307	448	425
Power Consumed, MW	32	30	42	38	55	36	46
Net System Power, MW	988	1,212	1,213	1,149	1,138	1,157	1,142
Overall System							
Process and Deaerater Makeup Water, gpm/1000 MW	2,207	834	497	1,031	157	362	1,072
Cooling Tower Makeup Water, gpm/1000 MW	5,698	5,882	6,125	6,003	7,439	7,588	7,255
Cooling Water Circulation Rate, gpm/MW	366	307	341	321	343	347	352
Cooling Tower Heat Rejection, % of Coal HHV	33.9	33.8	36.8	33.2	37.6	38.7	35.6
Air Cooler Heat Rejection, % of Coal HHV	6.5	4.7	3.2	7.2	4.9	5.2	4.6
Net Heat Rate, Btu/kWh	9,762	8,410	8,428	8,876	8,959	8,813	8,928
Overall System Efficiency (Coal→Power), % of Coal HHV	34.96	40.6	40.5	38.5	38.1	38.7	38.2

NOTES: (1) Dry basis, 100% O_2 for oxygen blown case

(2) Excluding the HHV of H_2S, COS and NH_3

(3) At generator terminals

(4) Average gasifier pressure

(5) Gasifier exit pressure is -0.5 inch H_2O

TABLE S-2

SUMMARY OF ECONOMIC RESULTS

PRODUCTION AT DESIGN CAPACITY	COAL FIRED POWER PLANT WITH FLUE GAS DESULFURIZATION	LURGI MACW	BGC SLAGGER MXSC	FOSTER WHEELER EAHC	FOSTER WHEELER EXHC	COMBUSTION ENGINEERING EALC	TEXACO EXTC (SLURRY FEED)	TEXACO EXTC (DRY FEED)
Net Power, MW(1)	1,000	988	1,212	1,213	1,149	1,138	1,157	1,142
Overall Plant Heat Rate, Btu/kWh	9,928	9,762	8,410	8,428	8,876	8,959	8,813	8,928
TOTAL CAPITAL(2)								
Total Capital @ $1/MM Btu Coal, $/kW	838	906	711	705	739	931	816	854
Total Capital @ $2/MM Btu Coal, $/kW	855	922	725	719	753	946	831	869
AVERAGE COSTS OF SERVICES(2)								
Annual Cost @ $1/MM Btu Coal, $1000/yr	250,690	249,573	243,474	241,328	239,851	288,563	262,088	267,861
Per Unit @ $1/MM Btu Coal, mills/kWh	40.88	41.20	32.79	32.53	34.05	41.35	37.21	38.25
Annual Cost @ $2/MM Btu Coal, $1000/yr	314,173	311,235	308,666	306,520	305,043	353,756	327,280	333,062
Per Unit @ $2/MM Btu Coal, mills/kWh	51.23	51.38	41.57	41.32	43.30	50.69	46.47	47.56

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

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INTRODUCTION AND SCOPE

The studies reported here represent a continuation and extension of earlier economic studies[1] done for EPRI by Fluor. The object of the new work was to investigate the economics associated with producing electricity from coal in new integrated gasification-combined cycle power plants. These plants are based on gasification processes using both current and advanced technologies, integrated with advanced gas turbine (2400°F combustor outlet) combined cycle power plants. A comparison is made of these cases with a coal fired boiler power plant using stack gas desulfurization.

The design of a fully integrated complex which exports electric power as its principal product is a substantially different undertaking from the design of a plant to produce fuel gas for pipeline transport. A new and different set of opportunities for using and reusing heat energy exists. It was envisioned at the outset of this study that this set of circumstances would be synergistic to a degree and would help overall plant efficiency. This, in fact, did happen. The reader will see that all the gasification cases considered are more efficient than the direct coal fired boiler case.

Designs for each of the gasification units, and for some of the Selexol® acid gas removal units, were based on information provided by appropriate licensors. The power systems were designed by Westinghouse. Plant costs were estimated by Fluor. Economic evaluation criteria were supplied by EPRI. For reference, a "benchmark" case is included which is based on a direct coal fired boiler power plant with stack gas scrubbing[2]. The reader should note that the cost estimate for this plant, while in the same range as those for the combined cycle plants, was not prepared by Fluor and, therefore, is probably not entirely consistent with the other estimates. Further, this plant is for a nominal 1000 MW capacity, whereas the gasification plants are all somewhat larger.

Cases reported here are referred to with a four letter acronym. These are defined in Appendix D. Block flow diagrams are given for each case to indicate the

1. "Economics of Current and Advanced Gasification Processes for Fuel Gas Production," EPRI AF-244, Project 239, Final Report July 1976.

2. "Coal Fired Power Plant Capital Cost Estimates," Prepared by Bechtel Corporation for EPRI (EPRI AF-342, Project SOA 76-239, January 1977).

overall plant flow scheme. Flow sheets are also provided for individual process units within each plant where necessary to depict what is included that is specific for a given case.

TECHNICAL CRITERIA

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

Insofar as possible, EPRI requested gasifier material and heat balances and equipment requirements from organizations which developed or licensed gasification processes representative of the technologies being considered. These organizations are as follows (see also Appendix D):

- MACW: American Lurgi Corporation
- MXSC: British Gas Corporation
- EXHC and EAHC: Foster Wheeler Energy Corporation
- EXTC: Texaco Development Corporation
- EALC: C-E (Combustion Engineering Power Systems)
- Electric Generation Systems: Westinghouse Electric Corporation

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

Two different coals are used in these studies. Their analyses are given in Table C-1. The western coal was chosen for Case MACW because it was felt that the choice of this type of coal would be most beneficial for the Lurgi gasification system. The other cases all use Illinois Number 6. In all the cases, coal was assumed delivered to the site washed and sized. If experience were to demonstrate that this assumption was not reliable, then each of the cases presented here would require additional coal handling equipment. This would slightly affect overall plant costs, but would not alter the comparisons between cases.

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

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

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

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

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

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

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

TABLE C-1

COAL ANALYSES

Type	Illinois No. 6	New Mexico
PROXIMATE ANALYSIS (Wt %)		
Moisture	4.2	12.4
Ash	9.6	25.6
Fixed Carbon	52.0	)
		) 62.0
Volatile Matter	34.2	)
	100.0	100.0
ULTIMATE ANALYSIS - DAF COAL (Wt %)		
Carbon	77.26	76.61
Hydrogen	5.92	5.71
Oxygen	11.14	14.81
Nitrogen	1.39	1.35
Sulfur	4.29	1.47
Other	-	0.05
	100.00	100.00
HEATING VALUE - AS RECEIVED		
Higher Heating Value (HHV) (Btu/lb)	12,235	8,325
Net Heating Value (LHV) (Btu/lb)	11,709	-

TABLE C-2

SITE CONDITIONS

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

TABLE C-3

WATER ANALYSIS

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

ECONOMIC CRITERIA

A consistent set of criteria for estimating capital requirements and cost of services was supplied by EPRI. Criteria for gasification - combined cycle power plants are summarized in Tables C-4 and C-5. A separate set of criteria was supplied for a conventional coal fired power plant with flue gas desulfurization. These criteria are summarized in Tables C-6 and C-7.

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

Plant investment estimates contain a contingency. The contingency has been divided into two parts. First is a project contingency which is intended to cover additional equipment that would result from a more detailed design of a definitive project at an actual site. The second is a process contingency which is applied to unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. Historically, as a new technology develops from the conceptual stage to commercial reality, a variety of technical problems which were not considered during the early stages of the development emerge. Solution of these problems generally results in an increase in the cost of the technology due to the need for more expensive materials of construction, more complex equipment specifications and sometimes the need for additional processing equipment. A total plant process contingency is arrived at by applying a separate contingency to individual process units based on their state of development and accumulating the results. The process contingency allowances, shown as a percentage of the installed plant costs before any project or other process contingencies have been added, are listed in Table C-8.

TABLE C-4

CAPITAL INVESTMENT BASIS FOR GASIFICATION -
COMBINED CYCLE PLANTS

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

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

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

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

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

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

TABLE C-5

COST OF SERVICES BASIS FOR GASIFICATION -
COMBINED CYCLE PLANTS

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

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

ITEM	BASIS

Process Unit	Maintenance % of Total Plant Investment/Yr
Support Facilities	1.5
Combined Cycle	1.5

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

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

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

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

TABLE C-6

CAPITAL INVESTMENT BASIS FOR THE COAL FIRED BOILER WITH STACK GAS SCRUBBER

ITEM	BASIS
Total Plant Investment	- Mid-1976 dollars with no escalation. - Illinois location. - Clear and level site. - Plant investment estimates for this plant are to be taken directly from the Bechtel study, EPRI report AF-342 using the estimates for Plant No. 1. When extracting total plant investment costs from the Bechtel report (Pages 8-5) one item included by Bechtel was removed, i.e., switchyard bulk materials ($9.6 million). - 1,000 MW design capacity.
Total Plant Investment Definition	- The total plant investment is defined as the sum of: (a) Process (or onsite) plant investment costs (b) General facilities (or offsites) investment costs (c) Contingencies
Project Contingency	- This contingency factor is intended to cover additional equipment that would result from a more detailed design of a definitive project at an actual site. An allowance of 15% of the sum of the Process Plant Investment and the General Facilities cost was used.

ITEM	BASIS
Process Contingency	- This contingency factor is to be applied to unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. A process contingency of 0% is to be applied to all sections of this plant with the exception of the stack gas scrubbing section. A process contingency of 5% was applied to the stack gas scrubbers.
Total Capital Requirements	- The total capital requirement includes all capital necessary to complete the entire project. These items include: (a) Total Plant investment (b) Royalties (c) Preproduction costs (d) Construction loan interest (e) Initial chemical and catalyst charge (f) Working capital
Paid-up Royalties	- 0% of total plant investment.
Preproduction Costs	- One month variable operating costs excluding coal. Variable costs are sludge disposal and ash disposal utilities, and maintenance materials. - Two months fixed costs excluding income taxes. Fixed costs are operating and maintenance labor, administrative and support labor, general and administrative expense, and property taxes and insurance. - 5% of total plant investment (this charge allows for possible changes in

ITEM	BASIS
	process equipment and charges associated with depreciation, bond interest, and return on equity during the preproduction period). - 25% of full load coal costs for one month.
Construction Loan Interest	- 0.1931x Total Plant Investment (based on compounded 8%/year interest over the plant construction expenditure schedule).
Construction Expenditures	see table below

Year	Percent of Total Plant Investment
1	25
2	35
3	30
4	10

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

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

TABLE C-7

COST OF SERVICES BASIS FOR COAL FIRED BOILER
WITH STACK GAS SCRUBBER

ITEM	BASIS
Operating Load Factor	- 70%
Cost of Coal Delivered	- $1.00 MM/Btu and $2.00 MM/Btu
Cost of Sludge Disposal	- 2 mills/kWh
Cost of Ash Disposal	- $1.00/ton
Cost of Limestone Delivered	- $10.00/ton
Maintenance	- Annual maintenance costs for the entire power plant are estimated at 2 mills/kWh plus 4% of the total plant investment for the stack gas scrubbing system.
Maintenance Labor/Materials Ratio	- 40/60
Operating Labor	- 9 men per shift at $11 per hour manhour (this labor rate corresponds to a direct labor charge of $8/hour plus a 35% payroll burden).
Administrative and Support Labor	- 30% of operating and maintenance labor.
General and Administrative Expense	- 60% of operating and maintenance labor.
Property Taxes and Insurance	- 2.5%/yr. of total plant investment.
Cost of Capital	- The capital charges (income taxes, interest on debt, return on equity, and depreciation) are computed on a levelized basis with a 10% discount rate. The discount rate is based on the average cost of money. Using this

ITEM	BASIS
	basis, the capital charges will be 15.6% per year of the Total Capital Requirement. The investment factors that form the basis for the 15.6%/yr. capital charge are shown below:

Depreciation	Straight Line
Tax Life	25 years
Plant Life	25 years
Debt/Equity Ratio	50/50
Bond Interest	8% annually
Bond Life	25 years
Return on Equity after Taxes	12% annually
Income Tax Rate	52%
Escalation Rate	Not included
Investment Tax Credit	Not included

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

TABLE C-8

PROCESS CONTINGENCIES

CASE	PERCENT MACW	MXSC	EALC	EAHC	EXHC	EXTC
Coal Handling	0	0	5	5	5	0
Oxidant Feed	0	0	0	0	0	0
Gasification	5	15	35	50	45	15
Ash Handling	5	5	5	5	5	5
Gas Cooling	0	0	10	10	10	0-15(1)
Acid Gas Removal	5	5	5	0	0	0
Sulfur Recovery (Claus)	0	0	NA	0	0	0
Tail Gas Treating	15	15	NA	15	15	15
Process Condensate Treatment and Steam, Condensate and BFW	0-15(2)	0	0	0	0	0
Support Facilities	0	0	0	0	0	0
Fuel Gas Compression	NA	NA	0	NA	NA	NA
Combined Cycle	5	5	5	5	5	5

(1) 15% applied to waste heat boilers, 0% to remaining low temperature gas cooling equipment.

(2) 15% applied to byproducts boiler, 0% applied to all other equipment.

DISCUSSION OF RESULTS

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

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

As a baseline for comparison, a case has been included here for a coal fired boiler power plant with a stack gas scrubber. This case is based on data contained in an EPRI report[1]. The reader will note, on examining the data, that the thermal efficiency for this type of plant is the lowest of any of the cases presented; and the overall cost of services is as high as that for the most expensive of the gasification cases. This case has been included because it represents the conventional alternative to gasification based systems. The cost estimate for this case was prepared by another contractor, and it may not be entirely consistent with the Fluor estimates. The technology is better known for this type of plant, and since the estimate was based on an actual plant design, it may reflect a lesser degree of optimism than the gasification estimates. This

1. "Coal Fired Power Plant Capital Cost Estimates," EPRI AF-342, Project SOA 76-329, January 1977, Prepared by Bechtel Power Corporation, San Francisco, California.

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is partially reflected by the fact that no process contingency has been included for the boiler or the power generating equipment.

In choosing between gasification or direct coal firing for electricity generation, in the future, the utility plant owner must be able to realize a clear advantage for the route selected. We believe the gasification has such advantages. For some of the cases shown here (MXSC, EAHC, EXHC), the overall cost of services is significantly below that for coal firing. If it is true that the gasification plants have a potential for problems that are not fully known, it is also true they have untapped potential for improvement. As machinery and process development proceeds, such improvement is inevitable.

The stack gas scrubbing process used in the Bechtel plant is a "throwaway" process (limestone based). It requires the plant owner to handle large volumes of chemicals and solid waste to make the operation go. If a change in environmental restrictions forces the owner to change his handling of these chemicals and wastes, it could well force the plant to shut down. Additionally, it is not easy to modify such plants if the degree of sulfur removal is changed (EPA is considering such a change at this writing); whereas it is practical to modify gasification plants to meet practically any level of sulfur emission contemplated. Gasification plants emit essentially no particulates from the stack. Recent literature[2] further indicates that because of the low Btu gas combustion characteristics, NO_x emissions may be much less than for direct coal firing.

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

The reader's attention is directed to a subtle but important difference between this work and that presented in a previous EPRI report (AF-244) on fuel gas production.

The term "gasifier cold efficiency" has been eliminated from consideration here, as an essentially meaningless quantity.

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2. "Recent Experimental Results on Gasification Combustion of Low Btu Gas for Gas Turbines," Crouch, Schlinger, Klapatch, Vitti; Combustion, April 1974, pp 32-38.

For engineering studies of this type, much importance is placed on the thermal efficiency of major processing units such as the coal gasifier. Most published reports addressing the subject of gasification-combined cycle power generation, quote gasifier efficiencies and use these estimates to justify the fact that one type of gasifier appears to be more efficient in a power generating system than another. It is further noted, that the method for describing the thermal efficiency of a gasifier is via the "cold gas efficiency," i.e., the ratio of the product gas HHV to the coal feed HHV. This type of efficiency favors gasifiers having low offtake temperatures and penalizes high temperature devices.

The "cold gas efficiency" of a gasifier is not a correct measure of thermal efficiency, as it neglects a number of important energy inputs to and outputs from the gasifier, i.e.:

a) It neglects to account for the power required to produce the oxidant (for oxygen blown devices) and the power required to compress the oxidant to gasifier pressure.

b) Cold gas efficiency does not account for the energy entering the gasifier via the steam feed.

c) Sensible and latent heat in the product fuel gas (which can be converted into steam and, thus power) is not included in cold gas efficiency calculations.

To illustrate the point that a consideration of "cold gas efficiency" only can lead to misrepresentation of gasifier performance, a new gasifier efficiency, termed "actual efficiency" can be defined as follows:

$$\text{Actual Efficiency} = \frac{\text{Total Fuel Gas Useful Energy}}{\text{Total Energy Input}}$$

where total fuel gas useful energy is represented by the sum of fuel gas HHV, sensible heat and latent heat. Total energy input is the sum of coal HHV plus sensible heats, air sensible plus latent heats (to air separation plant), enthalpy of live steam to gasifier, auxiliary power requirements and enthalpy of steam required to operate an air separation plant and oxidant compressors.

As an illustrative example, "cold gas efficiencies" and "actual efficiencies" of the BGC slagging gasifier (MXSC) and the Texaco gasifier (EXTC) are shown below:

	BGC Slagger (MXSC)	Texaco (EXTC)
Gasifier Cold Gas Efficiency	93.9%	79.9%
Gasifier Actual Efficiency	87.4%	85.7%

The above table indicates that on a cold gas efficiency basis, the slagging gasifier appears to be far superior to the Texaco device. However, if all inputs and outputs are considered, both gasifiers appear to have equivalent efficiencies.

In a highly integrated system (such as a gasification-combined cycle (GCC) power plant), however, it does not make sense to place much importance on the thermal efficiency of any one unit or any partial combination of units in the system. The only quantity of importance is the overall system efficiency (i.e., coal to electricity) which is a complex function of not only the individual component efficiencies, but also of the manner in which these components are linked together. To illustrate this point, it might appear to make sense to split GCC power plant into two parts, i.e., the gas processing section and the power production (combined cycle) section. It would be possible to calculate an overall thermal efficiency for the gas processing section of the plant by considering all steam and BFW inputs from the power section as inputs to the gas plant as well as all steam and condensate returns to the power plant as outputs from the gas plant. Once again the BGC slagging gasification plant and the Texaco gasification plant can be used as examples:

	BGC Slagging Plant (Gas Only) MXSC	Texaco Plant (Gas Only) EXTC
Gasifier Cold Gas Efficiency	93.9%	79.9%
Gasifier Actual Efficiency	87.4%	85.7%
Overall Gasification Plant Efficiency	82.3%	83.6%

This table demonstrates the fact that although the slagging gasifier is more efficient than the Texaco gasifier, the Texaco gasification plant is more efficient for energy production (including steam) than the slagging gasifier plant.

Finally, the following table includes the overall system efficiency (coal to electricity) for the two examples given above:

	BGC Slagging System MXSC	Texaco System EXTC
Gasifier Cold Gas Efficiency	93.9%	79.9%
Gasifier Actual Efficiency	87.4%	85.7%
Overall Gasification Plant Efficiency	82.3%	83.6%
Overall System Efficiency (Coal to Power)	40.5%	38.7%

Once again it can be seen that the overall system efficiency has contradicted the anticipated results based on the overall gasification plant efficiencies. The above table illustrates, therefore, the hazards involved with trying to infer overall system efficiencies from component or subsystem efficiencies. In this report, therefore, individual component or subsystem efficiencies will not be quoted as they are deemed to be misleading.

The gasification cases selected for evaluation include three entrained bed processes, as well as the standard Lurgi process (MACW), and the modified Lurgi Slagging gasifier developed by the British Gas Corporation (MXSC). The standard Lurgi case represents gasification technology that is available today as it has been widely used in commercial scale applications in Africa and Europe for years.

The BGC slagging gasifier has been successfully demonstrated on a pilot scale at the Westfield Development Center by the British Gas Corporation. On the basis of these trials, the slagger can be considered ready for the next step, incorporation into a demonstration plant.

The various entrained bed gasification processes, except for the Texaco process, will probably only be available in a somewhat later time frame requiring added development before they are ready for commercialization. The Texaco process has been commercially used in the United States for gasification of petroleum residue, and extensive pilot work has been done on coal. A 150 ton/day Texaco coal gasification plant has been built in West Germany and is now in preparation for start-up.

The entrained bed processes, in these evaluations, were not drastically different economically from the Lurgi or the Slagger. They do, however, have the advantage of simplicity. Because of their higher operating temperatures, none of these processes generate hydrocarbon by-products. This simplifies plant design, and operations, and may make these plants more adaptable to future environmental regulations. This would be particularly true if stricter criteria were enforced regarding exposure to aromatics, phenolics and/or hydrocarbons. These compounds are destroyed in the various entrained bed processes.

A brief discussion of salient points related to each case follows. Except where specifically noted, the capital costs for all these cases are surprisingly similar.

MACW - Lurgi, Western Coal

This case differs from all others reported in this study as it has been based on a Western Coal instead of Illinois #6. It employs 20 conventional air blown Lurgi gasifiers and a large excess of steam to keep the ash below slagging temperatures. Liquid by-products are collected and burned in a separate by-products boiler to generate high pressure steam for power generation. This handling was dictated by the doubtful feasibility of recycling the products, and by the ground rule that no by-products other than sulfur or ammonia were allowed. The by-products boiler requires a particulate and sulfur removal system for the stack gases. This case has the lowest thermal efficiency (highest heat rate) of any of the gasification cases. This is due to a variety of factors, including in approximate order of importance:

a) Excessive effluent heat losses due to large amounts of steam present and use of direct water quench in the gasifier outlet. This reduces the heat energy to a low temperature level, and renders a sizeable portion of it useless.

b) Unrecoverable heat in the ash.

c) Energy used in treating the relatively large volumes of process condensate produced.

This case recovers the lowest fraction of its total power in the gas turbine. Gas turbines are marginally more efficient than steam turbines for generating power, so this is a further penalty. This occurs in part because the liquid by-product energy is used in the steam cycle. The gasifier is air blown, and the sulfur content of the coal is low. This means that the volume of gasifier effluent is high because of the nitrogen in the stream, and the sulfur in the product gas is diluted. This in turn caused the design for the Selexol unit to be severe in this case, requiring a high Selexol circulation rate. Selexol unit's heat losses were 50 percent higher in this case than in any other case. This added penalty alone amounts to about 1 percent of overall thermal efficiency compared to the other cases. Because of the large number of gasifiers, this case suffered a minor penalty in added system heat losses.

<u>MXSC - BGC Slagger, Illinois #6 Coal</u>

This case uses 6 Lurgi type gasifiers, modified for slagging operation. It is among the most efficient studied. The by-products are separated, collected and recycled to the gasifier to extinction. This approach is deemed feasible due to the high temperature existing in the combustion zone. This allows the by-products to participate in the gas turbine power cycle. Almost 70 percent of the net power produced is produced by the gas turbine in this case. The low amount of steam used in the gasifier results in a high heating value product and reduced gasifier effluent cooling losses. This case, like other oxygen blown cases, uses a significant amount of energy to compress the required oxygen to gasifier pressures. Because of the efficient use of steam in this gasifier (±90% conversion), the requirement for oxygen is minimized and this case has the lowest energy loss associated with oxygen compression of any of the oxygen blown cases. This case produces less process condensate requiring treatment compared to case MACW, thus effecting approximately another 1-2 percent energy savings.

<u>EXHC, EAHC - Foster Wheeler, Illinois #6 Coal</u>

These cases were among the best in terms of overall thermal efficiency. Case EAHC achieved the lowest total cost of electrical services. Both these cases had 60 percent or more of their power produced in the gas turbine cycle. The air blown case achieved a slightly higher overall thermal efficiency; the difference was 2 percent and is probably significant.

Because of the larger volume of gas in the air blown case, some heat was shifted from the gas turbine to the steam turbine, and steam turbine condenser losses were higher. These losses were offset in the oxygen blown case by high losses in the oxygen compressor turbine and intercoolers. Case EXHC was more thermally efficient than case MACW because of its reduced heat losses in the ash (~1%), the lack of effluent treating requirements (~2%) and reduced load on the Selexol plant because of having a more concentrated sulfur content in the gasifier effluent (~1%).

This process, in a manner analogous to case MXSC, makes efficient use of steam and produces an energy-rich product gas. The steam serves as both a cooling and gasifying medium, and, in case EAHC, where large amounts of nitrogen are present, only a minimal amount of steam is required for cooling.

<u>EALC, EALC-LP - Combustion Engineering, Illinois #6 Coal</u>

These cases operate at atmospheric pressure and use no steam. Like case EXTC, it is a partial oxidation process. This case produces 74 percent of its net energy from the gas turbine. This is a reflection primarily, however, of the very large amount of steam energy used and lost in the fuel gas compressor. This compressor is required to take the product gas at slight vacuum conditions and deliver it at pressure to the combustor inlet. In these evaluations, the compressor is a conventional steam turbine driven centrifugal machine with intercooling. The intercooling alone uses nearly 8 percent of the coal HHV. This entire amount could not be recovered by a nonintercooled compressor, however, because of increased horsepower required to compress hot gases. This fuel gas compression system is very expensive adding enough to the plant capital cost to drive the overall cost of services for this case to the highest of any of the cases. It is important to realize that with the development of advanced fuel gas compressors, the cost of electricity from this system might be significantly decreased. Further study to confirm such an idea is warranted. The time lag for development of advanced nonintercooled (perhaps axial flow) compressors may mean that this process will be available in a later time frame than some of the others discussed in this report.

Case EALC-LP was investigated briefly to determine the value of selecting a low pressure ratio gas turbine in the Combined Cycle System. This turbine selection not only reduced the cost of the combined cycle power block, but reduced the fuel gas supply pressure requirement for the gas turbine, resulting in less expensive fuel gas compressors. It is estimated that the savings in these two areas would reduce the total capital requirements by about $68/kW resulting in a reduction in the cost of electricity of 2 mills/kWh. Case EALC-LP appears to be the more favorable of the two cases.

A Stretford sulfur removal system is used here instead of a Selexol unit, but this has no significant impact on the overall energy balance.

EXTC, EXTC-DF - Texaco, Illinois #6 Coal

These cases are intermediate in overall cost of services to the group MXSC, EXHC, EAHC, and the group MACW and EALC. A thermal efficiency of approximately 38.7 percent is achieved. This gasification process, despite the presence of some water in the gasifier as either slurry water or steam, uses mainly oxygen to achieve gasification. The process is analogous to partial oxidation and as such, it seems a relatively minor extension of Texaco's commercial experience in the partial oxidation of petroleum residue, so that this entrained bed process would be expected to be commercially available at an early date.

Because the gasifiers operate in the 600 psig range, which is higher than any of the other cases, the losses for oxygen compression are greater here. This pressure was selected to reduce the number of gasifiers required to five. These cases have the highest requirement for oxygen of any of the oxygen blown cases. This, coupled with the high pressure, cost this case about 5 percent of the coal HHV in intercooling for the oxygen compressors alone.

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

A conceptual dry feed case was developed by EPRI and Fluor in order to estimate what incentive there might be develop a method of feeding coal dry to the gasifier. Analysis of the dry feed case, however, shows that the overall thermal efficiency is not improved compared to the slurry feed case. It is concluded that there is no incentive to develop a dry feed system if the coal being considered can be slurried with less than 35% water.

TABLE D-1

OVERALL PERFORMANCE SUMMARY

CASE	LURGI MACW	BGC SLAGGER MXSC	FOSTER WHEELER EXHC	FOSTER WHEELER EAHC	COMBUSTION ENGINEERING EALC	TEXACO EXTC	COAL FIRED BOILER
OPERATING PARAMETERS							
Coal Type	Western	Illinois #6					
Oxidant	Air	Oxygen	Oxygen	Air	Air	Oxygen	--
Overall Efficiency, %	34.96	40.6	38.5	40.5	38.1	38.7	34.4
Total Capital Requirements, $/kW[1]	906	711	739	705	931	816	838
Cost of Services Mills/kWh[1]	41.2	32.8	34.1	32.5	41.4	37.2	40.9
% of Total Power from Gas Turbine	58	69	67.6	60	74.0	62.4	-
Lbs Oxygen/Lb m.f. Coal	0.36	0.48	0.61	0.67	1.02	0.86	-
Lbs Air/Lb m.f. Coal[2]	1.56	-	-	2.86	4.37	-	-
Total Lbs Steam/Lb m.f. Coal[3]	0.76	0.31	0.62	0.15	0	0.46	-
% Coal Carbon Converted to CH_4	11.1	11.2	11.1	9.5	0	0.16	-
HEAT LOSS FLOWS, MM BTU/HR							
Total Stack Gas	1899	2126	2084	1853	1541	1817	
Power Surface Condenser	2515	2221	2180	3099	1452	2687	
Compressor Surface Condenser	281	683	855	337	2053[4]	1067	
Compressor Intercoolers	-	267	305	-	819[4]	568	
Gasifier Effluent Cooling	506	270	268	224	-	25	
Ash Heat (Sensible and HHV)	254	123	159	157	138	81	
Process Condensate Treating	351	171	100	140	-	-	
Acid Gas Removal	319	132	251	212	179	78	

[1]Based on $1.00/MM Btu Coal.
[2]Dry Basis
[3]Excludes moisture in coal and oxidant streams, but includes moisture in char and coal slurries.
[4]Product Gas Compressor.

CAPITAL COST COMPARISONS

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

It can be seen from this table that all coal handling costs are similar except for Case EALC. Coal handling costs for this case are high due to the multiplicity of pulverization trains and coal storage silos necessary due to the dual zone feeding required by the gasifier. The oxidant feed system cost for the Texaco case (EXTC) is almost double that for any of the other oxygen blown systems. This is partially due to the fact that the Texaco gasifier consumes more oxygen per pound of coal than any of the others considered, and partially due to the fact that the Texaco gasifier operates at higher pressures than the others.

Costs for the gasification and ash handling sections of each plant indicate that Lurgi gasifiers would benefit greatly from increased capacity and that the Texaco gasifiers are indeed the simplest and least costly of all that were studied. However, gas cooling section costs indicate that the Texaco system pays part of the price of low gasifier cost by requiring costly gas cooling equipment.

Process condensate treating costs for both of the fixed bed systems (MACW and MXSC) are an order of magnitude higher than those for the entrained systems which are anticipated not to produce hydrocarbon liquid by-products. This cost of process water treatment indicates strongly the incentive to develop by-product free coal gasification devices.

Only one of the cases studied (EALC) required fuel gas compressors to compress gas to gas turbine inlet condition requirements. The cost of $157/kW demonstrates graphically the desirability for developing low cost gas compression equipment or restructuring power cycles for use with low pressure gasifiers.

The cost of the steam, condensate and BFW system for the Lurgi case (MACW) is considerably higher than that cost for any of the other systems. This is mainly due to the fact that costly boilers, stack gas scrubbers and electrostatic precipitators are required to dispose of the liquid hydrocarbon by-products produced in the gasifiers, showing once again the desirability of developing by-product free gasifiers.

Finally, it is of interest to examine the percentage contribution of each subsection of each plant to the installed plant cost. Such information is shown in Table D-3. These data indicate that the costs of gasification and ash handling contribute between 4 percent and 17 percent to the total installed plant cost with the average being less than 10 percent. This suggests that the economic incentive for developing new gasifiers is not great (i.e., reducing the cost of the gasification section of these types of systems is not terribly significant). Developing a gasifier, however, that will reduce the need for downstream processing equipment could have major economic significance.

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

TABLE D-2

SUMMARY OF CAPITAL INVESTMENT ESTIMATES - $/KW

PLANT INVESTMENT	LURGI MACW	BGC SLAGGER MXSC	FOSTER WHEELER EXHC	FOSTER WHEELER EAHC	COMBUSTION ENGINEERING EALC	TEXACO EXTC
Coal Handling	25.90	12.31	25.52	24.24	41.40	19.07
Oxidant Feed	19.36	47.85	58.23	17.74	1.85	101.48
Gasification and Ash Handling	101.34	40.99	30.99	35.30	78.63	20.97
Gas Cooling	19.40	7.38	16.88	26.58	12.32	57.91
Acid Gas Removal and Sulfur Recovery	30.35	19.41	23.74	30.10	21.40	24.71
Process Condensate Treating	59.80	25.74	6.66	8.13	--	--
Fuel Gas Compression	--	--	--	--	153.97	--
Steam, Condensate and BFW	42.51	1.70	2.40	1.36	0.74	0.71
Support Facilities	63.25	47.98	50.10	51.13	49.13	47.72
Combined Cycle	235.10	264.60	258.88	253.50	242.56	262.94
Contingency	107.88	85.40	101.65	100.65	125.88	102.14
TOTAL PLANT INVESTMENT	704.89	553.36	575.05	548.73	727.88	637.65
ILLINOIS SALES TAX	16.19	12.92	12.92	12.27	16.43	14.40
CAPITAL CHARGES						
Preproduction Costs	44.38	35.18	36.47	34.85	45.75	40.06
Paid-up Royalty	3.52	2.77	2.88	2.74	3.64	3.19
Initial Catalyst and Chemical Charges	1.09	0.66	0.74	1.12	0.47	0.44
Construction Loan Interest	89.94	69.11	71.82	68.54	90.91	79.64
TOTAL CAPITAL CHARGES	138.93	107.72	111.91	107.25	140.77	123.33
WORKING CAPITAL	45.66	37.12	38.67	36.90	47.57	41.15
TOTAL CAPITAL REQUIREMENTS	905.67	711.12	738.55	705.15	930.65	816.53

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

TABLE D-3

PERCENTAGE CONTRIBUTION OF PLANT SUBSECTIONS
TO INSTALLED PLANT COSTS

PERCENT

PLANT INVESTMENT	LURGI MACW	BGC SLAGGER MXSC	FOSTER WHEELER EXHC	FOSTER WHEELER EAHC	COMBUSTION ENGINEERING EALC	TEXACO EXTC
Coal Handling	4.36	2.63	5.39	5.41	6.88	3.56
Oxidant Feed	3.25	10.23	12.30	3.96	0.31	18.95
Gasification and Ash Handling	17.00	8.76	6.55	7.88	13.06	3.92
Gas Cooling	3.25	1.58	3.57	5.93	2.05	10.81
Acid Gas Removal and Sulfur Recovery	5.09	4.15	5.01	6.72	3.55	4.61
Waste Water Treating	10.03	5.50	1.41	1.81	--	--
Fuel Gas Compression	--	--	--	--	25.58	--
Steam, Condensate and BFW	7.13	0.36	0.51	0.30	0.12	0.14
Support Facilities	10.44	10.25	10.58	11.42	8.16	8.91
Combined Cycle	39.45	56.54	54.68	56.47	40.29	49.10
INSTALLED PLANT COST	100.00	100.00	100.00	100.00	100.00	100.00

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

CASE MACW - LURGI

PLANT DESCRIPTION - CASE MACW

GENERAL

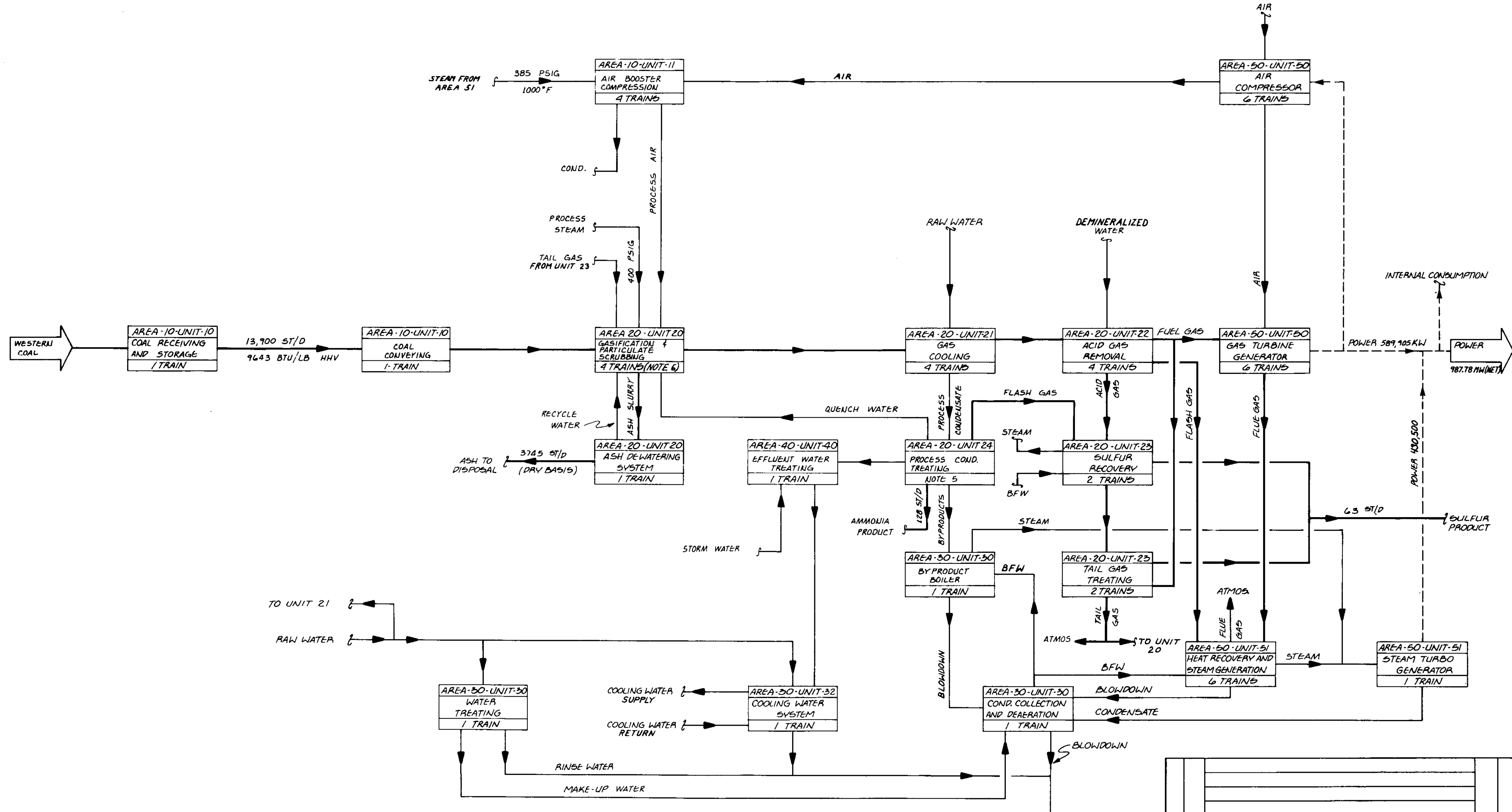
A grass roots plant for power generation based on dry ash moving bed air blown gasifiers of the Lurgi type is shown schematically on Block Flow Diagram MACW-1-1. This plant consumes 13,900 ST/day of western coal.

The main processing units are in four parallel and largely independent trains. Each process train consists of oxidant feed, gasification, gas cooling and acid gas removal units. Integration between processing trains is minimized. Complete trains may be shut down in order to maintain efficiency during reduced capacity operation. The impact of upset conditions is limited to the train in which the upset occurs.

In addition to the main processing trains, the complete plant includes necessary offsite, utility and environmental facilities. Coal receiving, storage, and conveying is accomplished in a single train to minimize space and operating labor requirements. Hydrogen sulfide removed from the crude fuel gas is processed through sulfur recovery facilities which produce elemental sulfur.

Other facilities in the plant include raw water treating, cooling water, process condensate treating and effluent water treating.

Table M-1 summarizes major equipment sections in the plant and shows the numbers of operating and spare sections.



NOTES:

1.) OPERATING TRAINS ONLY ARE SHOWN.
2.) FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3.) STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE, SENSIBLE HEAT ABOVE 60°F AND LATENT HEAT.
4.) EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
5.) PROCESS CONDENSATE TREATING CONSISTS OF TWO TAR OIL SEPARATION UNITS AND ONE PHENOL EXTRACTION UNIT.
6.) EACH TRAIN CONSISTS OF FIVE OPERATING GASIFIERS.

DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY B. GEORGE / R. KRESS
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
SUPERVISING ENGR FEERRAR

OVERALL BLOCK FLOW DIAGRAM
MOVING BED COAL GASIFICATION
AIR BLOWN - COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE
DRAWING NUMBER MACW-1-1
REVISION 1

TABLE M-1

MAJOR EQUIPMENT SECTIONS - CASE MACW

Unit No.	Name	Operating	Spare
10	Coal Preparation	1	0
11	Oxidant Feed System	4	0
20	Gasification	4*	0
20	Ash Handling	1	0
21	Gas Cooling	4	0
22	Acid Gas Removal	4	0
23	Sulfur Recovery & Tail Gas Treating	2	1
24	Process Condensate Treating		
	Tar Oil Separation	2	1
	Phenol Extraction	1	0
	Ammonia Recovery	1	0
30	Steam, BFW and Condensate System		
	By-product Boiler	1	1
	Condensate Collection & Deaeration	1	0
	Water Treating	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0
50	Gas Turbine/Generator	6	0
51	Heat Recovery Steam Generator	6	0
51	Steam Turbine/Generator	1	0

*Each train includes five parallel gasifiers resulting in a total of twenty operating gasifiers. Two additional spare gasifiers are provided for the entire plant.

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COAL PREPARATION

Process Flow Diagram MACW-10-1 depicts the process arrangement of equipment in this section.

Washed, 1-1/2" by 1/4" coal is received at the plant site by unit train. No crushing, grinding and refuse disposal systems are included. The coal is unloaded from 100 ton bottom dump cars into an unloading hopper, withdrawn from the hopper by two vibrating feeders and transported by belt conveyors to a tripper. The tripper distributes coal to a traveling belt stacking system. The stacker travels on tracks and forms storage piles on either side. The unloading and stacking system is designed to handle a three day supply in eight hours.

Coal is reclaimed from storage piles by a bridge type bucket wheel reclaimer rated at 700 tons per hour. This machine is a rail mounted bridge which supports a rotating bucket wheel and belt conveyor. The wheel moves across the face of the pile, making a vertical cut across the many layers of coal. At the end of each cut, the reclaimer moves ahead a predetermined distance and the wheel makes another cut in the opposite direction. The excavated coal is carried by a series of conveyors to a tripper, which distributes the feedstock to the coal hoppers above the operating gasifiers.

Equipment Notes

All equipment is commercially available.

10-BN-1
UNLOADING HOPPER

10-CV-1
BELT CONVEYOR

10-CV-2
BELT CONVEYOR

10-TR-1
TRIPPER

10-ME-1
STACKING SYSTEM

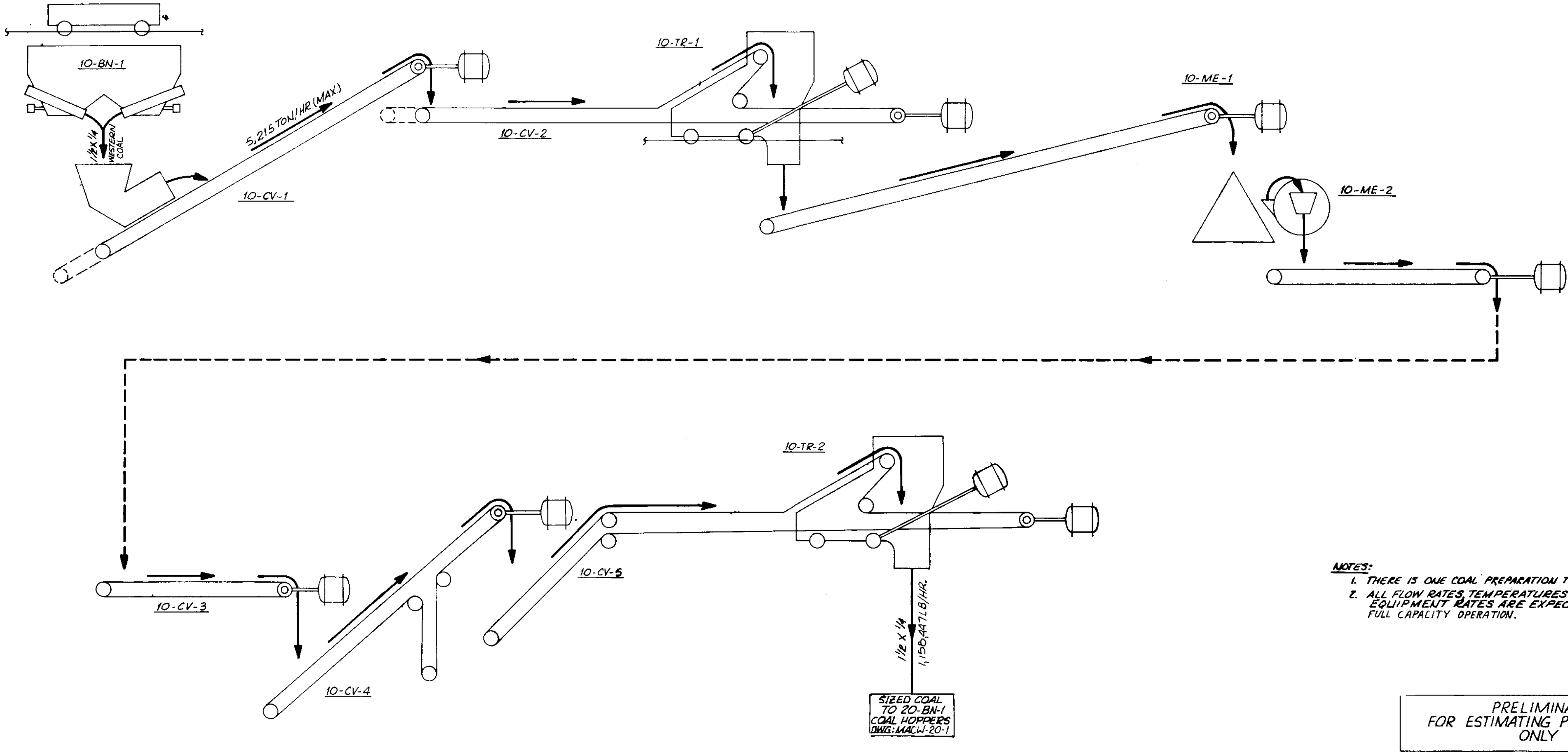
10-ME-2
RECLAIMING SYSTEM

10-CV-3
BELT CONVEYOR

10-CV-4
BELT CONVEYOR

10-CV-5
BELT CONVEYOR

10-TR-2
TRIPPER



NOTES:

1. THERE IS ONE COAL PREPARATION TRAIN.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES AT FULL CAPACITY OPERATION.

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FOR ESTIMATING PURPOSES
ONLY

REV NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY: M. SEKHON
CHECKED BY: R. RICE
SUPERVISOR: J. BARGER
SUPERVISING ENGR: COTTINGHAM
RELEASE DATE
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
COAL PREPARATION
MOVING BED-AIR BLOWN-COMBINED CYCLE
WESTERN COAL

ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: MACW-10-1
REVISION: 2

(5,X3) MACW-10-1 ASE83

OXIDANT FEED SYSTEM

Process Flow Diagram MACW-11-1 depicts one of the four parallel trains.

Air for the gasifier is obtained as a bleed stream from the gas turbine air compressor. Hot bleed air at 225 psig, 857°F, exchanges heat with cooled Therminol in 11-1-E-1 and after further cooling in an air cooler, 11-1-E-2, to 174°F, is compressed to 380 psig in Booster Air Compressor 11-1-C-1. Hot air at 380 psig and 300°F from 11-1-C-1 then flows to the gasifier.

The booster air compressor, 11-1-C-1, is a centrifugal type machine. The driving power, 5720 hp, for 11-1-C-1 is supplied by two condensing type steam turbines, 11-1-T-1 and 11-1-T-2. Turbine 11-1-T-1 uses superheated 100 psig steam at 710°F and Turbine 11-1-T-2 takes saturated 50 psig steam. The steam from the turbine drivers exhausts at 2-1/2" Hg absolute. The condensate produced in the surface condenser, 11-1-E-4, is pumped to the gas cooling unit (Flow Diagram: MACW-21-1).

The closed loop Therminol system is used to transfer heat indirectly from hot bleed air from the gas turbine air compressor to product fuel gas from the acid gas removal system. This way formation of an explosive mixture because of leakage in a direct heat exchanger is prevented. Therminol is circulated by Pump 11-1-P-1A or B through heat exchangers 11-1-E-1 and 11-1-E-3. In heat exchanger 11-1-E-1, Therminol is heated from 150°F to 600°F by recovering heat from hot bleed air which in turn is cooled down to 210°F. The hot Therminol after exchanging heat against product fuel gas in 11-1-E-3 flows back to Therminol Tank 11-1-V-1. Hot fuel gas flows to the gas turbines.

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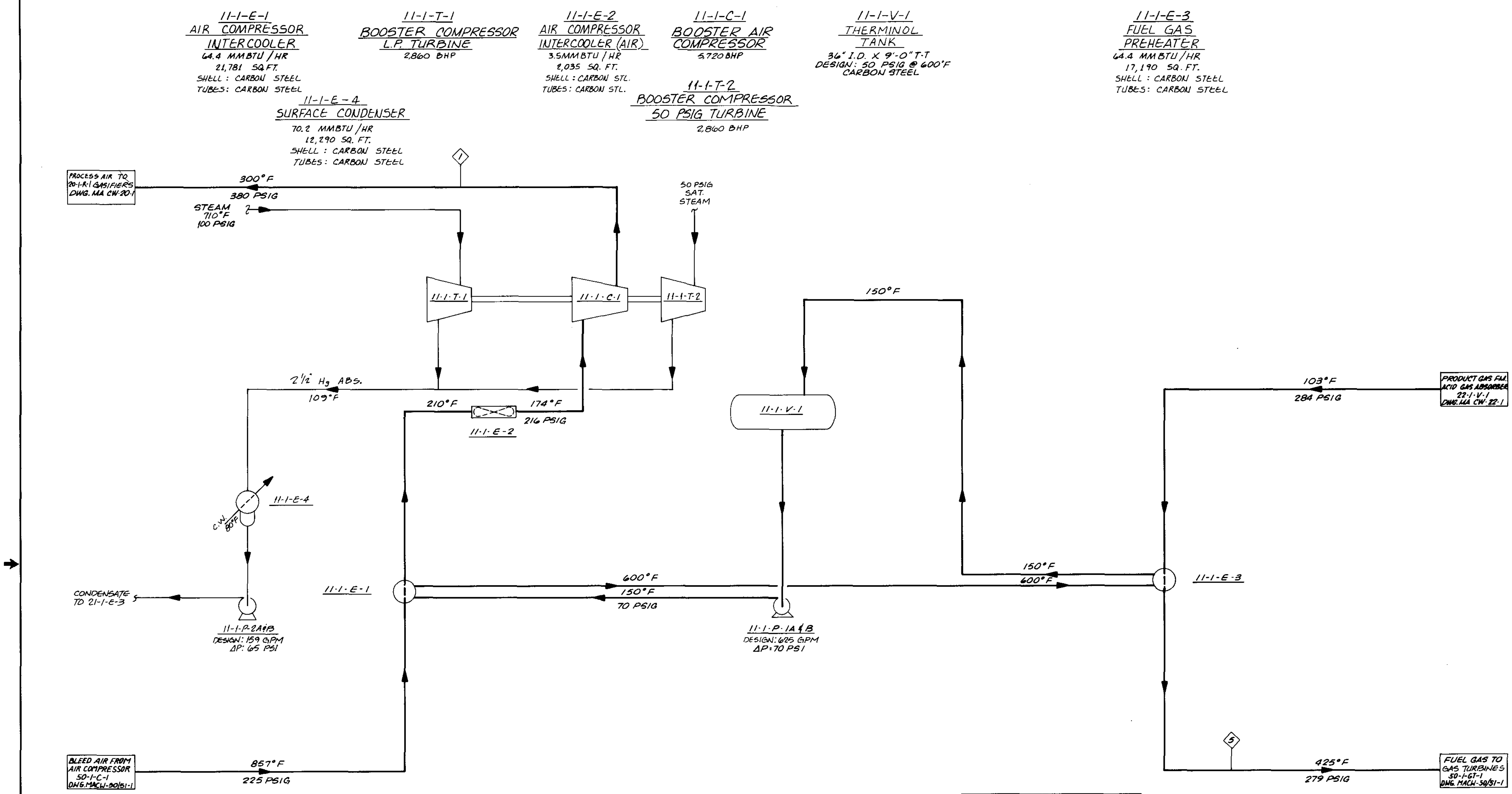

PRELIMINARY
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MATERIAL BALANCE				
STREAM NUMBER	1		5	
	PROCESS AIR		FUEL GAS	
COMPONENT	MPH	MOL%	MPH	MOL%
CH_4	—	—	1,251.9	4.89
C_2H_4	—	—	50.7	0.20
C_2H_6	—	—	33.6	0.13
H_2	—	—	6,033.4	23.57
CO	—	—	3,754.7	14.67
CO_2	—	—	3,584.3	14.00
H_2S	—	—	22.3	0.09
COS	—	—	—	—
NH_3	—	—	—	—
N_2	10,885.6	77.85	10,817.1	42.25
O_2	2,884.8	20.69	—	—
H_2O	203.7	1.46	51.6	0.20
TOTAL MPH	13,944.1	100.00	25,599.6	100.00
OTHERS LB/HR	—		—	
TOTAL LB/HR	400,082		602,310	
MOL. WT.	28.69		23.53	

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

8-74 DIETERICH-POST 2111X-8 .003 FILM

NOTES:

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

REV NO	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY T. BUJACICH
CHECKED BY R. RICE
SUPERVISOR R. RICE
RELEASE DATE
SUPERVISING ENGR COTTINGHAM
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
MOVING BED-AIR BLOWN-COMBINED CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE
DRAWING NUMBER MACW-11-1
REVISION 4

REDUCED PRINT SCALES

44E8SA-MW-CW-11-1 (S'x'E)

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GASIFICATION AND ASH HANDLING

Process Flow Diagram MACW-20-1 shows the gasification step. There are four parallel trains, each train having five parallel gasifiers. Two additional spare gasifiers are provided.

The Lurgi moving bed gasifier is a water jacketed pressurized unit composed of a series of vertically stacked vessels. There are, from top to bottom, a coal hopper, coal lock, water jacketed gasifier, ash lock and ash quench chamber.

Coal is conveyed from the coal preparation area to the coal hopper from which it is fed by gravity to the depressurized coal lock through a hydraulically operated valve. The lock is then isolated and pressurized with a slipstream of desulfurized tail gas (mainly N_2) and the coal is transferred to the gasifier through another hydraulically operated valve. The empty lock is isolated, depressurized through a bag filter and vented to the atmosphere. In addition, the gas displaced from the coal and lock hoppers during loading is vented to the atmosphere through the bag filter. Coal dust recovered in the filter is conveyed to the briquetting facility.

The coal flowing down through the gas producer represents a slowly moving bed which has several distinct zones. In the first zone at the top of the gasifier, coal is preheated and dried by contact with the hot crude gas leaving the reactor. As the coal moves down and is heated further, devolatilization occurs and gasification commences. The bottom of the bed is a combustion zone where carbon reacts with oxygen to form CO and CO_2. The oxidation provides the overall heat for the gasification and devolatilization reactions which are endothermic. Only a small amount of unburned carbon remains in the ash.

Air and steam enter the gasifier near the bottom and are heated as they rise upward to the combustion zone by the hot ash moving down from the combustion zone. Air flow rate is controlled to accomplish complete gasification of coal. Steam rate is controlled to maintain a specified maximum temperature near the bottom of the gasifier to prevent melting or clinkering of the ash.

Approximately 22% weight of the gasifier process steam is generated at 400 psig in the gasifier jacket. Each gasifier produces about 8844 lb/hr of saturated steam. The balance of the process steam demand of the gasifiers is met by

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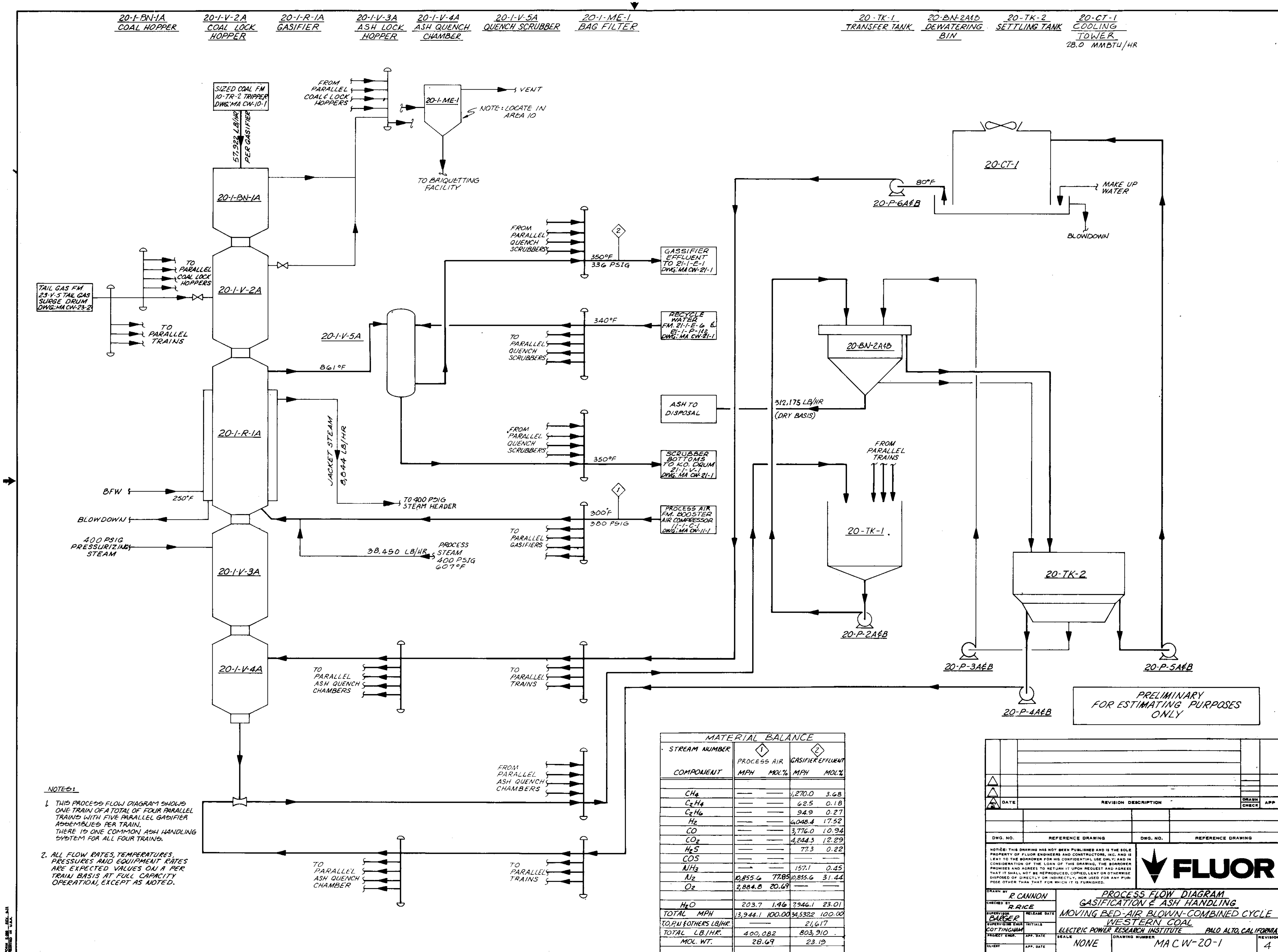

NOTES:

1. THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FOUR PARALLEL TRAINS WITH FIVE PARALLEL GASIFIER ASSEMBLIES PER TRAIN. THERE IS ONE COMMON ASH HANDLING SYSTEM FOR ALL FOUR TRAINS.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.

MATERIAL BALANCE

STREAM NUMBER / COMPONENT	1 PROCESS AIR MPH	MOL.%	2 GASIFIER EFFLUENT MPH	MOL.%
CH_4	—	—	1,270.0	3.68
C_2H_4	—	—	62.5	0.18
C_2H_6	—	—	94.9	0.27
H_2	—	—	6,048.4	17.52
CO	—	—	3,776.0	10.94
CO_2	—	—	4,244.3	12.29
H_2S	—	—	77.3	0.22
COS	—	—	—	—
NH_3	—	—	157.1	0.45
N_2	10,855.6	77.85	10,855.6	31.44
O_2	2,884.8	20.69	—	—
H_2O	203.7	1.46	7,946.1	23.01
TOTAL MPH	13,944.1	100.00	34,532.2	100.00
T,O,P,N & OTHERS LB/HR.	—		21,617	
TOTAL LB./HR.	400,082		803,910	
MOL. WT.	28.69		23.19	

63

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extraction at 400 psig from Turbine 51-T-1 (Flow Diagram: MACW-50/51-1) and steam generation in the sulfur plant (Flow Diagram: MACW-23-1).

The crude gas leaving the gasifier at 861°F contains appreciable quantities of tars, oils, naphtha, phenols, fatty acids, ammonia, hydrogen sulfide, sulfur compounds and a small amount of coal and ash dust. This gas flows through quench scrubber 20-1-V-5A where it is washed with a stream of process condensate. The washing process quenches the gas to 350°F and condenses the high boiling tar fractions. Coal and ash dust are removed with the condensed tar leaving the quenched effluent gas essentially free of particulate matter.

Ash from the process is continuously collected by a rotating ash grate and moved to the ash lock hopper. Ash collected in the lock is depressurized and discharged batchwise to an ash quench chamber, 20-1-V-4A, where it is cooled in water. The ash lock is pressurized with 400 psig steam.

The abrasive slurry from each gasifier train is educted to a common transfer tank, 20-TK-1, using water as the motive fluid. Ash grinders are provided to prevent large chunks of ash from plugging transfer lines. The ash slurry is dewatered in a bin, 20-BN-2A and B, producing an ash ready for disposal. Final cleaning of the water overflowing the dewatering bin, 20-BN-2, is accomplished in a settling tank, 20-TK-2, where ash fines settle and are pumped back to the dewatering bin. A portion of the clarified water is recycled to the ash quench chambers after it is cooled in an induced draft type cooling tower (20-CT-1). The balance of the water provides the motive fluid for the ash slurry transfer eductors.

Equipment Notes

The moving bed gasifier with associated coal and ash locks has been operated in commercial size plants on noncaking coals.

The ash slurry system is a commercially available system.

GAS COOLING

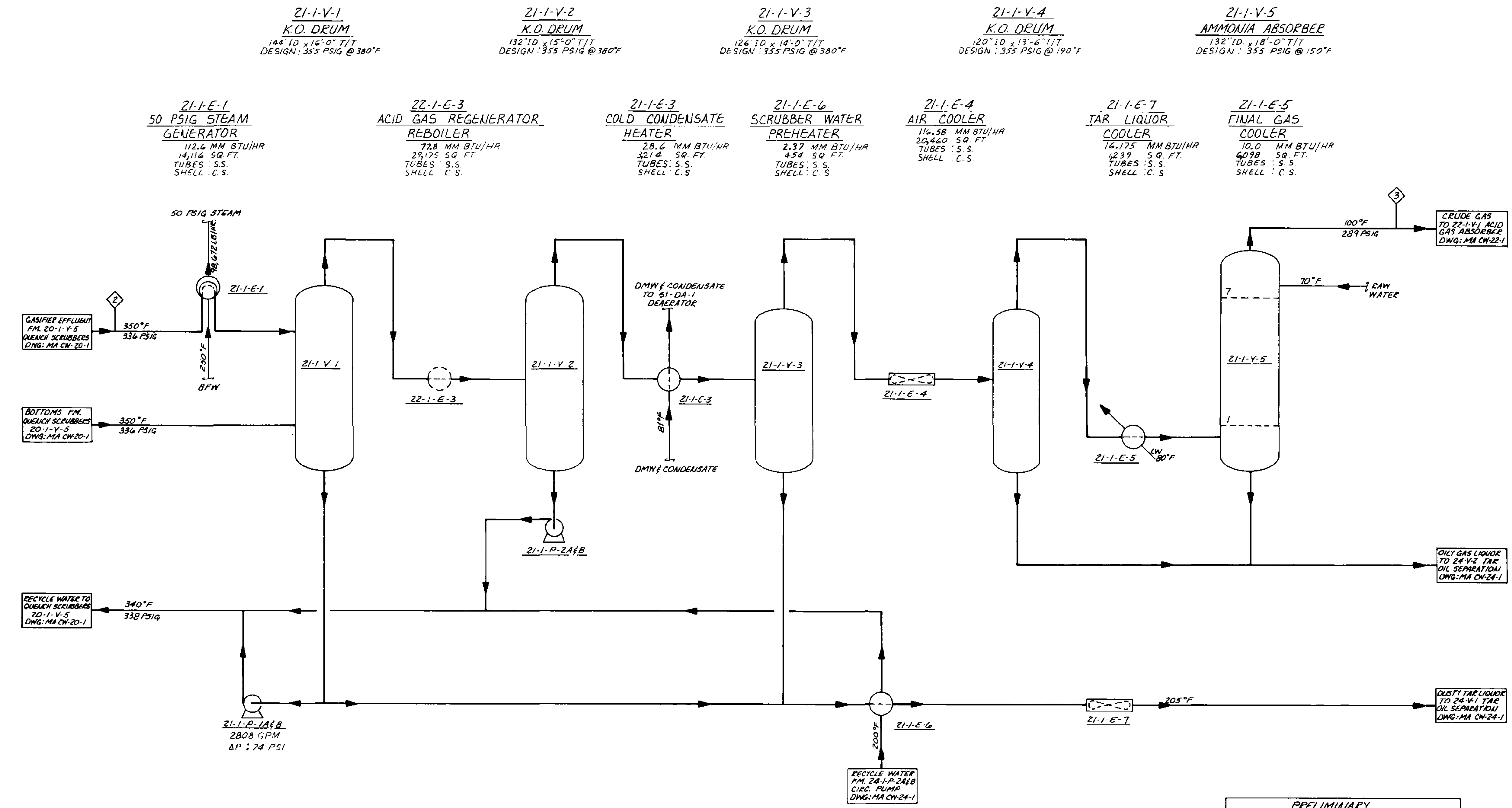
Process Flow Diagram MACW-21-1 depicts one of the four parallel gas cooling trains.

The crude gas from the quench scrubber is cooled from 350°F to approximately 105°F in a series of heat exchangers, 21-1-E-1, 22-1-E-3, 21-1-E-3 and 21-1-E-4.

Gas cooling to approximately 328°F is obtained in a kettle type boiler, 21-1-E-1, by 50 psig steam generation. After the separation of condensate produced in cooling in the knockout drum, 21-1-V-1, the crude gas flows to Regenerator Reboiler 22-1-E-3 located in the acid gas removal unit (Flow Diagram: MACW-22-1). The regenerator reboiler, 22-1-E-3, is a horizontal thermosyphon. Heat for solvent regeneration is extracted in the reboiler on the shell side from the crude gas which in turn is cooled down to 300°F on the tube side. The condensate produced in cooling is separated in the knockout drum, 21-1-V-2. The crude gas then exchanges heat with cold demineralized water and condensate in the shell and tube exchanger, 21-1-E-3, and the condensate produced is removed in the knockout drum, 21-1-V-3. Further cooling of the crude gas to 105°F is obtained by a combination of air (21-1-E-4) and water (21-1-E-5) cooling. Another knockout drum, 21-1-V-4, has been provided after the air cooler, 21-1-E-4, to separate the condensate produced in air cooling.

The cooled gas from 21-1-E-5 still contains ammonia which must be removed. This is achieved by water scrubbing in an ammonia absorber, 21-1-V-5, where gas contacts water countercurrently on trays. The ammonia-free overhead at approximately 289 psig and 100°F then flows to a Selexol® unit for further processing. The ammonia-rich water from the absorber bottom is combined with cooled oily gas condensate from the knockout drum, 21-1-V-4, and further processed in the tar oil separation unit.

The dusty tar liquor from the knockout drum, 21-1-V-1, flows to the tar oil separation unit (Flow Diagram: MACW-24-1) after heat exchange against recycle water in 21-1-E-6 and air cooling (21-1-E-7). Hot recycle water from 21-1-E-6 is combined with the gas liquor separated in the knockout drum, 21-1-V-2, and flows to the quench scrubbers located in the gasification area. A portion of the hot dusty tar liquor from the knockout drum, 21-1-V-1, is also pumped to the header transporting the recycle water to the quench scrubbers.



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MATERIAL BALANCE				
STREAM NUMBER	② GASIFIER EFFLUENT		③ CRUDE GAS	
COMPONENT	MPH	MOL.%	MPH	MOL.%
CH_4	1,270.0	3.68	1,270.0	4.79
C_2H_4	62.5	0.18	62.5	0.24
C_2H_6	94.9	0.27	94.9	0.36
H_2	6,048.4	17.52	6,048.4	22.80
CO	3,776.0	10.94	3,776.0	14.24
CO_2	4,244.3	12.29	4,244.3	16.00
H_2S	77.3	0.22	77.3	0.29
COS	—	—	—	—
NH_3	157.1	0.45	—	—
N_2	10,855.6	31.44	10,855.6	40.92
H_2O	7,946.1	23.01	96.5	0.36
TOTAL MPH	34,532.2	100.00	26,525.5	100.00
T,O,P,N & OTHERS LB/HR.	21,617		3,049	
TOTAL LB/HR.	803,910		641,248	
MOL. WT.	23.19		24.17	

FORM E-360 REV. 9-72 PRINTED IN U.S.A.

8-74 DIETERICH-POST 2111X-8 .003 FILM

NOTES:

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY: K. WILSON
CHECKED BY: R. RICE
SUPERVISOR: R. RICE
SUPERVISING ENGR: COTTINGHAM
PROJECT ENGR.:
CLIENT:
RELEASE DATE / INITIALS / APP. DATE / APP. DATE

PROCESS FLOW DIAGRAM
GAS COOLING
MOVING BED-AIR BLOWN-COMBINED CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: MA CW-21-1
REVISION: 4

REDUCED PRINT SCALES

(S.X.3) 1-12-MA CW-21-1 ... 4483-AM-CW-21-1 (S.X.3)

Equipment Notes

All equipment is commercially available.

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ACID GAS REMOVAL

Process Flow Diagram MACW-22-1 depicts one of the four acid gas removal trains.

The acid gas removal system employs Allied Chemical Corporation's Selexol® process for selective removal of hydrogen sulfide (H_2S). Hydrogen sulfide in the crude gas is absorbed in Selexol® solvent in order to reduce sulfur in the treated gas to 1.0 pound sulfur dioxide (SO_2) equivalent per million Btu (HHV) coal charged to the plant.

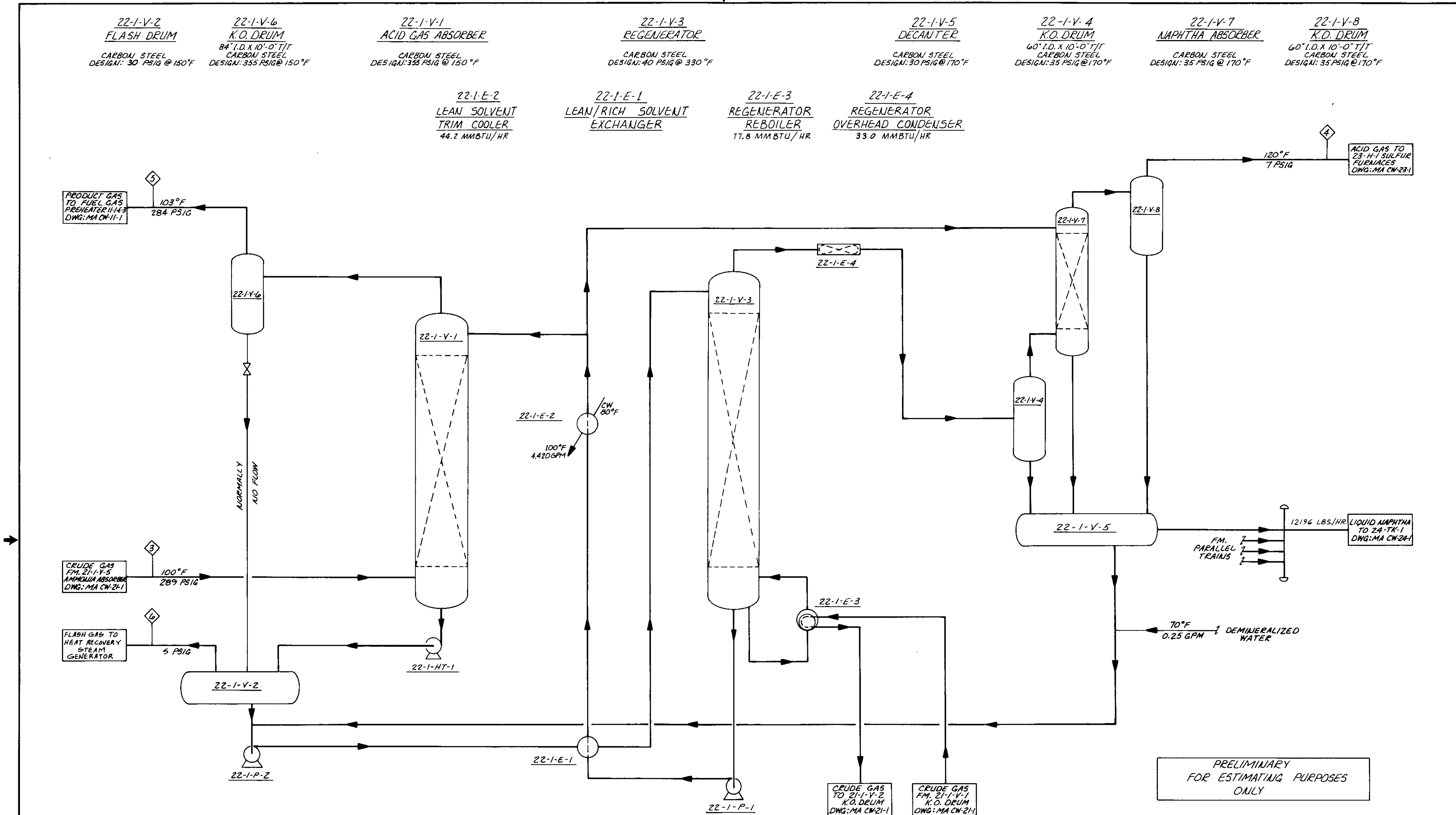
The cooled ammonia-free crude gas flows up through an acid gas absorber, 22-1-V-1, where it contacts Selexol® solvent countercurrently over a packed bed. The treated gas from the top of the absorber flows through Knockout Drum 22-1-V-6, to Unit 11 (Flow Diagram: MACW-11-1) where it is heat exchanged against the hot Therminol in 11-1-E-3. The hot fuel then flows to the gas turbines.

The rich solvent from the bottom of the absorber is let down through a hydraulic turbine, 22-1-HT-1, which supplies a portion of the power required by the lean solution pump 22-1-P-1. It then flows to a flash drum, 22-1-V-2, where most of the hydrocarbon gases in the solvent flash off. About 75 mol % of dissolved H_2S is however retained in the solvent because of its selective absorption in the Selexol solvent. The flashed gases flow to the heat recovery steam generator (HRSG) provided in the combined cycle area.

The rich solvent solution from the flash drum exchanges heat with hot regenerated solvent in 22-1-E-1 and flows to the top of the regenerator, 22-1-V-3. In the regenerator, the absorbed H_2S and CO_2 are stripped from the solvent. Reboil heat is supplied by heat exchange with the gasifier effluent in a thermosyphon reboiler, 22-1-E-3.

Hot, regenerated solvent is pumped back to the absorber, 22-1-V-1, through the exchangers, 22-1-E-1 and 22-1-E-2. Heat is first exchanged with rich solution in 22-1-E-1 in order to reduce reboiler duty. Then the lean solution is cooled down to operating temperature in a water cooler, 22-1-E-2.

Acid gas from the regenerator overhead is cooled to 120°F in an air fan cooler, 22-1-E-4. The condensate produced in cooling is separated in knockout drum, 22-1-V-4, and flows to a decanter, 22-1-V-5, by gravity.



MATERIAL BALANCE								
STREAM NUMBER	3		4		5		6	
	CRUDE GAS		ACID GAS		FUEL GAS		FLASH GAS	
COMPONENT	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	1270.0	4.79	0.5	0.18	1251.9	4.89	17.6	2.76
C_2H_4	62.5	0.24	1.8	0.63	50.7	0.20	10.0	1.57
C_2H_6	94.9	0.36	9.7	3.41	33.6	0.13	51.6	8.08
H_2	6,048.4	22.80	0.1	0.03	6,033.4	23.57	14.9	2.33
CO	3,776.0	14.24	0.3	0.11	3,754.7	14.67	21.0	3.29
CO_2	4,244.3	16.00	208.7	73.41	3,584.3	14.00	451.3	70.70
H_2S	77.3	0.29	41.0	14.42	22.3	0.09	14.0	2.19
COS	—	—	—	—	—	—	—	—
N_2	10,855.6	40.92	0.3	0.11	10,817.1	42.25	38.2	5.98
H_2O	96.5	0.36	21.9	7.70	51.6	0.20	19.8	3.10
TOTAL MPH	26,525.5	100.00	284.3	100.00	25,599.6	100.00	638.4	100.00
NAPHTHA LB./HR.	3,049		—		—		—	
TOTAL LB./HR.	641,248		11,343		602,310		24,489	
MOL. WT.	24.17		39.90		23.53		38.38	

NOTES:

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY K. WILSON
CHECKED BY R. RICE
SUPERVISOR R. RICE
RELEASE DATE
SUPERVISING ENGR COTTINGHAM
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
MOVING BED-AIR BLOWN-COMBINED CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE
DRAWING NUMBER MACW-22-1
REVISION 3

The cooled acid gas from 22-1-V-4 contains approximately 14.5 mol % H_2S and a small quantity of naphtha. Higher hydrocarbons (C_3 plus) cause problems in the downstream sulfur plant as they do not burn completely. Instead, the higher hydrocarbons undergo partial cracking resulting in carbon deposition on the sulfur converter catalyst and production of black sulfur. Naphtha is therefore removed from the acid gas. This is achieved by reabsorption of hydrocarbons (C_4-C_7), from the overhead gases of the knockout drum, 22-1-V-4, in a naphtha absorber, 22-1-V-7. The acid gas contacts a slip stream of cooled lean Selexol® solvent countercurrently over the packing in 22-1-V-7. The bottoms from the absorber then go to the decanter, 22-1-V-5. Phase separation of naphtha and Selexol® solvent is obtained in 22-1-V-5 because of the high solubility of Selexol® solvent in water. The condensate from the knockout drum, 22-1-V-4, dissolves the Selexol solvent and the water-rich phase settles in the bottom of 22-1-V-5. The lighter naphtha forms the top layer and is continuously removed from the decanter.

The hydrocarbon-free Selexol® solvent and condensate stream from 22-1-V-5, is then combined with the Selexol® solvent from 22-1-V-2 and pumped back to the regenerator. A small quantity of demineralized water is added to the Selexol® solvent downstream of 22-1-V-5 to maintain the water balance in the absorption system.

The naphtha-free acid gas from the absorber then flows through a knockout drum 22-1-V-8 to the sulfur plant.

Equipment Notes

Most of the equipment in this unit is all carbon steel. The equipment has been used in very similar services for a number of years.

The naphtha absorber is in operation in a Selexol plant for sweetening natural gas in Texas. Heavy hydrocarbons have been successfully removed from the acid gas in this equipment.

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SULFUR RECOVERY AND TAIL GAS TREATING

Process Flow Diagrams MACW-23-1, MACW-23-2 and MACW-23-3 depict the process arrangement for these units.

There are two 50% parallel operating sulfur recovery trains, each followed by a tail gas treating unit. Sulfur recovery is 63 short tons/day for both trains. There is a third (spare) train because of the important environmental requirements these units fulfill.

Sulfur Recovery Unit

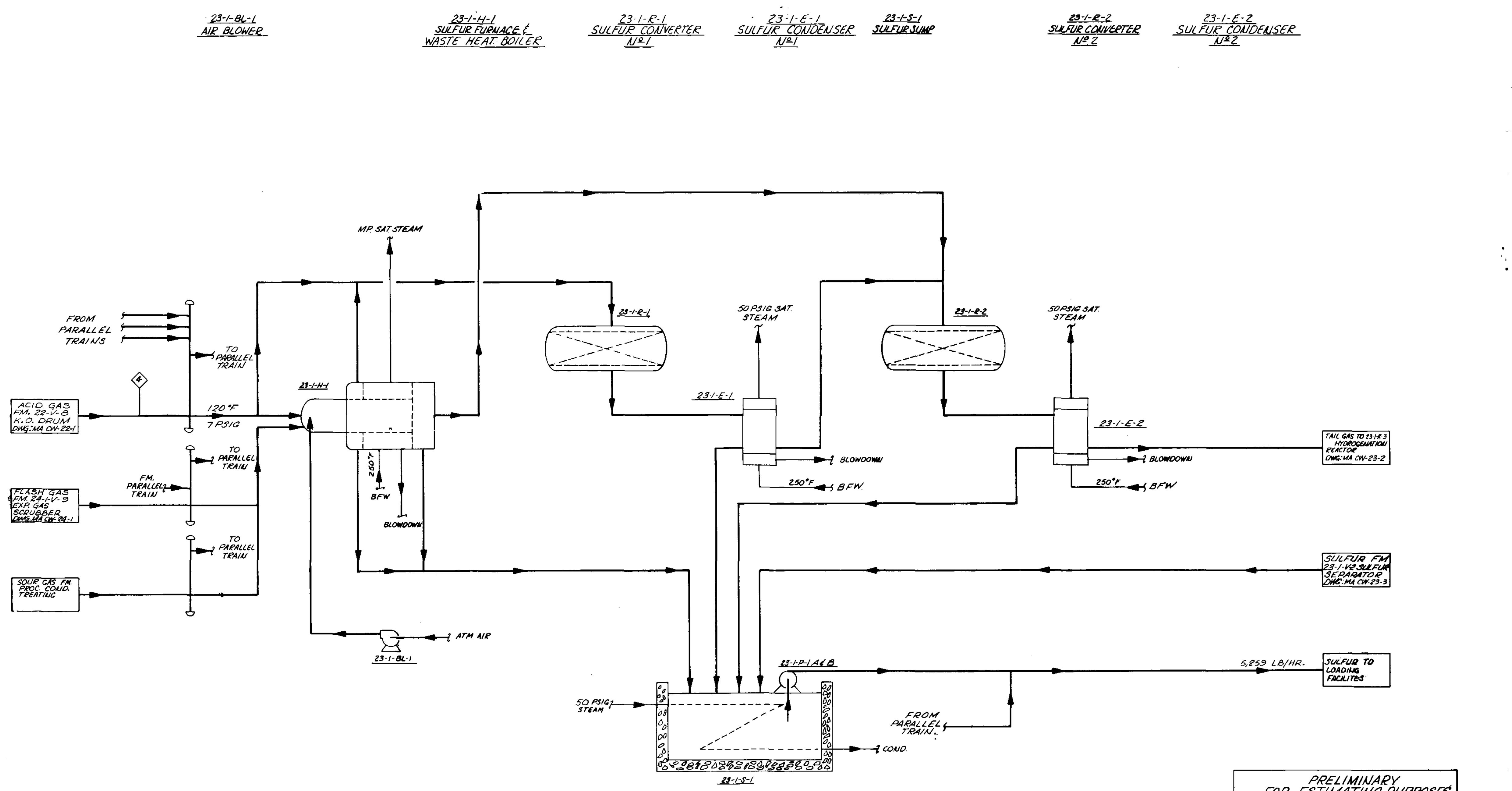
The sulfur recovery unit is a two stage, acid gas bypass type Claus unit with hot gas reheat. About one-third of the gas from the acid gas removal (Selexol) unit and the sour acid gas (containing ammonia) from the process condensate treating unit are burned in a sulfur furnace to convert H_2S to SO_2. Air is supplied to the furnace by a blower. Heat is recovered from the combustion products by generating 400 psig steam in a waste heat boiler. The cooled material from the sulfur furnace is mixed with the two-thirds portion bypassed around the furnace and fed to the first sulfur converter. The amount of acid gas bypassing the furnace is controlled to maintain a ratio of H_2S to SO_2 slightly more than 2:1 to ensure proper material balance for the sulfur formation reactions.

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

$$H_2S + 0.5\ SO_2 \rightarrow 1.5\ S + H_2O$$

This reaction is catalyzed by a bauxite or alumina catalyst contained in the converter. The reaction is exothermic and results in a temperature rise in the gas flowing through the converter. As this reaction is limited by thermodynamic equilibrium it does not proceed to completion.

The sulfur produced in the converter, as well as the heat of reaction, are recovered by cooling the effluent below its sulfur dew point against boiler feedwater to produce 50 psig steam. The sulfur condenses and flows by gravity to a concrete sulfur sump. Sulfur, which is a solid at ambient temperature, is kept molten in the sump by condensing low pressure steam in pipe coils that cover the bottom of the sump.



PRELIMINARY
FOR ESTIMATING PURPOSES
ONLY

MATERIAL BALANCE		
STREAM NUMBER	4 ACID GAS	
COMPONENT	MPH	MOL.%
CH_4	0.5	0.18
C_2H_4	1.8	0.63
C_2H_6	9.7	3.41
H_2	0.1	0.03
CO	0.3	0.11
CO_2	208.7	73.41
H_2S	41.0	14.42
COS	—	—
N_2	0.3	0.11
H_2O	21.9	7.70
TOTAL MPH	284.3	100.00
TOTAL LB./HR.	11,343	
MOL. WT.	39.90	

NOTES:
1. THIS PROCESS FLOW DIAGRAM SHOWS ONE PARALLEL TRAIN OF A TOTAL OF THREE-50% PARALLEL TRAINS. (TWO OPERATING AND ONE SPARE)
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN CHECK	APP

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

FLUOR

DRAWN BY R.McCLEARY
CHECKED BY R.WHISNANT
SUPERVISOR R.RICE
RELEASE DATE
SUPERVISING ENGR COTTINGHAM
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
SULFUR PLANT
MOVING BED - AIR BLOWN - COMBINED CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE
DRAWING NUMBER MACW-23-1
REVISION 2

CN-53-1

FORM E-300 REV. 9-72 PRINTED IN U.S.A.
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The cooled gases from the condenser are reheated to reaction temperature by mixing with a hot stream of combustion gases drawn off the sulfur furnace before entering a second converter stage. In the second converter the sulfur reaction proceeds further. Again, the converter effluent is similarly cooled to condense sulfur which flows to the sulfur sump.

Tail Gas Treating Unit

The gas stream from the last converter, called tail gas, contains unreacted H_2S and SO_2. To meet the environmental emissions limits, the tail gas is processed further to remove these sulfur compounds.

The tail gas treating unit employs a proprietary design called Beavon Stretford process. This process is a modification of the Stretford process. The Stretford process is a widely accepted process for the removal of hydrogen sulfide as elemental sufur from effluent gas streams at atmospheric pressure. However, by itself this procss is not adequate for handling other sulfur compounds such as SO_2, COS and CS_2.

In the Beavon Stretford process all the sulfur compounds present in the effluent gas (from the Claus plant in this study) are converted to H_2S in a catalytic reactor according to the following hydrogenation and hydrolysis reactions:

$$SO_2 + 3H_2 \rightarrow H_2S + 2H_2O$$
$$S_8 + 8H_2 \rightarrow 8H_2S$$
$$COS + H_2O \rightarrow CO_2 + H_2S$$
$$CS_2 + 2H_2O \rightarrow CO_2 + 2H_2S$$
$$CO + H_2O \rightarrow CO_2 + H_2$$

The hydrogenation and hydrolysis reactions result in the conversion of essentially all sulfur compounds to hydrogen sulfide (H_2S) and a small residual - usually less than 50 parts per million by volume - of carbonyl sulfide and carbon disulfide. Removal of H_2S from the reactor effluents to produce elemental sulfur is then accomplished in a Stretford plant. The H_2S level in the spent tail gas leaving this unit is reduced to approximately 10 parts per million on volume basis.

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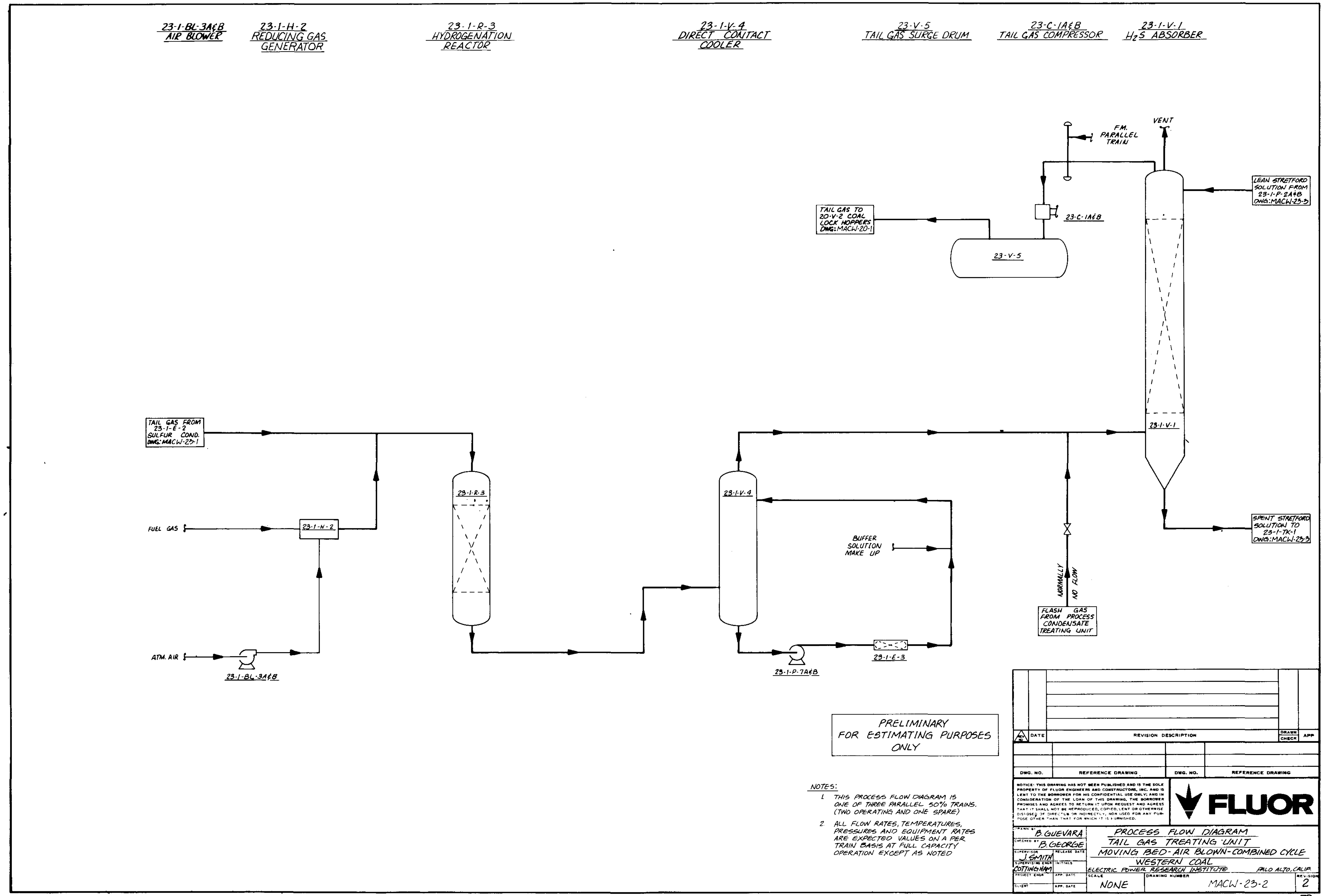

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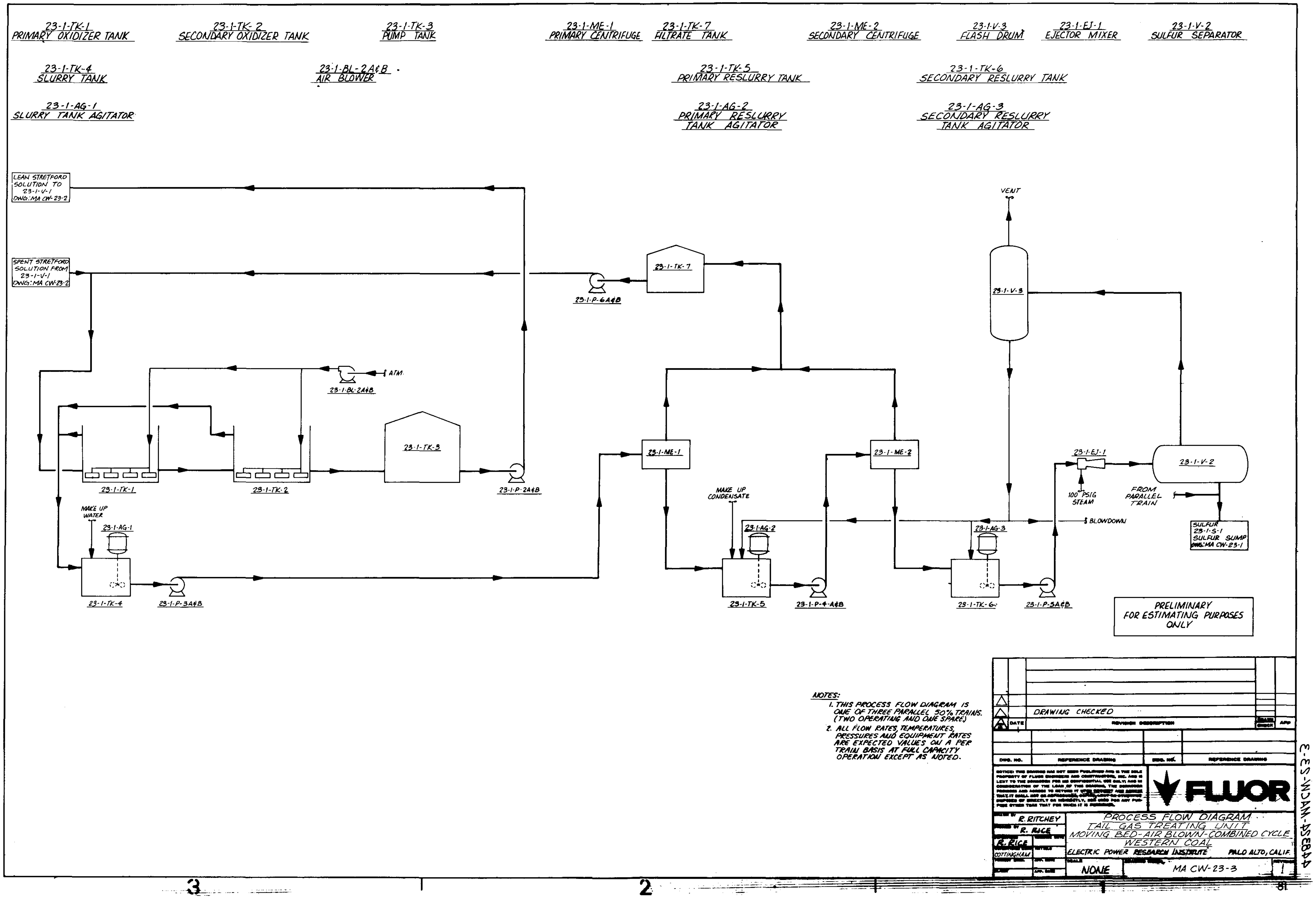

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Tail gas from the Claus plant (Flow Diagram: MACW-23-1) is heated to approximately 700°F by direct mixing of the combustion products from Reducing Gas Generator 23-1-H-2.

The hot tail gas flows through the hydrogenation reactor, 23-1-R-3, where SO_2, elemental sufur and other sulfur compounds are catalytically converted to H_2S. The effluent from 23-1-R-3 is cooled in a direct contact cooler, 23-1-V-4, and fed to the H_2S absorber, 23-1-V-1, where most of the H_2S is absorbed in the lean Stretford® solution. Active chemicals, sodium meta vanadate and anthra quinone disulfonic acid (ADA) in the solution oxidize absorbed H_2S to elemental sufur according the the following reactions:

$$2Na_2CO_3 + 2H_2S \rightarrow 2NaHCO_3 + 2NaHS$$
$$2NaHS + 2NaHCO_2 + 4NaVO_3 \rightarrow 2Na_2CO_3 + H_2O + S_2 + Na_2V_4O_9 + 2NaOH$$

The reduced vanadate is then oxidized to its original form by ADA in the solution. The reduced ADA is subsequently regenerated by air.

The absorber provides sufficient retention time to allow the reactions to elemental sulfur to go essentially to completion. The reacted solution flows from the bottom of the absorber to oxidizing tanks 23-TK-1 and 2, where it is regenerated with air sparged into the tanks. The air also provides a medium for floatation of the sulfur to the top of the oxidizers where it overflows into a slurry tank, 23-1-TK-4. The underflow from the oxidizers flows to a pump tank, 22-1-TK-3, from which it is pumped back to the absorber. Sulfur from the slurry tank is pumped to the primary centrifuge, 23-1-ME-1, which produces a wet sulfur cake that is reslurried and fed to a secondary centrifuge, 23-1-ME-2. The filtrate from the two centrifuges is recycled to the primary oxidizer through a filtrate tank, 22-1-TK-7. The sulfur cake from the secondary centrifuge is reslurried and pumped through an ejector mixer, 23-1-EJ-1, where sulfur is melted by the injection of steam. The molten sulfur is separated from the slurry medium (water) in a sulfur separator, 23-1-V-2. From the separator, sulfur is pumped to the loading facilities. The water portion is recycled to the reslurry tanks.

Equipment Notes

The Claus sulfur process is an established commercial process and consequently the equipment requirements are well known. Burning ammonia in the sulfur furnace

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is a recent development, but it has been demonstrated commercially. Tail gas treating units are a more recent development; however, the equipment has been operated in many commercial size plants successfully.

PROCESS CONDENSATE TREATING

Process Flow Diagrams MACW-24-1, MACW-24-2, MACW-24-3 and MACW-24-4 schematically represent the flow through the tar oil separation unit and phenol extraction plant, respectively, as described in applications to the Federal Power Commission by Transwestern Pipeline Company[1] and El Paso Natural Gas Company[2].

Process condensate from the quench scrubbers and gas cooling trains is processed through a tar oil separation unit which separates the condensate into the tar, oil and gas liquor fractions. There are three parallel 50 percent trains - two operating and one spare. The tar and oil fractions are pumped to a liquid fuels storage tank.

The gas liquor fractions from the two operating trains are combined and processed through a single train phenol extraction unit. Crude phenols recovered in this unit are pumped to the liquid fuels storage tank where they are combined with the tar and oil fractions. Storage capacity for 5 days at 100% plant capacity has been provided for phenolic water. This storage capacity is anticipated to cover any outage of the phenol extraction unit adequately. Naphtha recovered in the acid gas unit is also pumped to the fuels storage tank.

The liquid hydrocarbon by-products are burned in the by-product boilers, 30-1-B-1A or B, to generate supplementary 1,500 psig steam for the complex. The boilers are provided with facilities (electrostatic precipitators and stack gas treating units) to meet the particulate and sulfur emission standards required in the Chicago area.

The dephenolized condensate from the phenol extraction unit is fed to an acid gas and ammonia recovery system based on U.S. Steel's PHOSAM-W process. Surge capacity is provided for air cooled, flashed condensate to provide adequate

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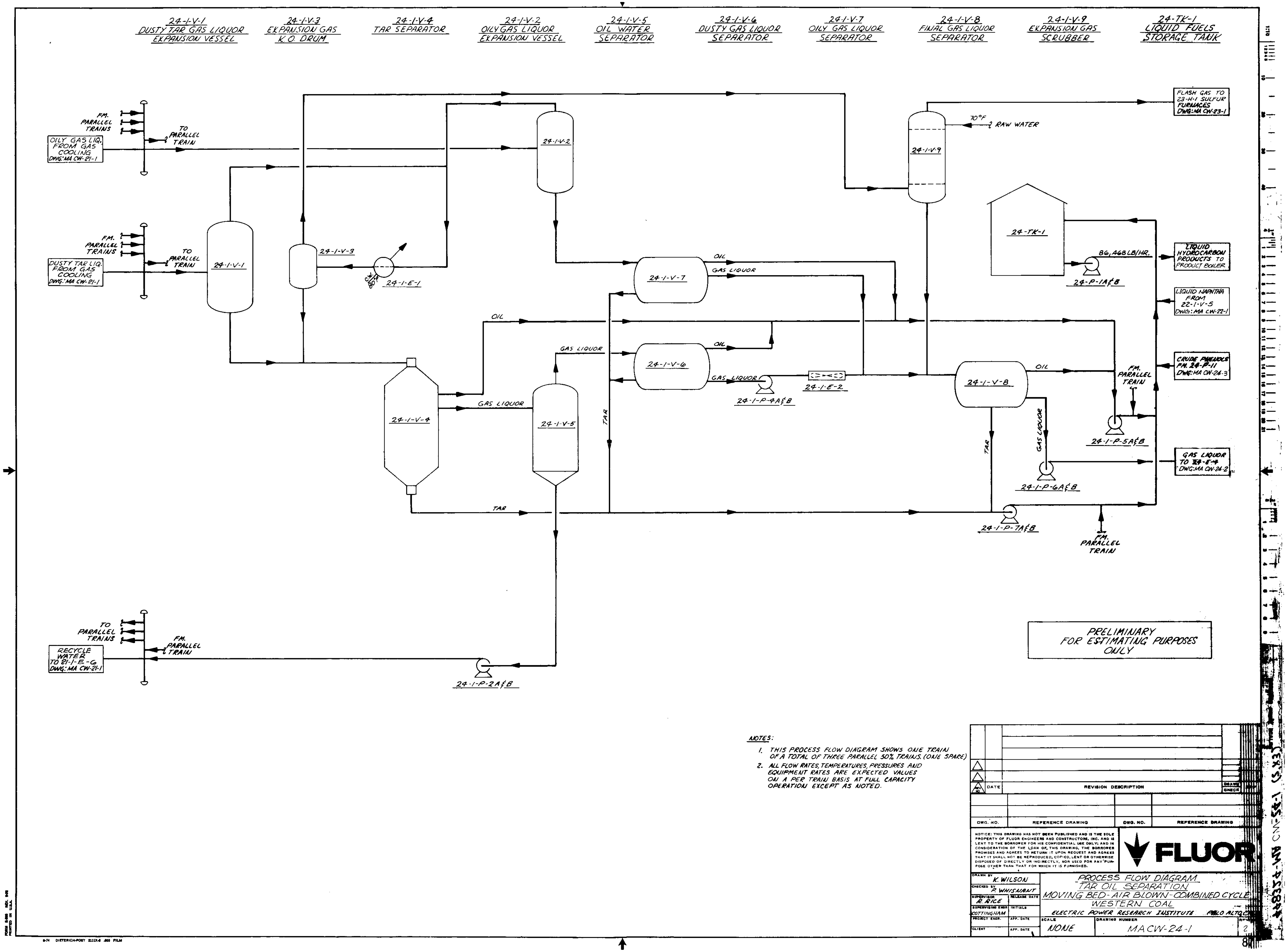
1. Transwestern Pipeline Company, et al., "Ammended Application for Certificate of Public Convenience and Necessity," Federal Power Commission Docket No. CP73-211.

2. El Paso Natural Gas Company, "Second Supplement to Application of El Paso Natural Gas Company for a Certificate of Public Convenience and Necessity," Federal Power Commission Docket No. CP73-131.



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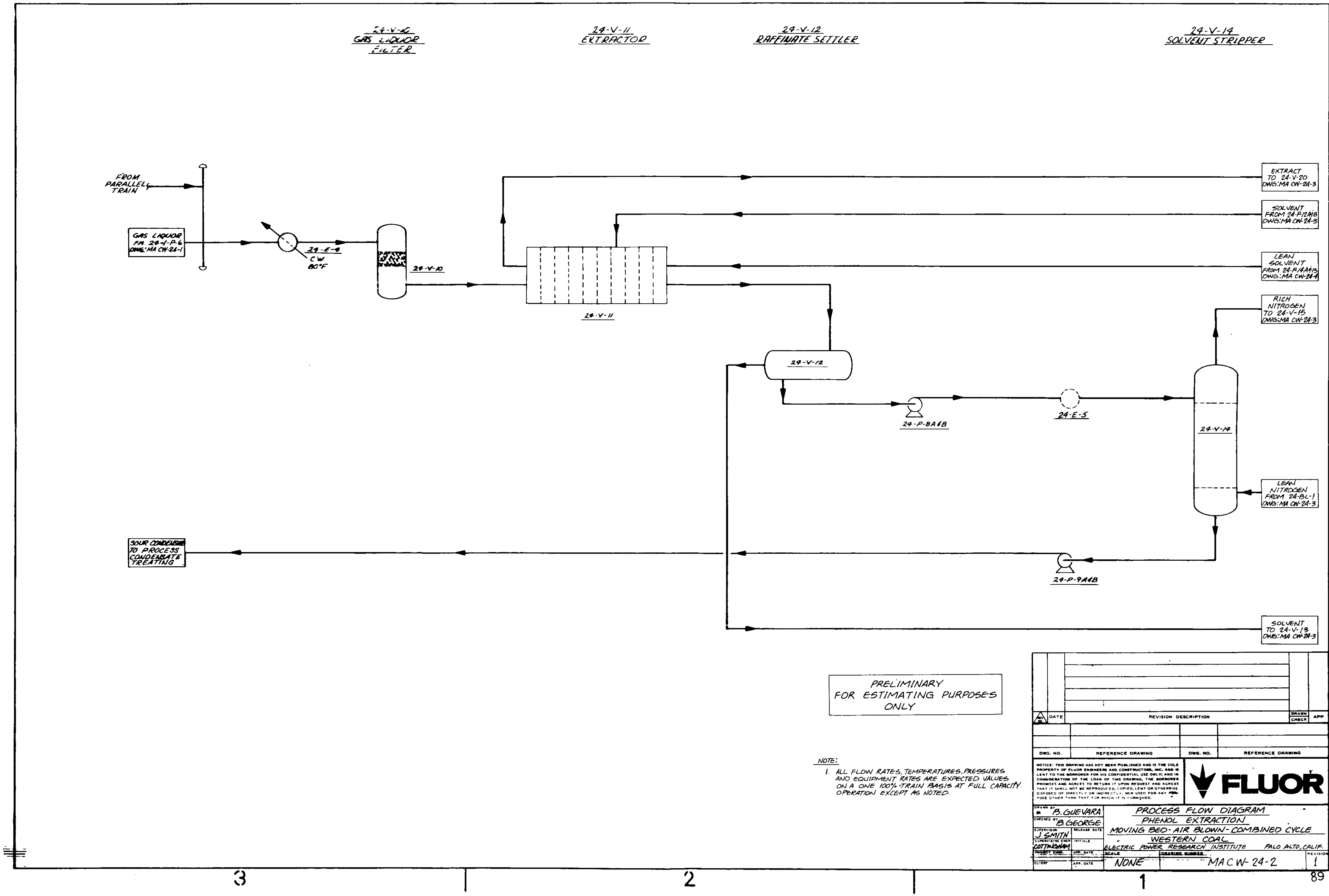

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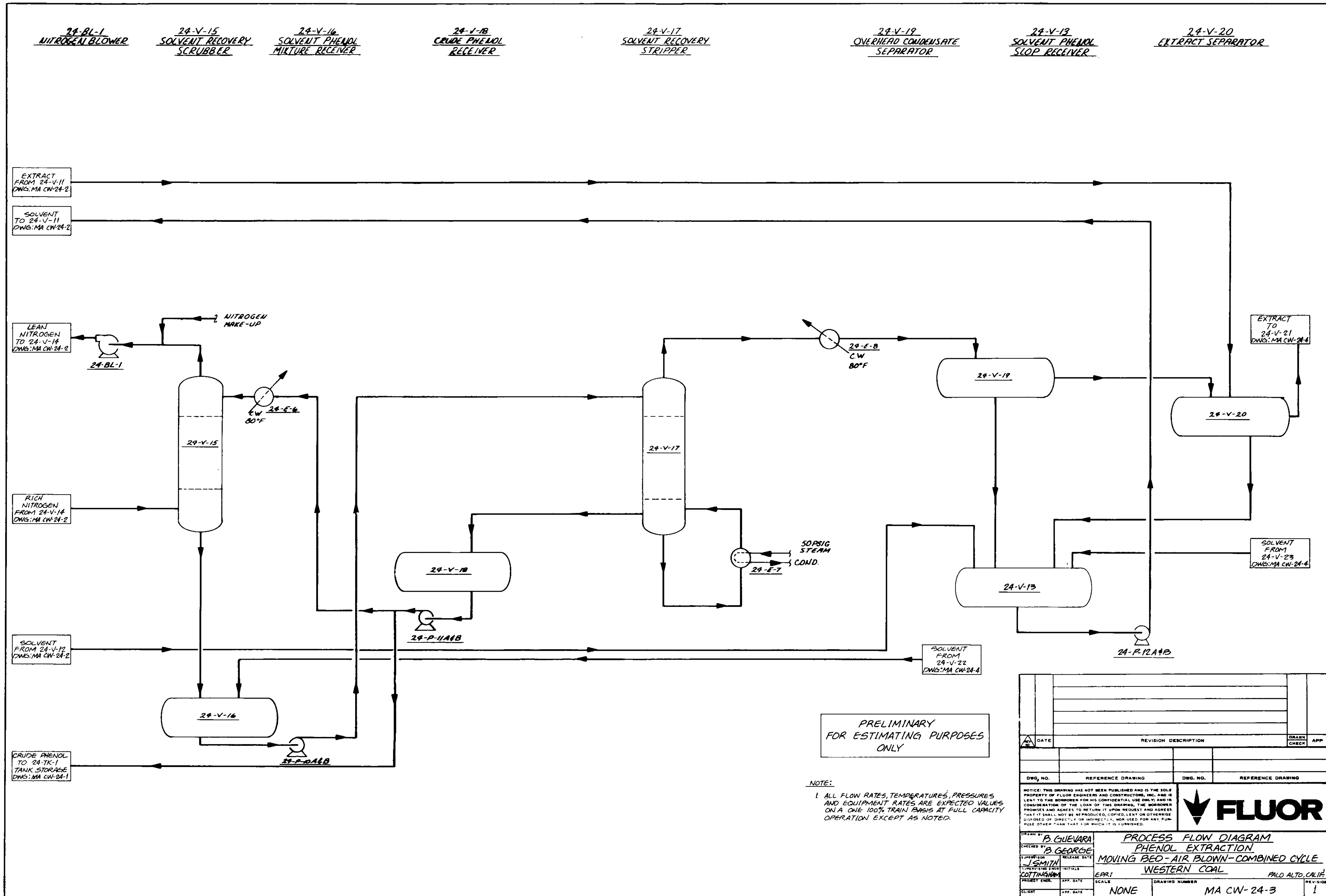

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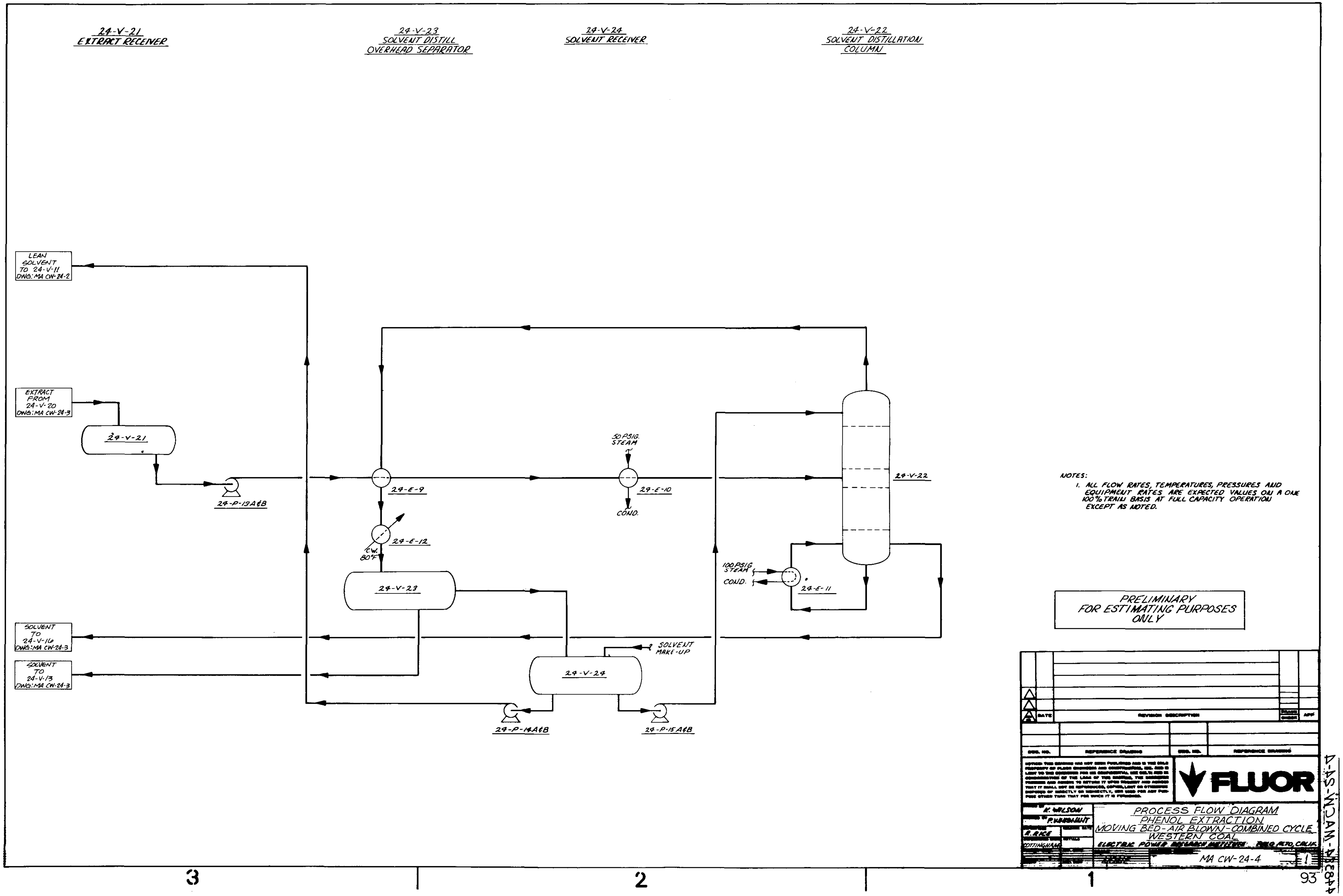

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storage during unscheduled shutdowns. Acid gases, which contain a small amount of ammonia, that are recovered from the process condensate are routed to the sulfur furnace in the sulfur recovery unit. A 99.99% pure anhydrous liquid ammonia stream of 128 tons per day, suitable for fertilizer and commercial uses, is recovered. Liquid ammonia storage and loading facilities are provided. Process condensate after acid gas and ammonia recovery is treated in a biological treating unit and then reused as cooling tower makeup water.

Equipment Notes

The tar oil separation and phenol extraction units are based on Lurgi's commercially proven processes. Similar units have been successfully operated for several years in Sasol's coal based gas and oil plant at Sasolburg, South Africa.

The PHOSAM-W is a widely accepted process to reclaim anhydrous ammonia from the sour process condensate.

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STEAM, BOILER FEEDWATER AND CONDENSATE

Process Flow Diagram MACW-30-1 schematically represents steam, boiler feedwater and condensate systems.

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

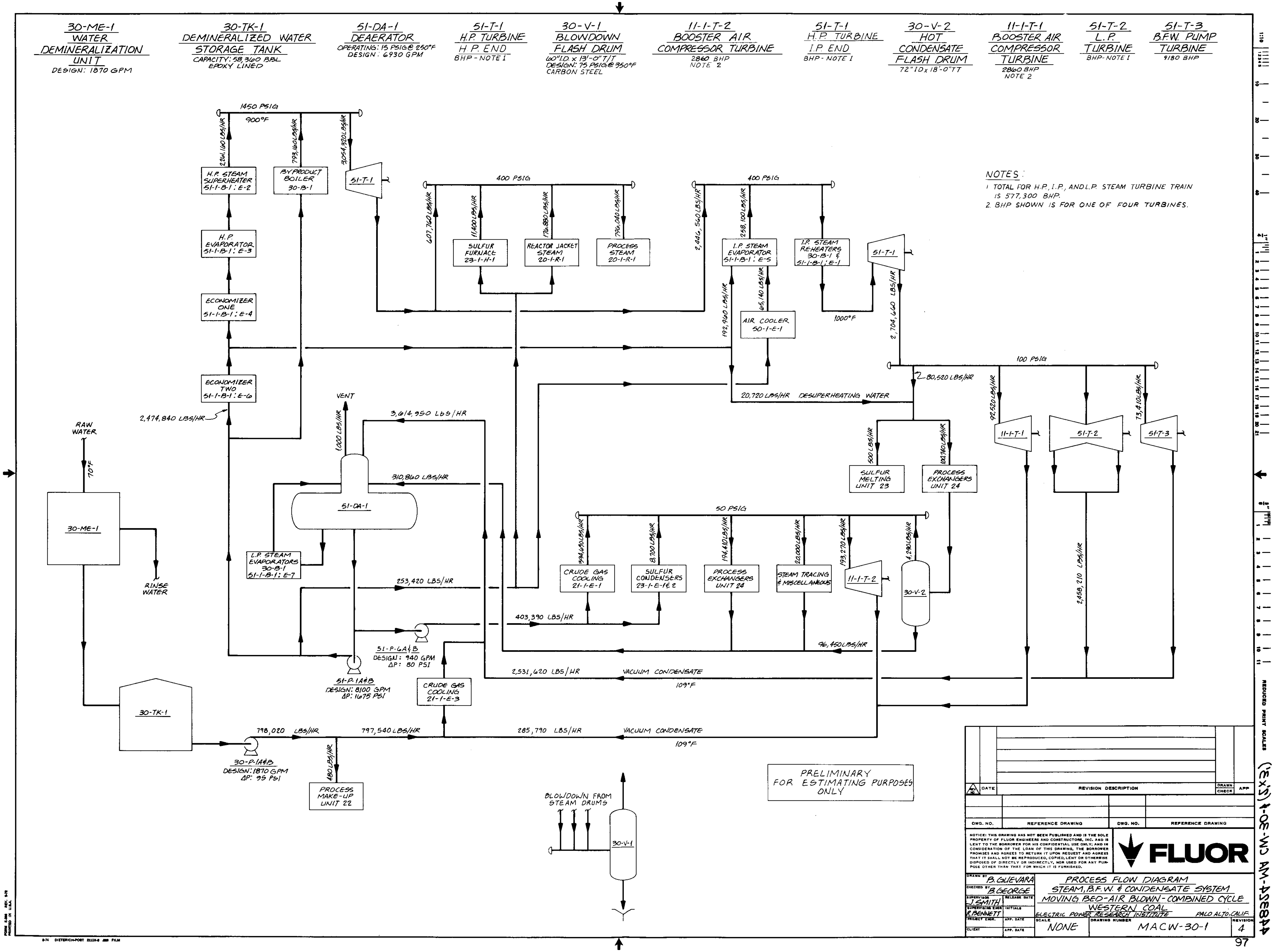
High Pressure	-	1450 psig, 900°F
Intermediate Pressure	-	400 psig
Low Pressure	-	100 psig
Low Pressure	-	50 psig

Major high pressure steam generation is carried out in the heat recovery steam generator (HRSG) of Gas Turbine 50-1-GT-1. There are six gas turbines and each gas turbine has its own HRSG. The HRSG is described in detail in Appendix A. Additional generation of high pressure superheated steam is obtained in the by-product boilers, 30-1-B-1A or B. All high pressure superheated steam is used to drive the steam turbine, 51-T-1.

The high pressure (HP) end of the turbine, 51-T-1, exhausts steam at 400 psig. Additional 400 psig steam generation is obtained in the sulfur furnace (23-1-H-1), the gasifiers' jackets (20-1-R-1) and the HRSGs (51-1-B-1:E-5). Part of the 400 psig steam flows to the gasification area to meet the process steam demand of the gasifiers, 20-1-R-1. The balance is superheated to 1000°F in the HRSGs and the by-product boilers. The superheated intermediate pressure (IP) steam is then used in the IP end of turbine 51-T-1, which exhausts at 115 psig.

A portion of the 100 psig exhaust steam is desuperheated and supplied to the process exchangers in Unit 24. A small quantity of desuperheated steam is also used for sulfur melting in Unit 23. The balance of the 100 psig exhaust steam is used in the condensing turbines, 11-1-T-1, 51-T-2 and 51-T-3.

The 50 psig steam header is supplied by the steam generated in the process exchangers, 21-1-E-1 and 23-1-E-1 and E-2. Hot condensate from the 100 psig steam users in Unit 24 is flashed in 30-V-2 at 50 psig and the flashed steam flows to the 50 psig header. The 50 psig steam is mainly used for the process exchangers in Unit 24, booster air compressor turbines, 11-1-T-1, and steam tracing and


30-ME-1 WATER DEMINERALIZATION UNIT
DESIGN: 1870 GPM
30-TK-1 DEMINERALIZED WATER STORAGE TANK
CAPACITY: 58,360 BBL EPOXY LINED
51-DA-1 DEAERATOR
OPERATING: 15 PSIG@ 250°F
DESIGN: 6930 GPM
51-T-1 H.P. TURBINE H.P. END
BHP-NOTE 1
30-V-1 BLOWDOWN FLASH DRUM
60"I.D. x 13'-0" T/T
DESIGN: 75 PSIG@ 350°F
CARBON STEEL
11-1-T-2 BOOSTER AIR COMPRESSOR TURBINE
2860 BHP
NOTE 2
51-T-1 H.P. TURBINE I.P. END
BHP-NOTE 1
30-V-2 HOT CONDENSATE FLASH DRUM
72" I.D x 18'-0" TT
11-1-T-1 BOOSTER AIR COMPRESSOR TURBINE
2860 BHP
NOTE 2
51-T-2 L.P. TURBINE
BHP-NOTE 1
51-T-3 B.F.W. PUMP TURBINE
9180 BHP
NOTES:
1 TOTAL FOR H.P., I.P., AND L.P. STEAM TURBINE TRAIN IS 577,300 BHP.
2. BHP SHOWN IS FOR ONE OF FOUR TURBINES.
1450 PSIG
900°F
2,641,160 LBS/HR
793,160 LBS/HR
3,054,320 LBS/HR
H.P. STEAM SUPERHEATER 51-1-B-1: E-2
BYPRODUCT BOILER 30-B-1
51-T-1
H.P. EVAPORATOR 51-1-B-1: E-3
ECONOMIZER ONE 51-1-B-1: E-4
ECONOMIZER TWO 51-1-B-1: E-6
2,474,840 LBS/HR
400 PSIG
607,760 LBS/HR
11,400 LBS/HR
176,880 LBS/HR
796,040 LBS/HR
SULFUR FURNACE 23-1-H-1
REACTOR JACKET STEAM 20-1-R-1
PROCESS STEAM 20-1-R-1
400 PSIG
2,446,540 LBS/HR
258,100 LBS/HR
I.P. STEAM EVAPORATOR 51-1-B-1: E-5
I.P. STEAM REHEATERS 30-B-1 & 51-1-B-1: E-1
51-T-1
192,960 LBS/HR
65,140 LBS/HR
AIR COOLER 50-1-E-1
1000°F
2,704,660 LBS/HR
100 PSIG
80,520 LBS/HR
92,520 LBS/HR
73,410 LBS/HR
20,720 LBS/HR DESUPERHEATING WATER
11-1-T-1
51-T-2
51-T-3
500 LBS/HR
100,740 LBS/HR
SULFUR MELTING UNIT 23
PROCESS EXCHANGERS UNIT 24
VENT
1,000 LBS/HR
3,614,950 LBS/HR
310,860 LBS/HR
51-DA-1
RAW WATER
70°F
30-ME-1
RINSE WATER
L.P. STEAM EVAPORATORS 30-B-1 51-1-B-1: E-7
253,420 LBS/HR
50 PSIG
394,640 LBS/HR
8,700 LBS/HR
194,410 LBS/HR
20,000 LBS/HR
193,270 LBS/HR
4,290 LBS/HR
CRUDE GAS COOLING 21-1-E-1
SULFUR CONDENSERS 23-1-E-1 & 2
PROCESS EXCHANGERS UNIT 24
STEAM TRACING & MISCELLANEOUS
11-1-T-2
30-V-2
2,458,210 LBS/HR
403,390 LBS/HR
96,450 LBS/HR
51-P-6A&B
DESIGN: 940 GPM
ΔP: 80 PSI
51-P-1A&B
DESIGN: 8100 GPM
ΔP: 1675 PSI
CRUDE GAS COOLING 21-1-E-3
2,531,620 LBS/HR
VACUUM CONDENSATE
109°F
30-TK-1
798,020 LBS/HR
797,540 LBS/HR
285,790 LBS/HR
VACUUM CONDENSATE
109°F
30-P-1A&B
DESIGN: 1870 GPM
ΔP: 95 PSI
480 LBS/HR
PROCESS MAKE-UP UNIT 22
BLOWDOWN FROM STEAM DRUMS
30-V-1
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
DATE
REVISION DESCRIPTION
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FLUOR
DRAWN BY B. GUEVARA
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
R. BENNETT
PROCESS FLOW DIAGRAM
STEAM, B.F.W. & CONDENSATE SYSTEM
MOVING BED-AIR BLOWN-COMBINED CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE
PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER MACW-30-1
REVISION 4
97
AM-30-4 (S,X,E)
44-788-E

other miscellaneous users. The condensate from the 50 psig steam users and 30-V-2 flows to the deaerator, 51-DA-1.

Raw water is treated in a semiautomatic, resin bed demineralization unit 30-ME-1 to produce demineralized water suitable for 1500 psig boiler. Demineralized water is stored in a tank, 30-TK-1. Storage equivalent to 24 hours of demineralized water production is provided. The demineralized water requirement is estimated at approximately 1596 gpm. Demineralized water from the storage tank is transported to the deaerator through Pumps 30-1-P-1A and B. A small quantity of the makeup water is withdrawn from the discharge of 30-1-P-1A and B transported to Unit 22. The balance of the demineralized water combines with the vacuum condensate from the turbines, 11-1-T-1 and 11-1-T-2, and flows through 21-1-E-3 to recover heat from the crude gas in the gas cooling unit (Flow Diagram: MACW-21-1). The hot condensate and demineralized water from the exchanger, 21-1-E-3, then combines with the condensate from the turbines, 51-T-2 and 51-T-3, and flows to the deaerator, 51-DA-1. The deaerator is a tray type unit operating at 28 psia. The deaerator provides 10 minute storage.

Boiler feedwater from the deaerator is pumped through high pressure boiler feedwater pumps (51-P-1A and B) to the high pressure and intermediate pressure evaporators of the HRSGs and the by-product boilers, 30-B-1A and B. Boiler feedwater is supplied to the low pressure (50 psig) steam generators by a separate set of pumps, 51-P-6A and B.

COMBINED CYCLE SYSTEM

Process Flow Diagram MACW-50/51-1 depicts the combined cycle system for Case MACW. The diagram shows the total power block flows.

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

One distinct feature of Case MACW is a provision for by-product boilers, 30-B-1A,B (one operating, one spare) to burn by-product tar, oil, phenol and naphtha produced in the process. High pressure (HP) superheated steam at 1,450 psig, 900°F produced in the by-product boilers is used in 51-T-1 to generate additional power.

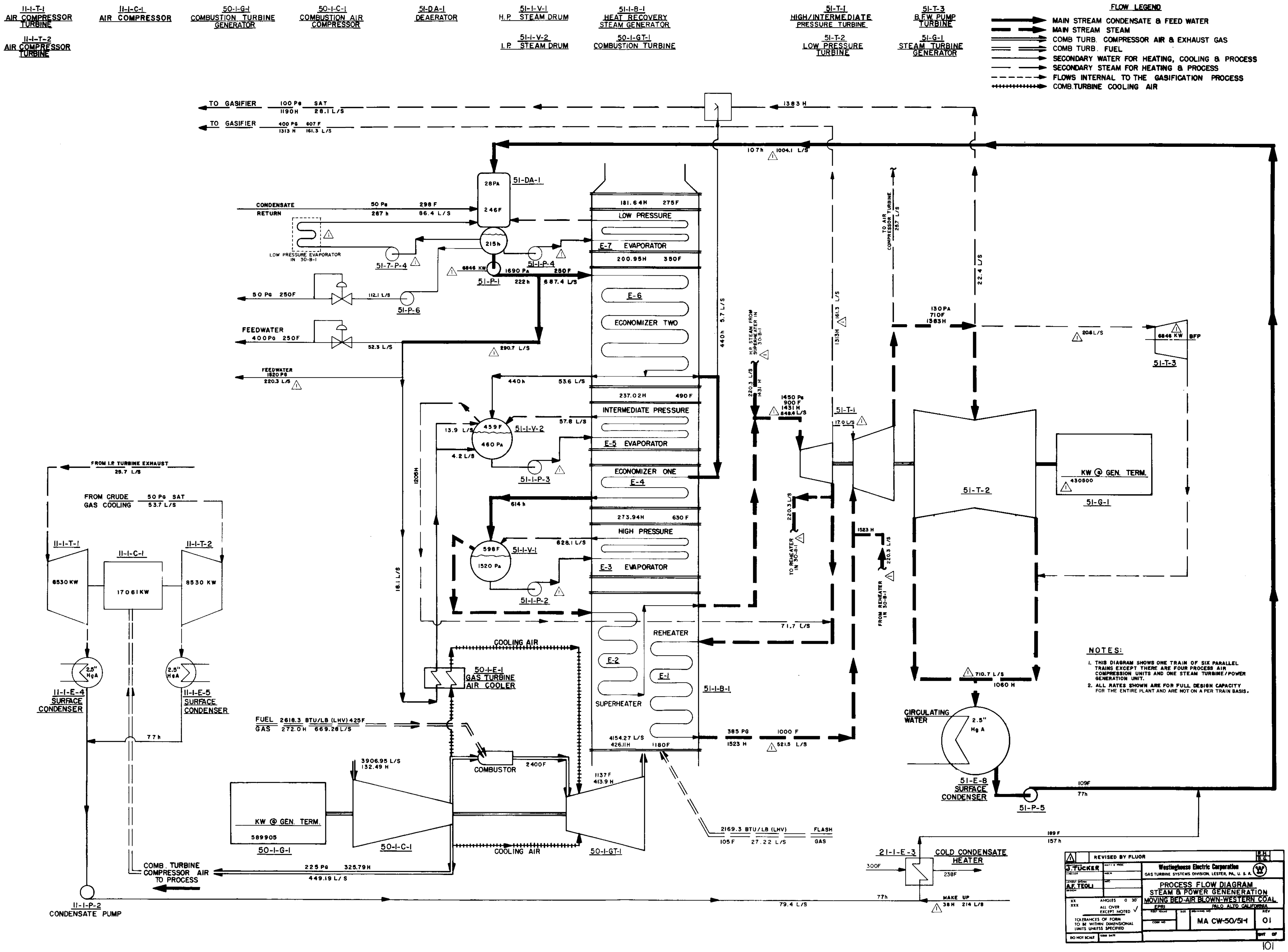
The by-product boiler design is similar to that of the HRSG. The following heating coils provided in 30-B-1 are listed in the direction of flue gas flow.

- IP steam reheater
- HP steam superheater
- Economizer
- LP evaporator

The by-product boiler supplies approximately 26 percent of the total superheated HP steam generated in the plant. A portion of the exhaust steam from the HP end of Turbine 51-T-1 is reheated to 1,000°F in the IP steam reheater of 30-B-1. The LP evaporator supplies a part of the deaerating steam.

The by-product boilers are provided with electrostatic precipitators and stack gas clean-up facilities to meet the environmental standard of 1.2 lbs SO_2/MM Btu on the stack gas.

Steam for the boiler feed water pump drive is taken from the steam cross over from the IP to LP steam turbine cylinders. One-half of the process air compressor power is supplied by the saturated 50 psig steam available from the


11-1-T-1 AIR COMPRESSOR TURBINE
11-1-T-2 AIR COMPRESSOR TURBINE
11-1-C-1 AIR COMPRESSOR
50-1-G-1 COMBUSTION TURBINE GENERATOR
50-1-C-1 COMBUSTION AIR COMPRESSOR
51-DA-1 DEAERATOR
51-1-V-1 H.P. STEAM DRUM
51-1-V-2 I.P. STEAM DRUM
51-1-B-1 HEAT RECOVERY STEAM GENERATOR
50-1-GT-1 COMBUSTION TURBINE
51-T-1 HIGH/INTERMEDIATE PRESSURE TURBINE
51-T-2 LOW PRESSURE TURBINE
51-T-3 B.F.W. PUMP TURBINE
51-G-1 STEAM TURBINE GENERATOR
FLOW LEGEND
MAIN STREAM CONDENSATE & FEED WATER
MAIN STREAM STEAM
COMB TURB. COMPRESSOR AIR & EXHAUST GAS
COMB TURB. FUEL
SECONDARY WATER FOR HEATING, COOLING & PROCESS
SECONDARY STEAM FOR HEATING & PROCESS
FLOWS INTERNAL TO THE GASIFICATION PROCESS
COMB.TURBINE COOLING AIR
TO GASIFIER 100 Pa SAT 1190H 28.1 L/S
TO GASIFIER 400 Pa 607 F 1313 H 161.3 L/S
1383 H
107 h 1004.1 L/S
CONDENSATE RETURN 50 Pa 287 h 298 F 86.4 L/S
28 PA 246 F
215h
LOW PRESSURE EVAPORATOR IN 30-B-1
51-7-P-4
51-1-P-4
6846 KW 51-P-1 1690 Pa 222h 250 F 687.4 L/S
50 Pa 250F
112.1 L/S
51-P-6
FEEDWATER 400 Pa 250F
52.3 L/S
290.7 L/S
FEEDWATER 1520 Pa 220.3 L/S
181.64H 275F
LOW PRESSURE
E-7 EVAPORATOR
200.95H 380F
E-6
ECONOMIZER TWO
440h 53.6 L/S
237.02H 490F
INTERMEDIATE PRESSURE
E-5 EVAPORATOR
ECONOMIZER ONE
E-4
273.94H 630 F
HIGH PRESSURE
E-3 EVAPORATOR
13.9 L/S 459 F 460 Pa 51-1-V-2
57.8 L/S
4.2 L/S
51-1-P-3
614 h
598F 1520 Pa 51-1-V-1
628.1 L/S
51-1-P-2
1206H
18.1 L/S
440h 5.7 L/S
H.P. STEAM FROM SUPERHEATER IN 30-B-1
220.3 L/S 1431 H
1313H 161.3 L/S
1450 Pa 900 F 1431 H 6484 L/S
TO AIR COMPRESSOR TURBINE 25.7 L/S
22.4 L/S
130 Pa 710F 1383H
208 L/S
6846 KW BFP
51-T-3
51-T-1
170 L/S
51-T-2
KW @ GEN. TERM. 430500
51-G-1
220.3 L/S TO REHEATER IN 30-B-1
1523 H
220.3 L/S
FROM REHEATER IN 30-B-1
71.7 L/S
REHEATER
E-2
E-1
SUPERHEATER
51-1-B-1
4154.27 L/S 426.11H 1180F
385 Pa 1523 H 1000 F 521.5 L/S
710.7 L/S 1060 H
NOTES:
1. THIS DIAGRAM SHOWS ONE TRAIN OF SIX PARALLEL TRAINS EXCEPT THERE ARE FOUR PROCESS AIR COMPRESSION UNITS AND ONE STEAM TURBINE/POWER GENERATION UNIT.
2. ALL RATES SHOWN ARE FOR FULL DESIGN CAPACITY FOR THE ENTIRE PLANT AND ARE NOT ON A PER TRAIN BASIS.
CIRCULATING WATER
2.5" Hg A
51-E-8 SURFACE CONDENSER
109F 77h
51-P-5
FROM I.P. TURBINE EXHAUST 25.7 L/S
FROM CRUDE GAS COOLING 50 Pa SAT 53.7 L/S
11-1-T-1 8530 KW
11-1-C-1 17061 KW
11-1-T-2 8530 KW
2.5" HgA
11-1-E-4 SURFACE CONDENSER
11-1-E-5 SURFACE CONDENSER
77h
COOLING AIR
50-1-E-1 GAS TURBINE AIR COOLER
FUEL GAS 2618.3 BTU/LB (LHV) 425F 272.0H 669.26 L/S
3906.95 L/S 132.49 H
COMBUSTOR 2400F
1137 F 413.9 H
KW @ GEN. TERM. 589905
50-1-G-1
50-1-C-1
COOLING AIR
50-1-GT-1
2169.3 BTU/LB (LHV) 105F 27.22 L/S FLASH GAS
COMB. TURBINE COMPRESSOR AIR TO PROCESS 225 Pa 325.79H 449.19 L/S
11-1-P-2 CONDENSATE PUMP
79.4 L/S
21-1-E-3 COLD CONDENSATE HEATER
300F 238F
189 F 157 h
77h
MAKE UP 38H 214 L/S
REVISED BY FLUOR
J. TUCKER
A.F. TEOLI
Westinghouse Electric Corporation
GAS TURBINE SYSTEMS DIVISION, LESTER, PA, U.S.A.
PROCESS FLOW DIAGRAM
STEAM & POWER GENENERATION
MOVING BED-AIR BLOWN-WESTERN COAL
EPRI PALO ALTO CALIFORNIA
MA CW-50/51-1
01
101

process. The balance of the air compressor horsepower is provided by the steam from the crossover line between the IP and LP steam turbine cylinders.

Process cooling loads are integrated where possible, into the condensate and makeup water system. Approximately 114.5 MM Btu/hr of low level heat from process gas cooling is recovered by heating the cold demineralized water and vacuum condensate in exchanger 21-1-E-3 (Flow Diagram: MACW-21-1).

Equipment Notes

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

PROCESS DISCUSSION

The table below summarizes pertinent heat and material balance results.

TABLE M-2

SUMMARY OF SYSTEM PERFORMANCE - CASE MACW

GASIFICATION AND GAS CLEANING SYSTEM	
Coal Feed Rate, lbs/hr (m.f.)	1,014,814
Oxygen or Air (1)/Coal Ratio, lbs/lb m.f.	1.562
Oxidant Temperature, °F	340
Steam/Coal Ratio, lbs/lb m.f.	0.758
Gasifier Exit Pressure, psig	340
Crude Gas Temperature, °F	861
Crude Gas HHV (dry basis). Btu/SCF (2)	189.1
Temperature of Fuel Gas to Gas Turbine, °F	425
POWER SYSTEM	
Gas Turbine Inlet Temperature, °F	2400
Pressure Ratio	17:1
Turbine Exhaust Temperature, °F	1137
Steam Conditions, psig/°F/°F	1450/900/1000
Condenser Pressure, inches Hg abs.	2.5
Stack Temperature, °F	275
Gas Turbine Power, MW (3)	590
Steam Turbine Power, MW (3)	430
Power Consumed, MW	32
Net System Power, MW	988
OVERALL SYSTEM	
Process and Deaerator Makeup Water, gpm/1000 MW	2207
Cooling Tower Makeup Water, gpm/1000 MW	5698
Cooling Water Circulation Rate, gpm/MW	366
Cooling Tower Heat Rejection, % of Coal HHV	33.9
Air Cooler Heat Rejection, % of Coal HHV	6.5
Net Heat Rate, Btu/kWh	9762
Overall System Efficiency (Coal→Power), % of Coal HHV	34.96

NOTES

(1) Dry basis, 100% O_2 for oxygen blown
(2) Excluding HHV of H_2S, COS and NH_3
(3) At generator terminals

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Gasifier Material Balance

Table M-3 presents the gasifier material balance for Case MACW.

This case is based on using western coal in Lurgi type gasifiers. Western coal is particularly suitable for the Lurgi process, which is a widely accepted process which has been commercially proven for the gasification of low to moderately caking coal. The process design is based on feeding approximately 12,178 tons per day of moisture free (m.f.) coal to the gasification units. It is important to keep in mind the fact that all other cases studied have been based on an Illinois #6 coal. As the ash and moisture contents of the western coal are significantly higher than in Illinois No. 6 coal, consumption of the bulk coal in the plant in Case MACW is 13,900 tons per day as compared to the coal consumption rate of 10,000 tons per day in other cases.

The by-product liquid hydrocarbons produced in the gasifiers are recovered in the tar oil separation and phenol extraction units. The recovered by-products are burned in the by-product boilers (equipped with stack gas scrubbers and electrostatic precipitators) to generate high pressure steam which is integrated with the main steam power systems. In an actual design, it would be necessary to evaluate this approach compared to providing capability in the HRSG for auxiliary firing of some of these by-products. These liquids (particularly the heavy tars) contain sulfur (Table M-3) and particulates; providing necessary modifications to the HRSG to handle such fuels could prove expensive. Such a system would have the advantage, however, of reducing the cost of auxiliary boilers, if the naphthas, oils and phenols were fired to the HRSG.

Other by-products produced in the overall plant are ammonia and elemental sulfur with production rates of 128 ton/day and 63 ton/day respectively.

TABLE M-3

MATERIAL BALANCE - CASE MACW

FEEDS

	T (°F)	lb/hr	lb mol/hr
Coal	77		
Moisture		143,633	7,973.4
Ash		296,533	
MAF Coal			
Carbon		550,227	45,809.9
Hydrogen		41,066	20,372.1
Oxygen		106,368	3,324.1
Nitrogen		9,696	346.1
Sulfur		10,565	329.5
Halogen		359	
TOTAL COAL		1,158,447	
Oxidant	340		
Oxygen		369,244	11,539.30
Nitrogen		1,216,408	43,422.50
Water		14,676	814.60
TOTAL OXIDANT		1,600,328	55,776.40
Steam			
Jacket	432	176,880	9,817.9
Makeup	590	592,160	32,866.3
TOTAL STEAM		769,040	42,684.2
TOTAL FEEDS		3,527,815	

EFFLUENTS

	psig	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	340	861			
CH_4			81,498	5,080.0	3.68
C_2H_4			7,010	250.0	0.18
C_2H_6			11,416	379.6	0.27
H_2			48,770	24,193.6	17.52
CO			423,069	15,104.0	10.94
CO_2			747,160	16,977.20	12.29
H_2S			10,535	309.2	0.22
COS			-	-	-
NH_3			10,701	628.4	0.45
N_2			1,216,408	43,422.40	31.44
H_2O			572,605	31,784.4	23.01
Subtotal			3,129,172	138,128.8	100.00
N+T+O(1)			81,153		
P+O(2)			5,315		
TOTAL GASIFIER EFFLUENT			3,215,640		
Ash		900			
Carbon			15,642		
Ash			296,533		
Subtotal			312,175		
TOTAL EFFLUENTS			3,527,815		

NOTES:

	(1) Naphtha, Tars, Oils Wt%	(2) Phenols, Others Wt%
Carbon	85.7	75.1
Hydrogen	6.5	6.5
Oxygen	5.9	18.0
Nitrogen	1.1	0.1
Sulfur	0.8	0.3
	100.0	100.0

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TABLE M-4

ENERGY BALANCE - CASE MACW

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

	MM BTU/HR					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	9,644	7				9,651
Compressor Suction Air		85	205			290
Demineralized Water		8				8
Auxiliary Power Inputs					111	111
TOTAL	9,644	100	205	0	111	10,060
HEAT OUT						
Ash	226	28				254
Gasifier Heat Loss				88		88
Sulfur Product	21					21
Ammonia Product	85					85
Generated Power					3,483	3,483
Power Block Losses (1)				37	114	151
Oxidant Compressor Surface Condensers			281			281
Power Surface Condensers			2,515			2,515
HRSG Stack Loss		814	898			1,712
Gasifier Effluent Cooling			506			506
Oxidant Compressor Cooling		14				14
Selexol Solvent Cooling		177				177
Regenerator Overhead Cooling		113	29			142
Process Condensate Cooling		65				65
Steam Heat Losses		18	2			20
Process Condensate Treating Unit			351			351
Waste Water Effluent		24				24
By-Product Boiler Stack Loss		113	74			187
Spent Tail Gas Loss		10	8			18
TOTAL	332	1,370	4,664	125	3,597	10,094

$$\frac{\text{Output - Input}}{\text{Input}} = 0.34\%$$

(1) Includes mechanical and electrical losses.

TABLE M-5

ENERGY BALANCE AS PERCENT COAL HHV - CASE MACW

	MM Btu/hr	Percent
IN		
Coal HHV	9,644	100.0
OUT		
Net Power	3,372	34.96
Sulfur Product, HHV	21	0.22
Ammonia Product, HHV	85	0.88
Selexol Sensible and Latent	319	3.31
Oxidant Compressor Cooling	14	0.15
Ash	254	2.63
HRSG Stack Gases	1712	17.75
Rejected at Condensers	2796	28.99
Other Sensible Losses	17	0.18
Other Latent Losses	662	6.86
Gasifier Heat Losses	88	0.91
Power Block Losses	151	1.56
By-Product Boiler Stack Loss	187	1.94
	9,678	100.34

Acid Gas Removal

Comments made for Case EXTC are applicable to this case as well.

Because this case employs a Lurgi gasification system, and naphtha and other oil products result, the Selexol® system has been modified slightly. It is expected that small quantities of naphtha will enter the Selexol® unit in the cooled gas stream. Higher hydrocarbons (C_3 plus) cause problems in downstream Claus plants as they do not burn completely. Instead the higher hydrocarbons undergo partial cracking resulting in carbon deposition on the sulfur converter catalyst and production of black sulfur. To prevent this material from building up in the Selexol® solvent (naphtha is soluble in Selexol), and/or passing through to the Claus plant, a separate absorber (Flow Diagram MACW-22-1) has been included, using a slipstream of the lean solution to absorb naphtha from the product gas. Naphtha is subsequently forced out of solution by mixing with water, and transported to the liquid hydrocarbon storage provided in Unit 24.

Process Energy Balances

Table M-4 presents the overall process energy balance at 100 percent capacity operation. The boundary for the balance encompasses the entire plant. Energy content of streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F, and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat energy at 3,413 Btu/kWh. These energy balances close to less than 0.5 percent. The discrepancies result from approximations used for some process units and for calculating some heat loads.

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

As shown in Table M-4, total power generation of 3,483 MM Btu/hr is obtained in this case. Coal charge rate based on high heating value (HHV) is 9,644 MM Btu/hr. This results in a conversion of 36.12 percent on a gross basis.

As shown in Table M-2, if all the power consumed in the plant is included, the system cold efficiency (net power at 3,413 Btu/kWh), as a percent of coal HHV, is 34.96 percent. The heat rate based on net power produced and coal HHV input is 9,762 Btu/kWh.

In the acid gas removal unit for this case, as previously discussed, a naphtha absorber has been included. There is a small energy penalty for this operation because some lean Selexol solvent is used in the naphtha absorber, resulting in slightly higher circulation rates.

The major energy losses from the plant are in the surface condensers and with the HRSG stack gases. These losses amount to 28.99 percent and 17.75 percent of the coal HHV respectively.

The liquid hydrocarbon by-products are combusted in by-product boilers to generate high pressure superheated and reheated steam resulting in a loss of 1.94 percent of coal HHV with the boiler stack and reheated gases. The steam from the by-product boiler has been added to the other plant steam and used to generate power in the steam turbine generator.

ECONOMICS

Important economic results are summarized below.

TABLE M-6

SUMMARY OF ECONOMICS - CASE MACW

NET PRODUCTION AT DESIGN CAPACITY	Case MACW
Net Power, MW (1)	987.8
Overall Plant Heat Rate, Btu/kWh	9,762
TOTAL CAPITAL (2)	
Total Capital @ $1.00/MM Btu Coal, $1000	893,616
Total Capital @ $1.00/MM Btu, $/kW	905
Total Capital @ $2.00/MM Btu Coal, $1000	909,803
Total Capital @ $2.00/MM Btu, $/kW	921
AVERAGE COST OF SERVICES (2)	
Annual Cost @ $1.00/MM Btu Coal, $1000/yr	249,573
Per Unit @ $1.00/MM Btu Coal, mills/kWh	41.20
Annual Cost @ $2.00/MM Btu Coal, $1000/yr	311,235
Per Unit @ $2.00/MM Btu Coal, mills/kWh	51.38

NOTES:

(1) At 100% Operating Load Factor.
(2) Mid-1976 dollars and 70% Operating Load Factor.

Total capital investment and cost of services are higher with increasing coal cost. Capital investment is nearly 2 percent higher and cost of services about 25 percent higher for $2.00/MM Btu coal when compared to $1/MM Btu coal.

Tables M-7 and M-8 give detailed breakdowns of plant investment, capital charges and working capital at 70% operating load factor for the two coal costs. Plant investment is the same at both coal costs. Capital charges are about 1 percent higher for the $2.00/MM Btu coal and working capital nearly 32 percent higher.

The accuracy of plant investment estimates is judged to be $\pm$25%. Since other capital charges and working capital are keyed to elements of plant investment, this accuracy is reflected in other capital charges as well. Therefore, caution must be exercised in comparing Case MACW with cases representing other gasification technologies.

Major elements of plant investment are Gasification and Ash Handling and the Combined Cycle System. Together they represent about 56 percent of the total plant investment. The steam, condensate and BFW costs are high since they contain tar boilers, precipitators and stack gas scrubbers for this case. Also, the coal handling section contains the cost of coal briquetting equipment.

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

Table M-9 summarizes cost of services based on coal charged at $1.00/MM Btu and $2.00/MM Btu HHV. Costs are computed in accordance with criteria given by EPRI (Criteria section). They are presented as averages for the plant.

Operating labor requirement is a function of the number of units and trains. Requirements on a per shift basis are:

	CASE MACW
"A" Operators	7
"B" Operators	23
Foremen	2
Lab and Instrument Technicians	4

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

Operating charges constitute nearly 44 percent of cost of services with coal at \$1.00/MM Btu and about 54 percent at a coal cost of \$2.00/MM Btu. Coal is the largest single operating charge, representing about 55 percent with coal at \$1.00/MM Btu and 71 percent at the higher coal cost. The relationships as percentages are summarized below:

	CASE MACW	
Coal Cost, $/MM Btu, HHV	1.0	2.0
Coal as % of Operating Charges	53.7	69.9
Coal as % of Total Cost of Services	23.7	38.0
Operating Charges as % of Total Cost of Services	44.1	54.4
Capital Charges as % of Total Cost of Services	55.9	45.6

TABLE M-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $1.00/MM Btu COAL

CASE MACW

	$1000 (1)	$/kW (2)	Percent
PLANT INVESTMENT			
Coal Handling	25,585	25.90	4.36
Oxidant Feed	19,121	19.36	3.25
Gasification and Ash Handling	100,101	101.34	17.00
Gas Cooling	19,158	19.40	3.25
Acid Gas Removal and Sulfur Recovery	29,982	30.35	5.09
Process Condensate Treating	59,067	59.80	10.03
Steam, Condensate and BFW	41,987	42.51	7.13
Support Facilities	61,493	63.25	10.44
Combined Cycle	232,231	235.10	39.45
Subtotal	588,725	597.01	100.00
Contingency	106,567	107.88	
TOTAL PLANT INVESTMENT	695,292	704.89	
ILLINOIS SALES TAX	15,993	16.19	
CAPITAL CHARGES			
Preproduction Costs	43,837	44.38	
Paid-up Royalties	3,476	3.52	
Initial Catalyst and Chemical Charges	1,072	1.09	
Construction Loan Interest	88,842	89.94	
TOTAL CAPITAL CHARGES	137,227	138.93	
DEPRECIABLE CAPITAL	848,512	860.01	
WORKING CAPITAL	45,104	45.66	
TOTAL CAPITAL	893,616	905.67	

NOTES
(1) Mid-1976 dollars.
(2) Based on 100% Operating Load Factor.

TABLE M-8

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR
AND $2.00/MM Btu COAL

CASE MACW

	$1000 (1)	$/kW (2)	Percent
PLANT INVESTMENT			
Coal Handling	25,585	25.90	4.36
Oxidant Feed	19,121	19.36	3.25
Gasification and Ash Handling	100,101	101.34	17.00
Gas Cooling	19,158	19.40	3.25
Acid Gas Removal and Sulfur Recovery	29,982	30.35	5.09
Process Condensate Treating	59,067	59.80	10.03
Steam, Condensate and BFW	41,987	42.51	7.13
Support Facilities	61,493	63.25	10.44
Combined Cycle	232,231	235.10	39.45
Subtotal	588,725	597.01	100.00
Contingency	106,567	107.88	
TOTAL PLANT INVESTMENT	695,292	704.89	
ILLINOIS SALES TAX	15,993	16.19	
CAPITAL CHARGES			
Preproduction Costs	45,592	46.16	
Paid-up Royalties	3,476	3.52	
Initial Catalyst and Chemical Charges	1,072	1.09	
Construction Loan Interest	88,842	89.94	
TOTAL CAPITAL CHARGES	138,982	140.71	
DEPRECIABLE CAPITAL	850,267	861.79	
WORKING CAPITAL	59,536	60.27	
TOTAL CAPITAL	909,803	922.06	

NOTES
(1) Mid-1976 dollars.
(2) Based on 100% Operating Load Factor.

TABLE M-9

COST OF SERVICES AT 70% OPERATING LOAD FACTOR

CASE MACW

COAL COST, HHV	\$1.00/MM Btu	\$2.00/MM Btu
NET PRODUCTION (1)		
Net Power, MW	987.78	987.78
By-product Ammonia ST/SD	63	63
By-product Sulfur ST/SD	128	128
OPERATING CHARGES, \$1000/YR		
Coal	59,137	118,274
Operating Labor	3,653	3,653
Catalyst and Chemicals	373	373
Utilities	1,150	1,150
Maintenance Labor	8,088	8,088
Maintenance Materials	12,132	12,132
Administration and Support Labor	3,522	3,522
General and Administration Expense	7,045	7,045
Ash Disposal	957	957
Property Tax and Insurance	17,382	17,382
By-product Ammonia	(3,270)	(3,270)
By-product Sulfur	(0)	(0)
TOTAL OPERATING CHARGES, \$1000/year	110,169	169,306
CAPITAL CHARGES, \$1000/YR		
Total Capital Charges	139,404	141,929
COST OF SERVICES		
Total, \$1000/yr	249,573	311,235
Per Unit Production, mills/kWh	41.20	51.38

NOTES

(1) At 100% Operating Load Factor.

CASE MXSC - BGC SLAGGER

PLANT DESCRIPTION - CASE MXSC

GENERAL

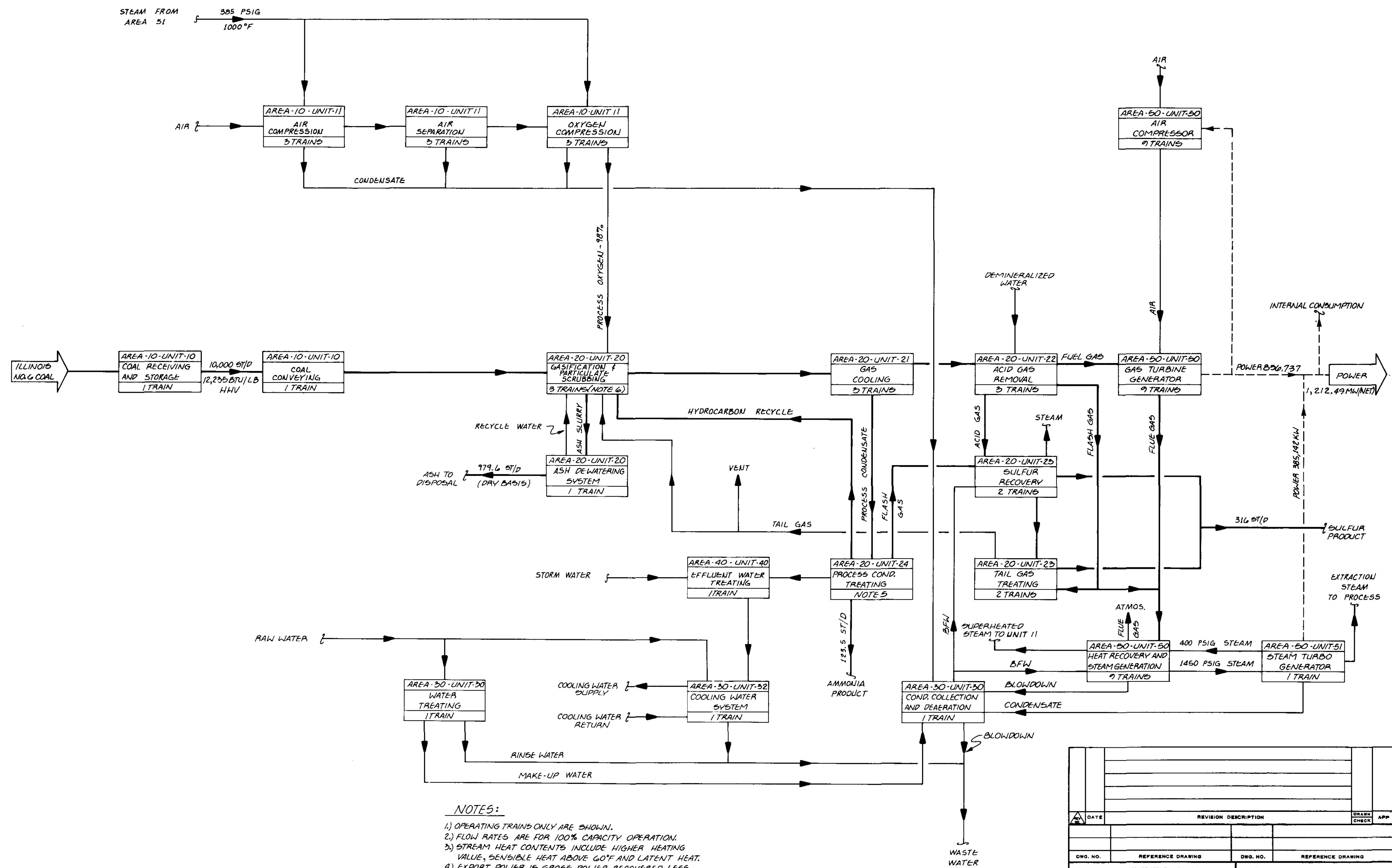
A grass roots plant for electric power generation is shown schematically on the block flow diagram, MXSC-1-1. This plant is based on gasifying 10,000 ST/day of Illinois No. 6 coal in a moving bed slagging bottom gasifier. The block flow diagram identified as Case MXSC represents integration of the fuel gas production process with a combined cycle power generation system. This particular gasifier is oxygen blown and is a representation of the device currently under development by the British Gas Corporation.

The main processing units are in three parallel and largely independent trains. Each process train consists of oxidant feed, gasification, gas cooling and acid gas removal units. Integration between processing trains is minimized. Complete trains may be shut down in order to maintain efficiency during reduced capacity operation. The impact of upset conditions is limited to the train in which the upset occurs.

In addition to the main processing trains, the complete plant includes necessary offsite, utility and environmental facilities. Coal receiving, storage, and conveying is done in a single train to minimize space and operating labor requirements. Hydrogen sulfide removed from gasified coal is processed through sulfur recovery facilities which produce elemental sulfur. All liquid products such as tar, oils, phenols, etc., are recovered from the crude gas and are recycled to the gasifier to extinction.

Other facilities in the plant are raw water treating, cooling water, process condensate treating and effluent water treating.

Table MS-1 summarizes major equipment sections in the plant and shows the number of operating and spare sections.



NOTES:

1.) OPERATING TRAINS ONLY ARE SHOWN.
2.) FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3.) STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE, SENSIBLE HEAT ABOVE 60°F AND LATENT HEAT.
4.) EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
5.) PROCESS CONDENSATE TREATING CONSISTS OF TWO TAR OIL SEPARATION TRAINS AND ONE PHENOL EXTRACTION TRAIN.
6.) EACH TRAIN CONSISTS OF TWO OPERATING GASIFIERS.

DATE	REVISION DESCRIPTION	DRAWN CHECK	APP

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

FLUOR

DRAWN BY B. GEORGE / R. KRESS
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
SUPERVISING ENGR. FERRAR

OVERALL BLOCK FLOW DIAGRAM
MOVING BED SLAGGING COAL GASIFICATION
OXYGEN BLOWN - COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE
DRAWING NUMBER MXSC-1-1
REVISION 1

MXSC-1-1

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TABLE MS-1

MAJOR EQUIPMENT SECTIONS - CASE MXSC

Unit No.	Name	Operating	Spare
10	Coal Preparation	1	0
11	Oxidant Feed System		
	. Air Compression	3	0
	. Air Separation	3	0
	. Oxygen Compression	3	0
20	Gasification	3*	1*
20	Ash Handling	1	0
21	Gas Cooling	3	0
22	Acid Gas Removal	3	0
23	Sulfur Recovery and Tail Gas Treating	2	1
24	Process Condensate Treating		
	. Tar Oil Separation	2	1
	. Phenol Extraction	1	0
	. Ammonia Recovery	1	0
30	Steam, BFW and Condensate System		
	. Condensate Collection and Deaeration	1	0
	. Water Treating	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0
50	Gas Turbine/Generator	9	0
51	Heat Recovery Steam Generators	9	0
51	Steam Turbine/Generator	1	0

* Each train includes two parallel gasifiers.

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COAL PREPARATION

Process Flow Diagram MXS/MXSC-10-1 depicts the process arrangement of equipment in this section.

Washed, 1-1/2" by 1/4" coal is received at the plant site by unit train. No crushing, grinding or refuse disposal systems are included. The coal is unloaded from 100 ton bottom dump cars into an unloading hopper, withdrawn from the hopper by two vibrating feeders and transported by belt conveyors to a tripper. The tripper distributes coal to a traveling belt stacking system. The stacker travels on tracks and forms storage piles on either side. The unloading and stacking system is designed to handle a three day supply in eight hours.

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

Equipment Notes

All equipment is commercially available.

10-BN-1 UNLOADING HOPPER

10-CV-1 BELT CONVEYOR

10-CV-2 BELT CONVEYOR

10-TR-2 TRIPPER

10-ME-1 STACKING SYSTEM

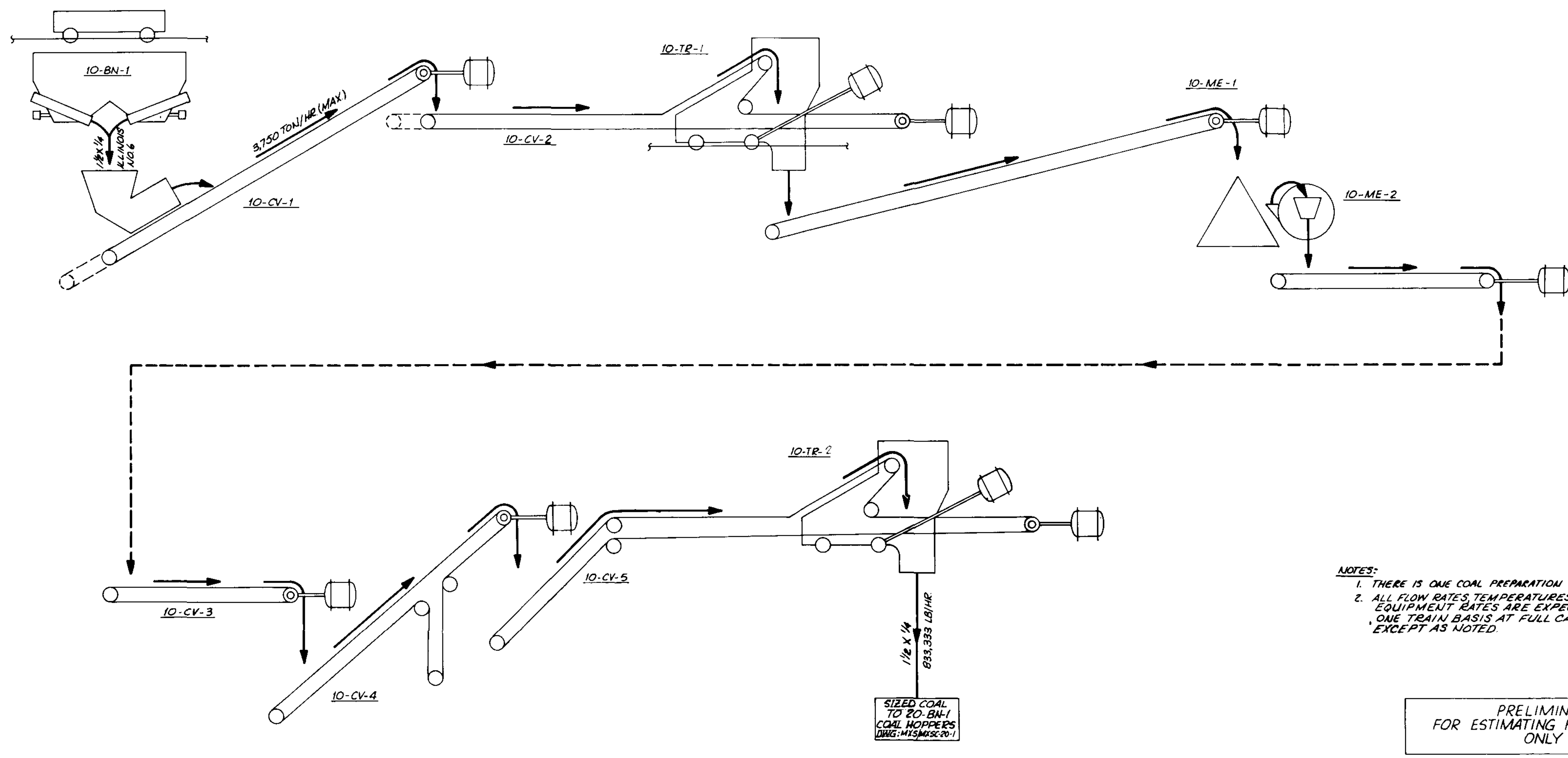
10-ME-2 RECLAIMING SYSTEM

10-CV-3 BELT CONVEYOR

10-CV-4 BELT CONVEYOR

10-CV-5 BELT CONVEYOR

10-TR-1 TRIPPER



NOTES:
1. THERE IS ONE COAL PREPARATION TRAIN.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A ONE TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

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ONLY

REV NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY: M. SEKHON
CHECKED BY: R. RICE
SUPERVISOR: J. BARGER
SUPERVISING ENGR: T.H. TAN
PROJECT ENGR.:
CLIENT:

PROCESS FLOW DIAGRAM
COAL PREPARATION
MOVING BED SLAGGING-OXYGEN BLOWN
ELECTRIC POWER RESEARCH INSTITUTE, PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: MXS/MXSC-10-1
REVISION: 1

OXIDANT FEED

Process Flow Diagram MXSC-11-1 shows the oxidant feed system for Case MXSC. There are three parallel operating trains. Each train has one air compressor, one air separation plant and one oxygen compressor. No spare train is provided in this section.

Atmospheric air is compressed to 95 psig in a three stage centrifugal machine, 11-1-C-1. Heat of compression is rejected to air in interstage airfan coolers 11-1-E-1 and 11-1-E-2, respectively.

The 31,100 hp required by each air feed compressor is supplied by a steam turbine. The steam turbine driver is a condensing type machine operating at inlet steam conditions of 385 psig, 1000°F, with exhaust pressure at 2-1/2" Hg abs. to meet the overall steam balance requirements. The steam turbine is designed with excess capacity to provide response capabilities during turndown or upset conditions. No spare machine is provided for this service.

The compressed air is processed in an air separation unit, 11-1-ME-1, which produces 1700 tons per day (100% basis) of 98% oxygen. Liquid oxygen storage of 5100 tons is provided, with attendant cryogenic pumps and vaporizer. Storage is equivalent to approximately three days of rated capacity operation of a single train. The three days of storage is anticipated to adequately cover any outage of the cryogenic unit.

The air separation plant produces oxygen at 2 psig and 90°F. The oxygen is compressed to 400 psig in six stages. As in the case of the air feed compressor, interstage heat of compression is rejected to interstage air coolers 11-1-E-5 through 9. The compressed oxygen at 400 psig and 214°F flows to the gasifiers.

The 9950 hp compression requirement for the oxygen is supplied by a condensing turbine. The inlet steam condition is 385 psig, 1000°F with backpressure at 2-1/2" Hg abs.

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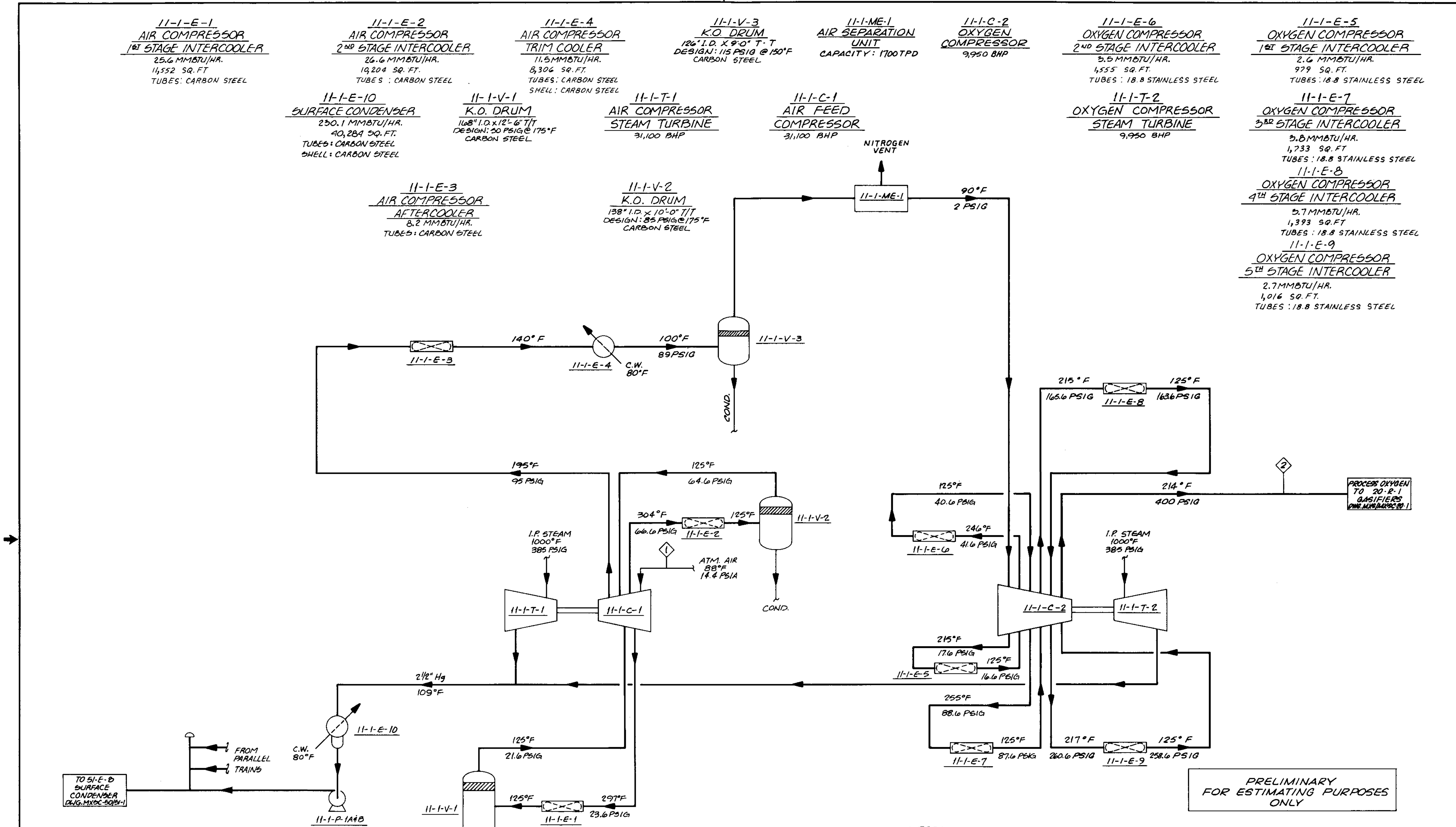

MATERIAL BALANCE				
STREAM NUMBER	1		2	
COMPONENT	AIR FEED MPH	MOL%	PROCESS OXYGEN MPH	MOL%
CH_4	—	—	—	—
C_2H_6	—	—	—	—
H_2	—	—	—	—
CO	—	—	—	—
CO_2	—	—	—	—
H_2S	—	—	—	—
COS	—	—	—	—
NH_3	—	—	—	—
N_2	16,264.2	77.05	81.6	2.00
O_2	4,323.4	20.48	3,999.4	98.00
H_2O	521.4	2.47	—	—
TOTAL MPH	21,109.0	100.00	4,081.0	100.00
TOTAL LB/HR	603,340		130,270	
MOL. WT.	28.58		31.92	

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

FLUOR

DRAWN BY T. BUJACICH
CHECKED BY R. RICE
SUPERVISOR R. RICE
SUPERVISING ENGR. R. BENNETT

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
MOVING BED SLAGGING-OXYGEN BLOWN
COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE
DRAWING NUMBER MXSC-11-1
REVISION 04

4484-MXSC-11-1 (2X3)

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Equipment Notes

The compressors and cryogenic air separation plant are commercially available. The condensing turbines with 1000°F inlet temperature represent an extension of the present state of the art for turbines. It is expected that such turbines would be available in the next few years.

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GASIFICATION AND ASH HANDLING

Process Flow Diagram MXS/MXSC-20-1 shows the gasification step for the Case MXSC. There are four parallel gasification trains (three operating, one spare), each train having two parallel gasifiers. There is one ash handling train.

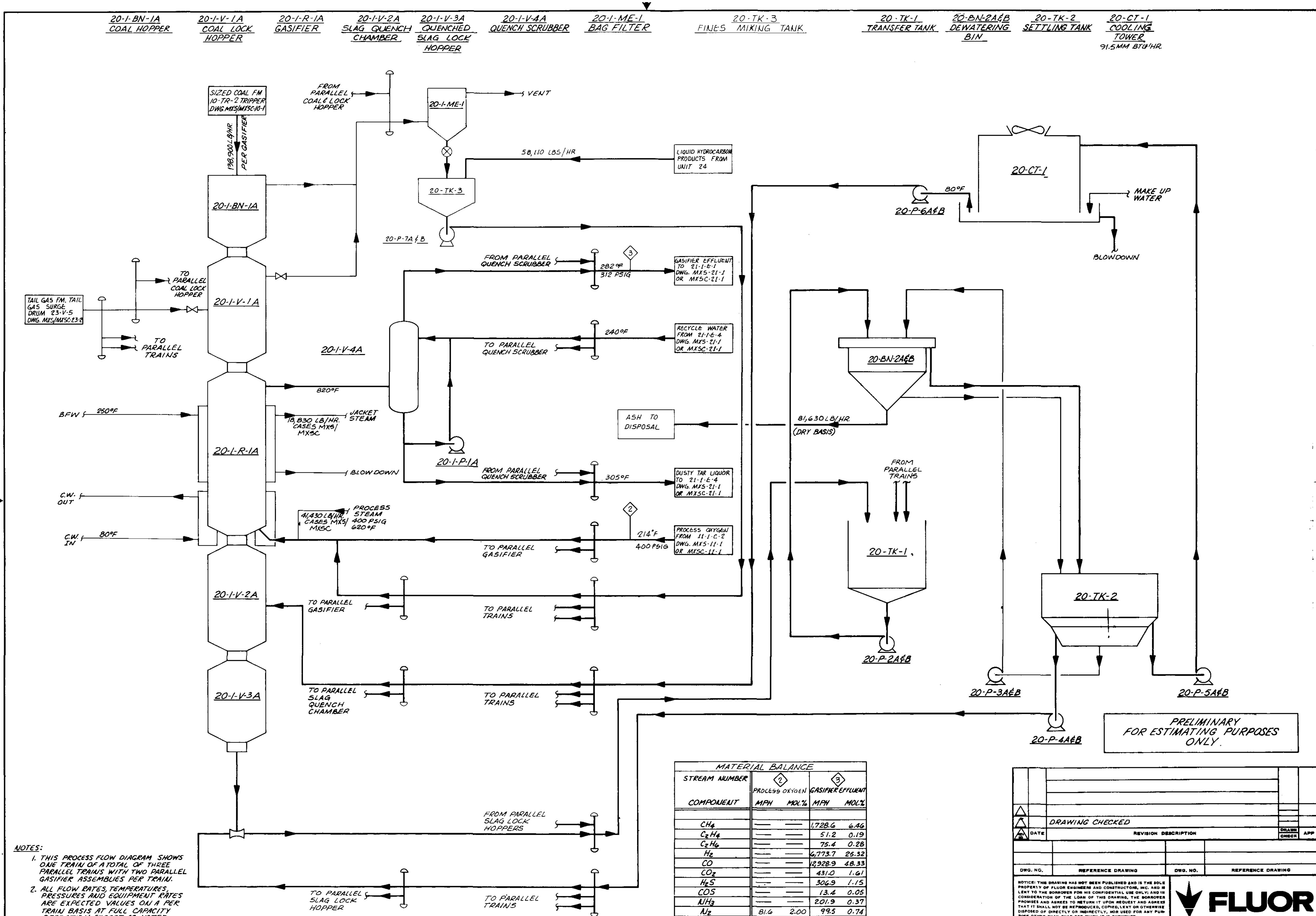
The moving bed slagging bottom gasifier is a water jacketed pressurized unit composed of a series of vertically stacked vessels. There are, from top to bottom, a coal hopper, coal lock, water jacketed gasifier, slag quench chamber, and slag lock hopper.

Coal is conveyed from the coal preparation area to the coal hopper from which it is fed by gravity to the depressurized coal lock through a hydraulically operated valve. The lock is then isolated and pressurized with a slipstream of tail gas and the coal is transferred to the gasifier through another hydraulically operated valve. The empty lock is isolated, depressurized through a bag filter and vented to the atmosphere. In addition, the gas displaced from the coal and lock hoppers during loading is vented to the atmosphere through the bag filter.

Coal dust recovered in the filter is fed to a fines mixing tank, 20-TK-3. Coal fines produced in the coal preparation area are also fed to 20-TK-3. The coal dust and fines are mixed with the liquid hydrocarbon byproducts recovered in Unit 24 and the liquid slurry is pumped to the bottom of the gasifiers through the tuyeres.

The coal flowing down through the gas producer represents a slowly moving bed which has several distinct zones. In the first zone at the top of the gasifier, coal is preheated and dried by contact with the hot crude gas leaving the reactor. As the coal moves down and is heated further, devolatilization occurs and gasification commences. The bottom of the bed is a combustion zone where carbon reacts with oxygen to form CO and CO_2. The oxidation provides the overall heat for the gasification and devolatilization reactions which are endothermic. Only a negligible amount of unburned carbon remains in the slag.

Oxidant and steam, together with recycle coal fines and hydrocarbon products, enter the gasifier near the bottom through an array of tuyeres. The intense heat created by the exothermic reactions in the "raceway" between opposing tuyeres maintains a temperature of approximately 3500°F in the bottom of the gasifier



MATERIAL BALANCE				
STREAM NUMBER	2 PROCESS OXYGEN		3 GASIFIER EFFLUENT	
COMPONENT	MPH	MOL.%	MPH	MOL.%
CH_4	—	—	1,728.6	6.46
C_2H_4	—	—	51.2	0.19
C_2H_6	—	—	75.4	0.28
H_2	—	—	6,773.7	25.32
CO	—	—	12,928.9	48.33
CO_2	—	—	431.0	1.61
H_2S	—	—	306.9	1.15
COS	—	—	13.4	0.05
NH_3	—	—	201.9	0.37
N_2	81.6	2.00	99.5	0.74
O_2	3999.4	98.00	—	—
H_2O	—	—	4,094.9	15.31
TOTAL MPH	4,081.0	100.00	26,705.4	100.00
T,O,P,N & OTHERS LB/HR.	—		4,840	
TOTAL LB./HR.	130,270		522,310	
MOL. WT.	31.92		19.38	

NOTES:

1. THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE PARALLEL TRAINS WITH TWO PARALLEL GASIFIER ASSEMBLIES PER TRAIN.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.

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REV.	DATE	REVISION DESCRIPTION	DRAWN	CHECK	APP

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FLUOR

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CHECKED BY R. RICE
SUPERVISOR BARGER
RELEASE DATE
SUPERVISING ENGR T.H. TAN
INITIALS
PROJECT ENGR. APP. DATE
CLIENT APP. DATE

PROCESS FLOW DIAGRAM
GASIFICATION & ASH HANDLING
MOVING BED SLAGGING - OXYGEN BLOWN

ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIFORNIA

SCALE NONE | DRAWING NUMBER MXS/MXSC-20-1 | REVISION 03

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allowing ash to be withdrawn as a molten slag. Oxidant flow rate is controlled to accomplish complete gasification of coal. Steam rate is controlled to allow the ash to form into a molten slag.

The crude gas leaving the gasifier contains appreciable quantities of tars, oils, naphtha, phenols, fatty acids, ammonia, hydrogen sulfide, sulfur compounds and a small amount of coal and ash dust. The crude gasifier effluent at 820°F flows through quench scrubber 20-1-V-4A, where it is washed with a stream of process condensate. The washing process quenches the gas to 282°F and condenses the high boiling tar fractions. Coal and ash dust are removed with the condensed tar leaving the quenched effluent gas essentially free of particulate matter.

Ash collected as a molten slag in the bottom of the gasifier is periodically discharged downward into slag quench chamber 20-1-V-2A. Slag is quenched with water to form small grained frit and passes into quenched slag hopper 20-1-V-3A.

When the quench slag hopper is full it is isolated from the slag quench chamber. The slag is then discharged through an eductor to a common transfer tank using water as the motive fluid. The quench slag hopper is then recharged with cold water, repressurized and put back into circuit by opening the appropriate valves.

The ash slurry from the transfer tank is pumped to dewatering bins 20-BN-2A&B to produce ash ready for disposal.

Final cleaning of the water overflowing the dewatering bin, 20-BN-2, is accomplished in a settling tank, 20-TK-2, where ash fines settle and are pumped back to the dewatering bin. A portion of the clarified water is recycled to the slag quench chambers after is is cooled in an induced draft type cooling tower (20-CT-1). The balance of the water provides the motive fluid for the ash slurry transfer eductors.

<u>Equipment Notes</u>

The coal feed, coal distribution, stirrer and gas quench technology associated with moving bed gasifiers is commercially proven for noncaking coals via the Lurgi experience. No fundamental problems are expected in this area for handling caking coals such as Illinois No. 6, which is the basis for this case. The

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slagging technology has been under development at the Westfield Development Centre of British Gas Corporation. A slagging bottom gasifier has been operating at Westfield for the past three years supported by a consortium comprising fourteen gas companies and the Electric Power Research Institute. The success of this pilot program has resulted in an ERDA contract of a 60 MM SCFD demonstration plant in Ohio for SNG production based on the slagging gasifier technology.

The ash slurry system is a commercially available system.

GAS COOLING

Process Flow Diagram MXSC-21-1 depicts one of the three parallel trains.

Gasifier quench scrubber effluent is cooled to approximately 262°F in exchanger 21-1-E-1 by cold product gas from 22-1-V-4 in the acid gas removal section. Product gas is then heated to 580°F by high pressure boiler feedwater in exchanger 21-1-E-8 and flows to the gas turbines. The condensate from the crude gas is separated in knockout drum 21-1-V-1.

The crude gas is further cooled to approximately 105°F by a combination of air (21-1-E-2) and water (21-1-E-3) cooling. Knockout drum 21-1-V-2, separates the oily gas liquor condensed in 21-1-E-2 which is further cooled in 21-1-E-9 by water cooling. Crude gas flow in all the exchangers is on the tube side.

The cooled gas from 21-1-E-3 still contains ammonia which must be removed. The ammonia is removed by water scrubbing in an ammonia absorber (21-1-V-3) where gas contacts water countercurrently on trays. The ammonia-free overhead gas from the absorber then flows to the acid gas removal unit for further processing. The ammonia-rich water from the absorber bottom is combined with cooled oily gas condensate from exchanger 21-1-E-9 and further processed in the process condensate unit.

Condensate from 21-1-V-1 is combined with dusty tar liquor from the quench scrubber. The combined stream of dusty tar liquor then exchanges heat with the tar liquor oil separation plant (Unit 24) quench scrubber recycle water in 21-1-E-5. The dusty tar liquor is cooled down to 243°F and flows to Unit 24 for further processing.

The recycle water is further heated in exchanger 21-1-E-4 to approximately 305°F by dusty tar liquor from the quench scrubbers and flows to the quench scrubbers located in the gasification unit.

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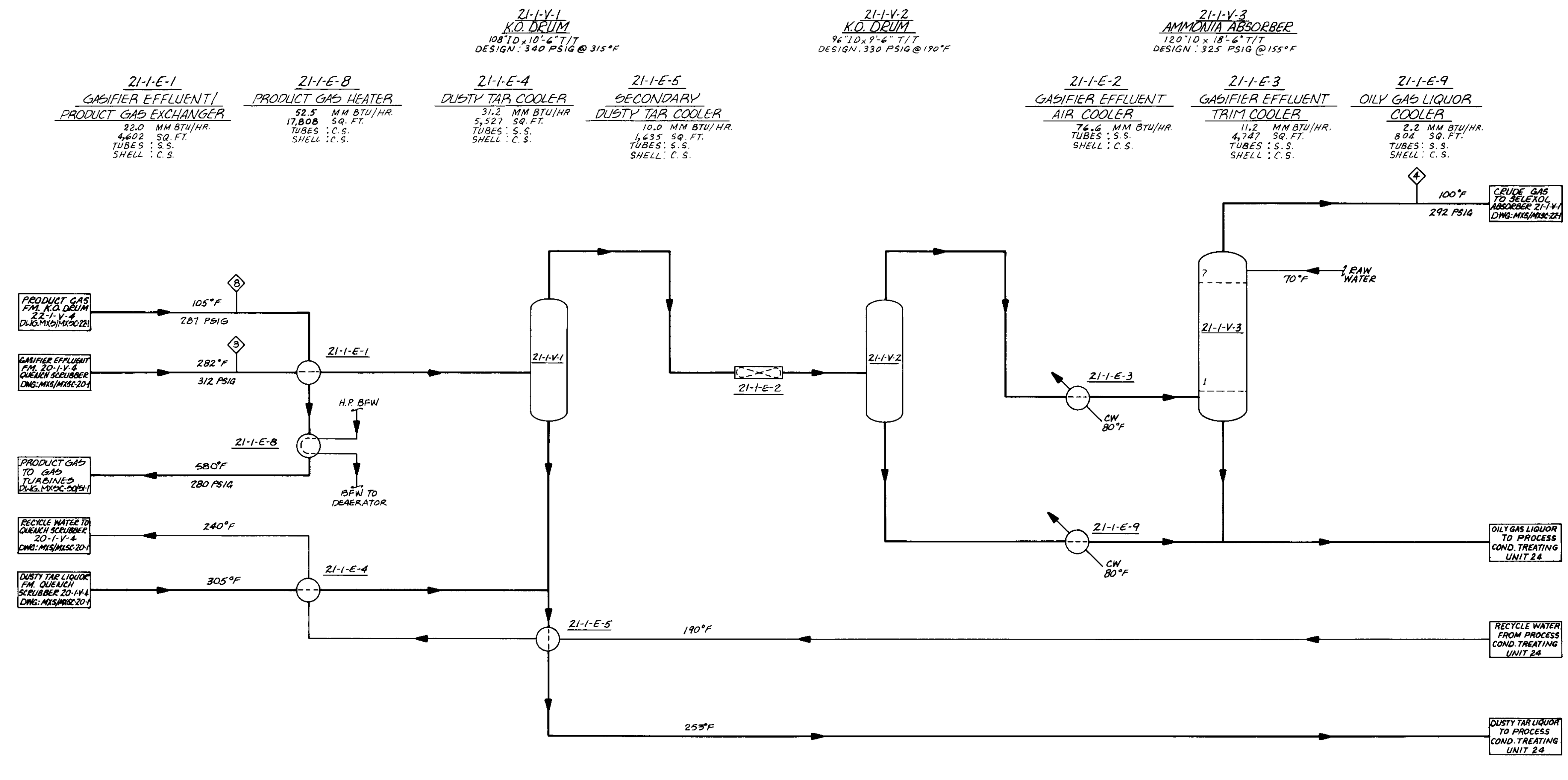

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ONLY

MATERIAL BALANCE						
STREAM NUMBER	3 GASIFIER EFFLUENT		8 PRODUCT GAS		4 CRUDE GAS	
COMPONENT	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	1,728.6	6.46	1,702.5	7.78	1,728.6	7.69
C_2H_4	51.2	0.19	47.8	0.22	51.2	0.23
C_2H_6	75.4	0.28	70.5	0.32	75.4	0.33
H_2	6,773.7	25.32	6,755.0	30.87	6,773.7	30.13
CO	12,928.9	48.33	12,851.5	58.73	12,928.9	57.52
CO_2	431.0	1.61	234.1	1.07	431.0	1.92
H_2S	306.9	1.15	34.6	0.16	306.9	1.37
COS	13.4	0.05	8.6	0.04	13.4	0.06
NH_3	201.9	0.74	—	—	—	—
N_2	99.5	0.37	99.1	0.45	99.5	0.44
H_2O	4,094.9	15.31	78.8	0.26	69.7	0.31
TOTAL MPH	26,705.4	100.00	21,882.5	100.00	22,478.3	100.00
T.O.P.N. & OTHER LB/HR	4,840		—		TRACE	
TOTAL LB/HR	522,310		420,560		441,510	
MOL. WT.	19.52		19.22		19.64	

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6-74 DIETERICH-POST 2111X-6 .003 FILM

NOTES:

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

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SUPERVISOR R. RICE | RELEASE DATE
SUPERVISING ENGR. R. BENNETT | INITIALS
PROJECT ENGR. | APP. DATE
CLIENT | APP. DATE

PROCESS FLOW DIAGRAM
GAS COOLING
MOVING BED SLAGGING OXYGEN BLOWN
COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE | DRAWING NUMBER: MXSC-21-1 | REVISION: 04

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44838A-MXSC-21-1 (S,X,E)

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ACID GAS REMOVAL

Process Flow Diagram MXS/MXSC-22-1 depicts one of the three parallel acid gas removal trains.

The acid gas removal system employs Allied Chemical Corporation's Selexol® process for selective removal of hydrogen sulfide (H_2S). Hydrogen sulfide in the crude gas is absorbed in Selexol® solvent in order to reduce sulfur in the treated gas to 1.0 pound sulfur dioxide (SO_2) equivalent per million Btu (HHV) coal charged to the plant.

The cooled ammonia-free gas flows through an acid gas absorber, 22-1-V-1, where it contacts Selexol® solvent countercurrently over a packed bed. The treated gas from the top of the absorber flows through knockout drum 22-1-V-4 back to the upstream gas cooling unit (Flow Diagram: MXSC-21-1).

The rich solvent from the bottom of the absorber is let down through a hydraulic turbine, 22-1-HT-1, which supplies a portion of the power required by the lean solution pump, 22-1-P-1. It then flows to a flash drum, 22-1-V-2, where most of the dissolved hydrocarbon gases in the solvent flash off. Approximately 98% of the dissolved H_2S and most of the dissolved COS are retained in the solvent because of their selective absorption in the Selexol®. The flash gas flows to the heat recovery steam generators (Flow Diagram: MXSC-50/51-1) and to the sulfur plant.

The rich solvent solution from the flash drum exchanges heat with hot regenerated solution in 22-1-E-1 and flows to the top of the regenerator, 22-1-V-3. In the regenerator the absorbed H_2S and CO_2 are stripped from the solution. Reboil heat is supplied by 100 psig steam in a thermosyphon reboiler, 22-1-E-3. Hot regenerated solvent is pumped back to the absorber, 22-1-V-1, through exchanger 22-1-E-1 in order to reduce reboiler duty. Then the lean solution is cooled down to operating temperature with cooling water in exchanger 22-1-E-2.

Acid gas from the regenerator overhead is cooled to 120°F in airfan cooler 21-1-E-4. The condensate produced in cooling is separated in a knockout drum, 22-1-V-6, and flows to a decanter, 21-1-V-8, by gravity.

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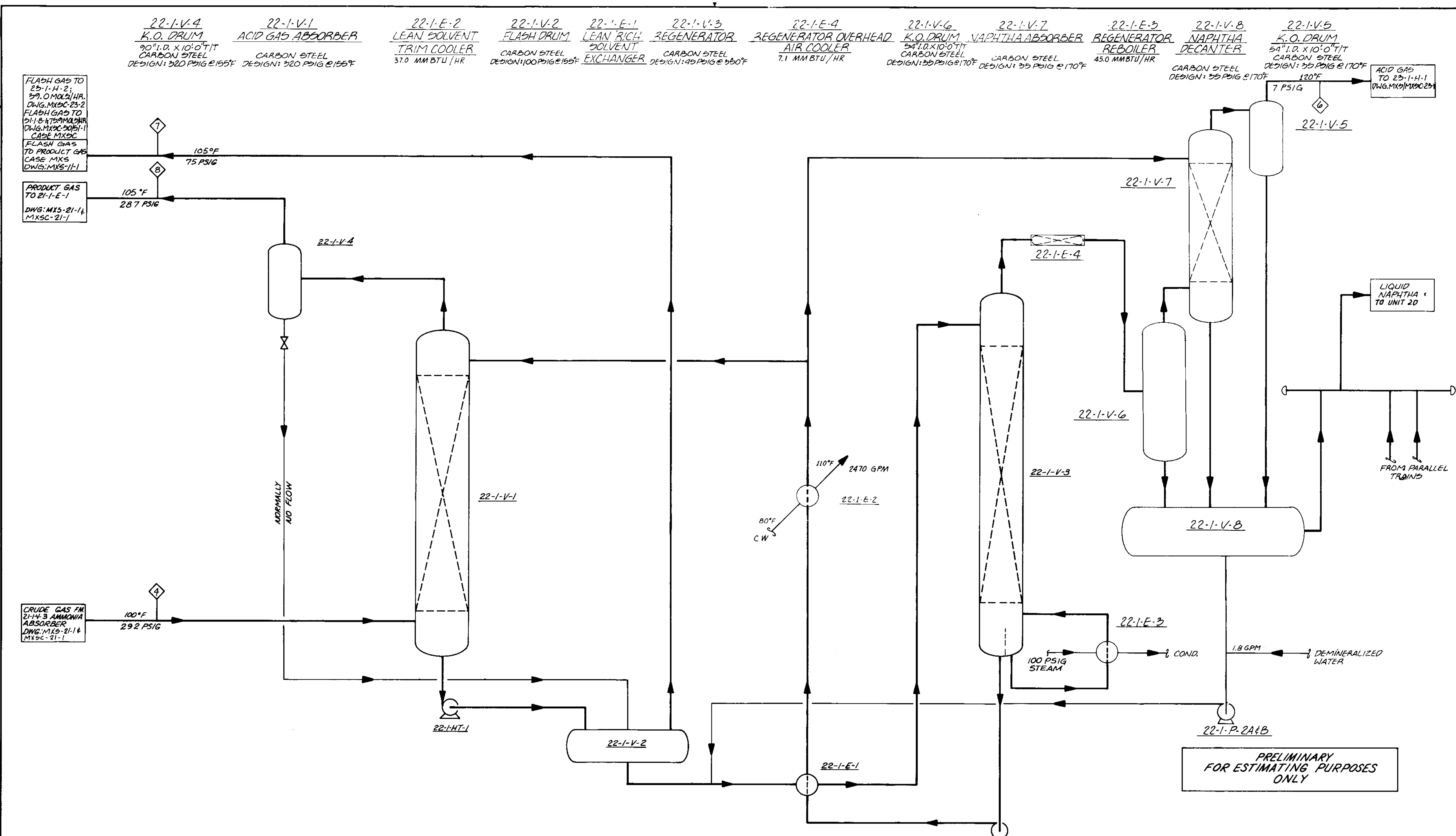

MATERIAL BALANCE

STREAM NUMBER / COMPONENT	4 CRUDE GAS MPH	MOL %	8 PRODUCT GAS MPH	MOL %	7 FLASH GAS MPH	MOL %	6 ACID GAS MPH	MOL %
CH_4	1,728.6	7.69	1,702.5	7.78	14.2	12.36	12.1	2.28
C_2H_4	51.2	0.23	47.8	0.22	0.8	0.70	2.6	0.49
C_2H_6	75.4	0.33	70.5	0.32	1.1	0.96	3.8	0.72
H_2	6,773.7	30.13	6,755.0	30.87	16.2	14.10	2.5	0.47
CO	12,928.9	57.52	12,851.5	58.73	58.0	50.48	19.4	3.66
CO_2	431.0	1.92	234.1	1.07	18.3	15.93	178.6	33.66
H_2S	306.9	1.37	34.6	0.16	4.5	3.92	267.8	50.47
COS	13.4	0.06	8.6	0.04	0.2	0.17	4.6	0.87
N_2	99.5	0.44	99.1	0.45	0.3	0.21	0.1	0.02
NAPHTHA	TRACE				TRACE			
H_2O	69.7	0.31	78.8	0.36	1.3	1.12	39.1	7.36
TOTAL MPH	22478.3	100.00	21,882.5	100.00	114.9	100.00	530.6	100.00
TOTAL LB/HR	441,510		420,560		2,943		18,900	
MOL. WT.	19.64		19.22		25.61		35.62	

NOTES:

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

REV	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY: BOB MCAULEY
CHECKED BY: R. RICE
SUPERVISOR: J. BARGER
SUPERVISING ENGR: T.H. TAN
PROJECT ENGR. | APP. DATE
CLIENT | APP. DATE
RELEASE DATE
INITIALS

PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
MOVING BED SLAGGING-OXYGEN BLOWN
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: MXS/MXSC-22-1
REVISION: 04

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MXS/MXSC-22-1

The cooled acid gas from 22-1-V-6 contains approximately 50.5 mol% H_2S and a small quantity of naphtha. Higher hydrocarbons (C_3 plus) cause problems in the downstream sulfur plant as they do not burn completely. Instead the higher hydrocarbons undergo partial cracking resulting in carbon deposition on the sulfur converter catalyst and production of black sulfur. Naphtha is therefore removed from the acid gas. This is achieved by reabsorption of hydrocarbons (C_4-C_7), from the overhead gases of the knockout drum, 22-1-V-6, in a naphtha absorber, 22-1-V-7. The acid gas contacts a slip stream of cooled lean Selexol® solvent countercurrently over the packing in 22-1-V-7. The bottoms from the absorber then go to the decanter, 22-1-V-8. Phase separation of naphtha and Selexol® solvent is obtained in the decanter because of the high solubility of Selexol® solvent in the water. The condensate from the knockout drum, 22-1-V-6, dissolves the Selexol® solvent and the water-rich phase settles in the bottom of 22-1-V-8. The lighter naphtha forms the top layer and is continuously removed from the decanter.

The hydrocarbon-free Selexol® solvent and condensate stream from 22-1-V-8 is then transported to 22-1-V-2 through pumps 22-1-P-A or B where it combines with the Selexol® solvent from 22-1-V-2. The combined stream then flows to the top of the regenerator through the exchanger, 22-1-E-1.

A small quantity of demineralized water is added to the Selexol® solvent at the discharge of 22-1-P-2A&B to maintain the water balance in the absorption system.

The naphtha free acid gas from the absorber then flows through a knockout drum, 22-1-V-5, to the sulfur plant.

Equipment Notes

The majority of the equipment in this unit is all carbon steel. The equipment has been used in similar services for a number of years.

The naphtha absorber is in operation in a Selexol® plant for sweetening natural gas in Texas. Heavy hydrocarbons have been successfully removed from acid gas in this equipment.

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SULFUR RECOVERY AND TAIL GAS TREATING

The processes used in these units are the same as in Case MACW. Refer to Case MACW and Process Flow Diagrams MACW-23-1, MACW-23-2, and MACW-23-3 for the detailed process description of these units.

There are two 50% parallel operating sulfur recovery trains each followed by a tail gas treating unit. Sulfur recovery per train is 158 short tons/day. There is a third (spare) train because of the important environmental requirements these units fulfill.

PROCESS CONDENSATE TREATING

The processing schemes for these units are similar to the ones used in Case MACW. Refer to Process Flow Diagrams MACW-24-1, MACW-24-2, MACW-24-3 and MACW-24-4, and Case MACW for the detailed process description of these units.

There are three parallel 50% trains, two operating and one spare, for the tar oil separation unit. There is a single train for the phenol extraction unit. Storage capacity for 5 days at 100% plant capacity has been provided for phenolic water. This storage capacity is anticipated to cover any outage of the phenol extraction unit adequately. The hydrocarbon liquids (tar, oil and crude phenols) extracted in these units are stored in the storage tank, 24-TK-1, and are pumped back to Unit 20 for gasification to extinction (Flow Diagram: MXS/MXSC-20-1).

An anhydrous ammonia stream of 123.5 tons/day suitable for fertilizer and commercial uses is recovered.

STEAM, BOILER FEEDWATER AND CONDENSATE

Process Flow Diagram MXSC-30-1 schematically represents steam, boiler feedwater, and condensate systems.

The steam system operates at four pressure levels:

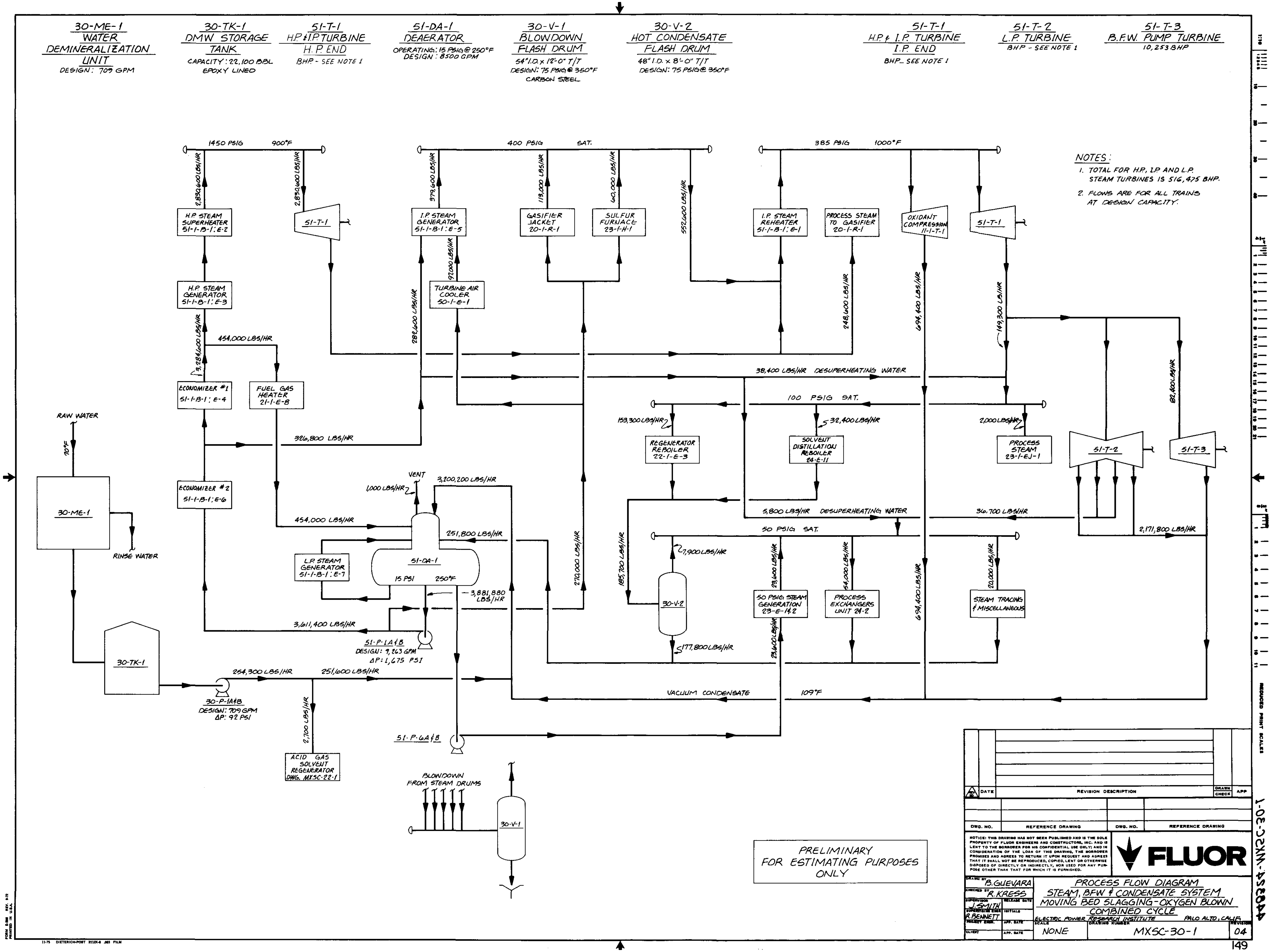
High Pressure	-	1450 psig
Intermediate Pressure	-	400 psig
Low Pressure	-	100 psig
Low Pressure	-	50 psig

Major high pressure (HP) steam generation is carried out in heat recovery steam generator (HRSG) 51-1-B-1 of Gas Turbine 50-1-GT-1. There are nine gas turbines. Each gas turbine has its own HRSG. The HRSG is described in detail in Appendix A. All the HP steam is used to drive the single backpressure type turbine, 51-T-1.

The HP end of 51-T-1 exhausts steam at 400 psig. Additional 400 psig steam generation is obtained in the sulfur plant (23-1-H-1), the gasifier jackets (20-1-R-1) and the HRSGs (51-1-B-1:E-5). Part of the 400 psig steam flows to the gasification area to meet the process steam demand of the gasifiers, 20-1-R-1. The balance is superheated to 1000°F in the IP steam reheaters (51-1-B-1:E-1) of the HRSGs. The superheated intermediate pressure (IP) steam is then used in the IP end of the turbine, 51-T-1, which exhausts to the 100 psig header.

A portion of the 100 psig steam is desuperheated and supplied to the process exchangers (22-1-E-3 and 24-E-11) in acid gas removal and tar oil separation units. A small quantity of desuperheated steam also flows to the steam ejectors (23-1-EJ-1) in the acid gas removal unit. The balance of the 100 psig steam is used in the low pressure (LP) turbine, 51-T-2, and a condensing type boiler feed water (BFW) pump turbine, 51-T-3. The LP turbine has two stages. The first stage operates at a backpressure of 50 psig and the second stage exhausts at 2-1/2" Hg abs. A portion of the steam is extracted after the first stage of the LP turbine and supplied to the 50 psig header. The balance of the steam flows to the second stage of the LP steam turbine.

The 50 psig steam is generated in the process exchangers (23-E-1&2) in the sulfur plant. Extraction steam from the turbine, 51-T-2, after desuperheating also


30-ME-1
WATER DEMINERALIZATION UNIT
DESIGN: 709 GPM
30-TK-1
DMW STORAGE TANK
CAPACITY: 22,100 BBL
EPOXY LINED
51-T-1
H.P. & I.P. TURBINE
H.P. END
BHP - SEE NOTE 1
51-DA-1
DEAERATOR
OPERATING: 15 PSIG @ 250°F
DESIGN: 8500 GPM
30-V-1
BLOWDOWN FLASH DRUM
54" I.D. x 12'-0" T/T
DESIGN: 75 PSIG @ 350°F
CARBON STEEL
30-V-2
HOT CONDENSATE FLASH DRUM
48" I.D. x 8'-0" T/T
DESIGN: 75 PSIG @ 350°F
51-T-1
H.P. & I.P. TURBINE
I.P. END
BHP - SEE NOTE 1
51-T-2
L.P. TURBINE
BHP - SEE NOTE 1
51-T-3
B.F.W. PUMP TURBINE
10,253 BHP
NOTES:
1. TOTAL FOR H.P., I.P. AND L.P. STEAM TURBINES IS 516,475 BHP.
2. FLOWS ARE FOR ALL TRAINS AT DESIGN CAPACITY.
1450 PSIG 900°F
400 PSIG SAT.
385 PSIG 1000°F
100 PSIG SAT.
50 PSIG SAT.
H.P. STEAM SUPERHEATER 51-1-B-1: E-2
H.P. STEAM GENERATOR 51-1-B-1: E-3
ECONOMIZER #1 51-1-B-1: E-4
ECONOMIZER #2 51-1-B-1: E-6
FUEL GAS HEATER 21-1-E-8
I.P. STEAM GENERATOR 51-1-B-1: E-5
TURBINE AIR COOLER 50-1-E-1
GASIFIER JACKET 20-1-R-1
SULFUR FURNACE 23-1-H-1
I.P. STEAM REHEATER 51-1-B-1: E-1
PROCESS STEAM TO GASIFIER 20-1-R-1
OXIDANT COMPRESSION 11-1-T-1
REGENERATOR REBOILER 22-1-E-3
SOLVENT DISTILLATION REBOILER 24-E-11
PROCESS STEAM 23-1-EJ-1
L.P. STEAM GENERATOR 51-1-B-1: E-7
50 PSIG STEAM GENERATION 23-E-1&2
PROCESS EXCHANGERS UNIT 24-2
STEAM TRACING & MISCELLANEOUS
ACID GAS SOLVENT REGENERATOR DWG. MXSC-22-1
2,830,600 LBS/HR
3,284,600 LBS/HR
454,000 LBS/HR
326,800 LBS/HR
979,600 LBS/HR
97,000 LBS/HR
288,600 LBS/HR
118,000 LBS/HR
60,000 LBS/HR
552,000 LBS/HR
248,600 LBS/HR
694,400 LBS/HR
149,300 LBS/HR
38,400 LBS/HR DESUPERHEATING WATER
82,400 LBS/HR
153,300 LBS/HR
32,400 LBS/HR
2,000 LBS/HR
5,800 LBS/HR DESUPERHEATING WATER
36,700 LBS/HR
2,171,800 LBS/HR
RAW WATER
70°F
RINSE WATER
VENT
1,000 LBS/HR
3,200,200 LBS/HR
251,800 LBS/HR
15 PSI 250°F
3,881,880 LBS/HR
3,611,400 LBS/HR
270,000 LBS/HR
185,700 LBS/HR
7,900 LBS/HR
177,800 LBS/HR
23,600 LBS/HR
54,000 LBS/HR
20,000 LBS/HR
51-P-1A&B
DESIGN: 9,263 GPM
ΔP: 1,675 PSI
30-P-1A&B
DESIGN: 709 GPM
ΔP: 92 PSI
254,300 LBS/HR
251,600 LBS/HR
2,700 LBS/HR
VACUUM CONDENSATE
109°F
51-P-6A&B
BLOWDOWN FROM STEAM DRUMS
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
REVISION DESCRIPTION
REFERENCE DRAWING
FLUOR
B. GUEVARA
R. KRESS
J. SMITH
R. BENNETT
PROCESS FLOW DIAGRAM
STEAM, BFW & CONDENSATE SYSTEM
MOVING BED SLAGGING-OXYGEN BLOWN COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE
PALO ALTO, CALIF.
NONE
MXSC-30-1
04
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flows to the 50 psig steam header. Additional 50 psig steam generation is obtained in 30-V-2 by flashing the condensate from the 100 psig steam users. The flashed steam from 30-V-2 is vented to the 50 psig steam header.

The major portion of the 50 psig process steam is supplied to the process exchangers in Unit 24. The balance of the 50 psig steam is used for steam tracing and other miscellaneous users. The low pressure (LP) condensate produced combines with the flashed condensate from 30-V-2 and flows to the deaerator, 51-DA-1.

Raw water is treated in a semiautomatic, resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for a 1500 psig boiler system. Storage equivalent to 24 hours of demineralized water production is provided. The demineralized water requirement is estimated at approximately 509 gpm. Some demineralized water is also used to satisfy the process requirement. The demineralized water is combined with the vacuum condensate returned from the surface condensers of the turbines, 11-1-T-1, 51-T-2 and 51-T-3. The combined stream then flows to the deaerator, 51-DA-1.

The deaerator is a tray type unit operating at 28 psia. The deaerating steam is supplied by the LP evaporators (51-1-B-1:E-7) of the HRSGs. The deaerator provides for ten minutes storage capacity.

Boiler feedwater (BFW) make-up for the HP and IP steam generation is pumped from the deaerator by high pressure BFW pumps (51-P-1A&B) to the HRSG units. BFW heating, IP and HP steam generation and superheating is accomplished in the HRSGs by heat recovery from the flue gases produced in gas turbines, 50-1-GT-1. BFW is first heated in economizers #2 (51-1-B-1:E-6). Make-up BFW for IP steam generation is withdrawn and transported to the IP evaporators (51-1-B-1:E-5). The balance of the BFW is further heated to approximately 598°F in economizers #1 (51-1-B-1:E-4). Part of the hot BFW from economizers #2 is circulated through product gas heaters, 21-1-E-8, and flows back to the deaerator. The make-up BFW for high pressure steam generation flows to the HP evaporators (51-1-B-1:E-3). The HP steam is superheated in 51-1-B-1:E-2 and flows to the superheated HP steam header.

The low pressure (50 psig) steam generators are supplied boiler feedwater by a separate set of pumps, 51-P-6A&B.

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COMBINED CYCLE SYSTEM

Process Flow Diagram MACW-50/51-1 depicts the combined cycle system for Case MXSC.

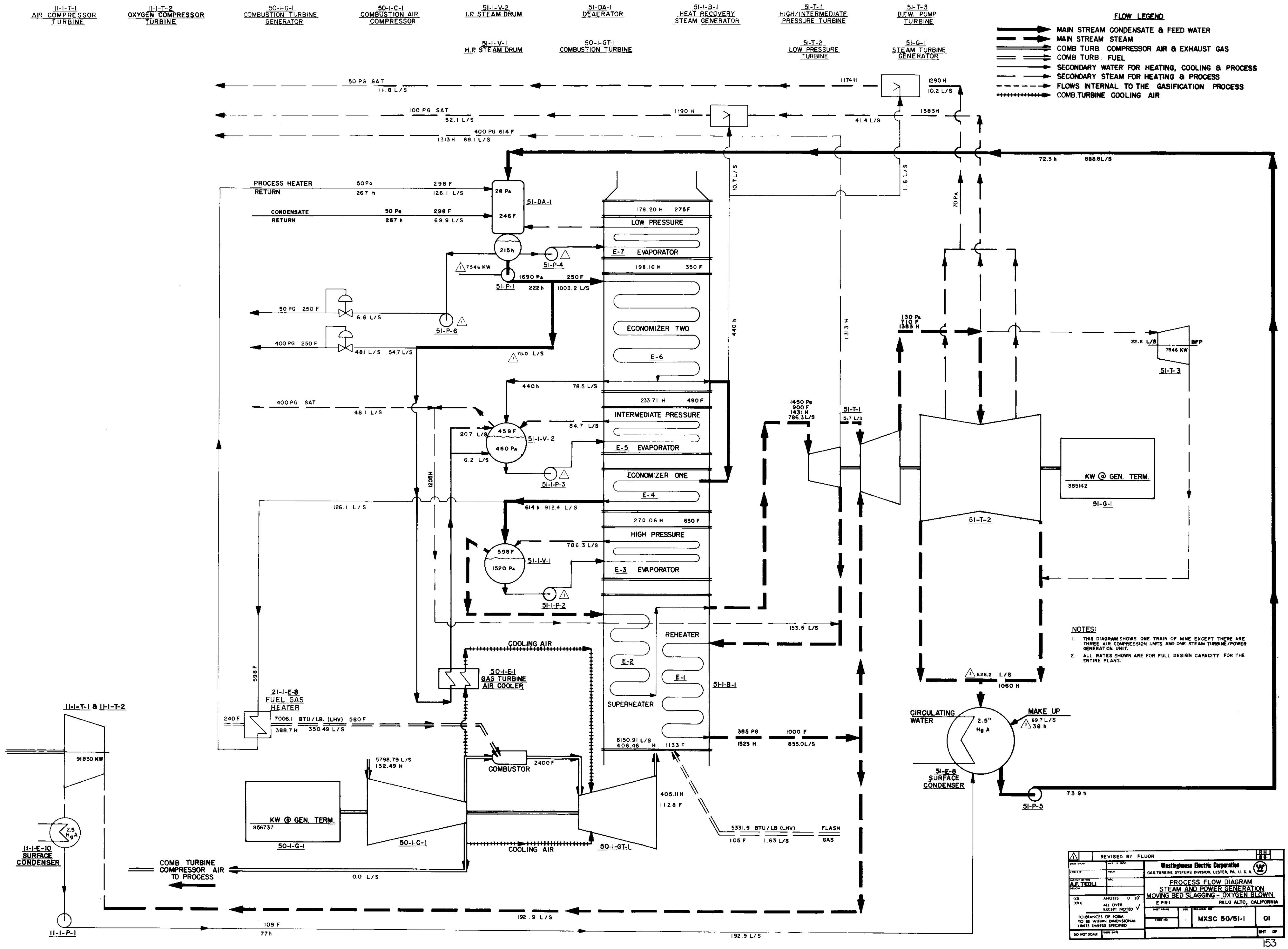
There are nine parallel trains of gas turbines 50-1-GT-1, generators 50-1-G-1 and heat recovery steam generators (HRSG) 51-1-B-1, one 100 percent steam turbine 51-T-1&2, and generator unit 51-G-1. Refer to Appendix A for a detailed description of the combined cycle system. Detailed performance information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

A distinct feature of Case MXSC is that the fuel gas supplied to the gas turbines from the process plant is heated to 580°F in 21-1-E-8 against feed water at 598°F extracted from economizers #1 (51-1-B-1:E-4) of the HRSG.

Steam for boiler feed water pump drives is taken from the steam cross over from the IP to LP steam turbine cylinders. Steam for the air and oxygen compressors' drives is taken from the hot reheat line.

Equipment Notes

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



11-1-T-1 AIR COMPRESSOR TURBINE
11-1-T-2 OXYGEN COMPRESSOR TURBINE
50-1-G-1 COMBUSTION TURBINE GENERATOR
50-1-C-1 COMBUSTION AIR COMPRESSOR
51-1-V-2 I.P. STEAM DRUM
51-1-V-1 H.P. STEAM DRUM
51-DA-1 DEAERATOR
50-1-GT-1 COMBUSTION TURBINE
51-1-B-1 HEAT RECOVERY STEAM GENERATOR
51-T-1 HIGH/INTERMEDIATE PRESSURE TURBINE
51-T-2 LOW PRESSURE TURBINE
51-T-3 B.F.W. PUMP TURBINE
51-G-1 STEAM TURBINE GENERATOR
FLOW LEGEND
MAIN STREAM CONDENSATE & FEED WATER
MAIN STREAM STEAM
COMB TURB. COMPRESSOR AIR & EXHAUST GAS
COMB TURB. FUEL
SECONDARY WATER FOR HEATING, COOLING & PROCESS
SECONDARY STEAM FOR HEATING & PROCESS
FLOWS INTERNAL TO THE GASIFICATION PROCESS
COMB.TURBINE COOLING AIR
50 PG SAT 11.8 L/S
100 PG SAT 52.1 L/S
400 PG 614 F 1313H 69.1 L/S
PROCESS HEATER RETURN 50 Pa 267 h 298 F 126.1 L/S
CONDENSATE RETURN 50 Pa 267 h 298 F 69.9 L/S
28 Pa 246 F
LOW PRESSURE
E-7 EVAPORATOR
ECONOMIZER TWO
E-6
INTERMEDIATE PRESSURE
E-5 EVAPORATOR
ECONOMIZER ONE
E-4
HIGH PRESSURE
E-3 EVAPORATOR
REHEATER
E-2
E-1
SUPERHEATER
51-P-1
51-P-2
51-P-3
51-P-4
51-P-5
51-P-6
7546 KW
1690 Pa 250 F 222 h 1003.2 L/S
50 PG 250 F 6.6 L/S
400 PG 250 F 48.1 L/S 54.7 L/S
75.0 L/S
400 PG SAT 48.1 L/S
459 F 460 Pa
598 F 1520 Pa
COOLING AIR
50-1-E-1 GAS TURBINE AIR COOLER
21-1-E-8 FUEL GAS HEATER
7006.1 BTU/LB. (LHV) 580 F 388.7 H 350.49 L/S
COMBUSTOR 2400 F
5798.79 L/S 132.49 H
KW @ GEN. TERM. 856737
50-1-G-1
50-1-C-1
50-1-GT-1
405.11H 1128 F
5331.9 BTU/LB (LHV) 105 F 1.63 L/S FLASH GAS
6150.91 L/S 406.46 H 1133 F
385 PG 1523 H 1000 F 855.0L/S
COMB. TURBINE COMPRESSOR AIR TO PROCESS 0.0 L/S
11-1-T-1 & 11-1-T-2 91830 KW
11-1-E-10 SURFACE CONDENSER
11-1-P-1
109 F 77h
192.9 L/S
1450 Pa 900 F 1431 H 786.3 L/S
130 Pa 710 F 1383 H
22.8 L/S 7546 KW BFP
KW @ GEN. TERM. 385142
626.2 L/S 1060 H
CIRCULATING WATER 2.5" Hg A
MAKE UP 69.7 L/S 38 h
51-E-8 SURFACE CONDENSER
72.3 h 888.8 L/S
73.9 h
NOTES:
1. THIS DIAGRAM SHOWS ONE TRAIN OF NINE EXCEPT THERE ARE THREE AIR COMPRESSION UNITS AND ONE STEAM TURBINE/POWER GENERATION UNIT.
2. ALL RATES SHOWN ARE FOR FULL DESIGN CAPACITY FOR THE ENTIRE PLANT.
REVISED BY FLUOR
Westinghouse Electric Corporation
GAS TURBINE SYSTEMS DIVISION, LESTER, PA., U.S.A.
A.F. TEOLI
PROCESS FLOW DIAGRAM
STEAM AND POWER GENERATION
MOVING BED SLAGGING - OXYGEN BLOWN
EPRI PALO ALTO, CALIFORNIA
MXSC 50/51-1 01
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PROCESS DISCUSSION

The table below summarizes the pertinent heat and material balance results.

TABLE MS-2

SUMMARY OF SYSTEM PERFORMANCE - CASE MXSC

GASIFICATION AND GAS CLEANING SYSTEM	
Coal Feed Rate, lbs/hr (m.f.)	798,333
Oxygen or Air (1)/Coal Ratio, lbs/lb m.f.	0.481
Oxidant Temperature, °F	214
Steam/Coal Ratio, lbs/lb m.f.	0.31
Gasifier Exit Pressure, psig	320
Crude Gas Temperature, °F	820
Crude Gas HHV (dry basis), Btu/scf (2)	379.0
Temperature of Fuel Gas to Gas Turbine, °F	580
POWER SYSTEM	
Gas Turbine Inlet Temperature, °F	2,400
Pressure Ratio	17:1
Turbine Exhaust Temperature, °F	1,128
Steam Conditions, psig/°F/°F	1450/900/1000
Condenser Pressure, inches Hg abs.	2.5
Stack Temperature, °F	275
Gas Turbine Power, MW (3)	857
Steam Turbine Power, MW (3)	385
Power Consumed, MW	30
Net System Power, MW	1,212
OVERALL SYSTEM	
Process and Deaerator Makeup Water, gpm/1000 MW	834
Cooling Tower Makeup Water, gpm/1000 MW	5,882
Cooling Water Circulation Rate, gpm/MW	307
Cooling Tower Heat Rejection, % of Coal HHV	33.8
Air Cooler Heat Rejection, % of Coal HHV	4.7
Net Heat Rate, Btu/kWh	8,410
Overall System Efficiency (Coal → Power), % of Coal HHV	40.6

NOTES

(1) Dry Basis, 100% O_2 for Oxygen Blown
(2) Excluding HHV of H_2S, COS, and NH_3
(3) At Generator Terminals

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Gasifier Material Balance

Table MS-3 details the material balance around the gasifiers for this case. The figures are based on 100% capacity operation.

The liquid hydrocarbon by-products separated from the crude gas and waste water streams are recovered and mixed with the coal fines produced in the gasification and coal preparation units. This slurry is recycled to the gasifier tuyeres, and is assumed to be completely gasified in the high temperature section of the gasifier. A further assumption made in this case is that the total production of coal fines in the plant is low enough so that, when combined with the total liquid recycled to the gasifiers, the slurry concentration will not exceed 50% solids. With approximately 50% weight, maximum, slurry composition the coal fines recovery would be limited to approximately 7% wt of the total coal feed to the plant. This number was judged reasonable and therefore fines briquetting facilities were not provided for this case.

This gasifier, based on the British Gas Corporation's slagging gasifier development, is unique from two points of view. First, 84.94% weight of the carbon in the coal is converted to CO resulting in a fuel gas containing very little CO_2, thus preserving a major fraction of the chemical heat in the coal for use in the gas turbine combustor. This gasifier also decomposes 90% of the steam feed to hydrogen thereby simplifying the gas cooling system and resulting in a more efficient system.

By-products of the overall process are ammonia and elemental sulfur, production of which are 123.5 ST/D and 316 ST/D respectively.

TABLE MS-3

MATERIAL BALANCE - CASE MXSC

FEEDS

	T(°F)	lb/hr	lb mol/hr
Coal	77		
Moisture		35,000	1,942.8
Ash		80,000	
MAF Coal			
Carbon		554,984	46,205.9
Hydrogen		42,525	21,094.6
Oxygen		80,022	2,500.8
Nitrogen		9,985	356.4
Sulfur		30,817	961.1
TOTAL COAL		833,333	
Oxidant	214		
Oxygen		383,934	11,998.3
Nitrogen		6,863	245.0
TOTAL OXIDANT		390,797	12,243.3
Steam	620	248,593	13,798.5
Liquid Hydrocarbons	157		
N + T + O (1)		48,990	
P + O (2)		9,123	
TOTAL LIQUID HYDROCARBONS		58,113	
TOTAL FEEDS		1,530,836	

EFFLUENTS

	psig	T(°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	320	820			
CH_4			83,205	5,185.8	7.29
C_2H_4			4,312	153.7	0.22
C_2H_6			6,805	226.3	0.32
H_2			40,971	20,321.1	28.56
CO			1,086,550	38,786.7	54.51
CO_2			56,914	1,293.0	1.82
H_2S			31,384	920.7	1.29
COS			2,415	40.2	0.06
N_2			8,365	298.6	0.41
NH_3			10,313	605.6	0.85
H_2O			59,858	3,322.5	4.67
Subtotal			1,391,092	71,154.1	100.00
N + T + O (1)			48,990		
P + O (2)			9,123		
Subtotal			58,113		
TOTAL GASIFIER EFFLUENT			1,449,205		
Ash		2,800			
Carbon			1,631		
Ash			80,000		
TOTAL ASH			81,631		
TOTAL EFFLUENTS			1,530,836		

NOTES:

	(1) Naphtha, Tars, Oils Wt %	(2) Phenols + Others Wt %
Carbon	85.80	74.10
Hydrogen	6.80	6.40
Oxygen	4.35	17.00
Nitrogen	1.12	1.00
Sulfur	1.93	1.50
	100.00	100.00

TABLE MS-4

ENERGY BALANCE: CASE MXSC

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

	MM Btu/hr					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Oxidant Compressor Suction Air		12	30			42
Turbine Compressor Suction Air		141	340			481
Demineralized Water		3				3
Auxiliary Power Inputs					101	101
TOTAL	10,196	161	370	0	101	10,828
HEAT OUT						
Ash/Slag	23	100				123
Gasifier Heat Loss				21		21
Gasifier Cooling		41				41
Sulfur Product	106	1				107
Ammonia Product	100					100
Generated Power					4,239	4,239
Power Block Losses (1)				50	196	246
Oxidant Compressor Surface Condenser			683			683
Power Surface Condenser			2,221			2,221
HRSG Stack Losses		1,177	949			2,126
Gasifier Effluent Cooling		46	224			270
Oxidant Compressor Cooling		239	28			267
N_2 Vent from Oxygen Plant		14				14
Selexol Solvent Cooling		111				111
Selexol Regeneration Overhead Cooling		2	19			21
Process Condensate Cooling		7				7
Steam Heat Losses		18	3			21
Tail Gas Unit Cooling		17				17
Process Condensate Treating Unit		171				171
Spent Tail Gas	22	2	7			31
Waste Water Effluent		14				14
TOTAL	249	1,984	4,134	71	4,435	10,851

$$\frac{\text{Output - Input}}{\text{Output}} = 0.21\%$$

(1) Includes mechanical and electrical losses.

TABLE MS-5

ENERGY BALANCE AS PERCENT OF COAL HHV - CASE MXSC

	mm Btu/hr	Percent
IN		
Coal HHV	10,196	100.00
OUT		
Net Power	4,138	40.58
Sulfur Product Heat	107	1.05
Ammonia Product HHV	100	0.98
Spent Tail Gas HHV	22	0.22
Selexol (Sensible and Latent)	132	1.29
Oxidant Interstage Cooling	267	2.62
Ash/Slag Heat	123	1.20
HRSG Stack Gases	2,126	20.85
Rejected at Condensers	2,904	28.48
Other Sensible Losses	169	1.65
Other Latent Losses	(136)	(1.33)
Gasifier Heat Losses	21	0.21
Power Block Losses	246	2.41
	10,219	100.21

Acid Gas Removal

A distinct feature of Case MXSC is the production of a smaller quantity of CO_2 in the gasifier effluent compared to other cases reported here. Selective removal of H_2S over CO_2 is therefore not as important for this case. Allied Chemical's Selexol® process was selected for Case MXSC to have a common basis for this section with other cases reported here.

The Selexol® process compares favorably with other acid gas removal processes economically.

The Selexol® process results in an H_2S concentration over 20 percent in the acid gas feed to the sulfur recovery unit. At H_2S concentrations in this range, a sulfur plant design commonly referred to as "split flow" may be employed that avoids use of fuel gas in the sulfur furnace. Fuel gas must be burned in the furnace to sustain a flame if H_2S concentration is under 15 percent. In the split flow design the flame can be sustained by burning acid gas with flash gas from the process condensate treating unit.

A naphtha absorber is provided in the acid gas removal unit to recover heavy hydrocarbons from the acid gas feed to the sulfur plant and to prevent naphtha accumulation in the solvent. The naphtha product is decanted from the solvent and recycled to the gasifier with liquid hydrocarbon products recovered in Unit 24. A small energy penalty is incurred by use of the lean Selexol® solvent in the naphtha absorber, which results in a slightly higher circulation rate in the acid gas unit.

Process Energy Balances

Table MS-4 presents an overall process energy balance for this case at 100% capacity operation. The boundary for the balance encompasses the entire plant. Energy content of the stream crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F, and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kWh. The energy balance closes to less than 0.5 percent. The discrepancies result from approximations used for some units and for calculating some heat loads.

Data from Table MS-4 is shown in MMBtu/hr and as percent of coal higher heating value in Table MS-5.

As shown by the tables, this scheme produces a total power generation of 4239 MM Btu/hr from a coal charge of 10,196 MM Btu/hr. The system cold efficiency (net power export at 3413 Btu/kWh as a percentage of the Coal Charge HHV) is 40.58%. The heat rate based upon net power production is 8410 Btu/kWh.

Substantial amounts of liquid hydrocarbon by-products are produced in the form of naphthas, oils, tars and phenols. These by-products are recovered, and this recovery represents an energy penalty due to the process heat required for the tar and phenol units (171 MM Btu/hr). The liquid hydrocarbon by-products are recycled back to the gasifiers' combustor zone through tuyeres.

The steam to oxygen mole ratio for this case is approximately 1.12. The steam decomposition is 90%. The high utilization of steam in the process allows for smaller equipment in the downstream units.

The two major heat losses in the system are at the surface condenser and in the exiting flue gases from the heat recovery and steam generation (HRSG) unit. A condensing turbine with 2.5 in. Hg abs. exhaust pressure is used for maximum power recovery. The surface condenser's duty represents about 28.48% of the coal heating value. The HRSG flue gases exit temperature of 275°F is selected to avoid sulfur condensation. The heat loss with the flue gases from the HRSG units represent 20.85% of the heating value of the coal feed.

All fuel gas produced goes to the combustion gas turbines except a portion needed to fire the combustors (16.8 MM Btu/hr) in the Beavon section of the tail gas treating unit. This represents an energy penalty of 0.16% of the coal feed HHV. The furnace in the sulfur plant is fired with acid gas from the Selexol unit.

ECONOMICS

Important economic results are summarized below.

TABLE MS-6

SUMMARY OF ECONOMICS - CASE MXSC

PRODUCTION AT DESIGN CAPACITY	
Net Power, MW (1)	1,211.86
Overall Plant Heat Rate, Btu/kWh	8,410
TOTAL CAPITAL (2)	
Total Capital @ $1/MM Btu Coal, $1000	861,052
Total Capital @ $1/MM Btu, $/kW	711
Total Capital @ $2/MM Btu Coal, $1000	878,167
Total Capital @ $2/MM Btu, $/kW	725
AVERAGE COST OF SERVICES (2)	
Annual Cost @ $1/MM Btu Coal, $1000/yr	243,474
Per Unit @ $1/MM Btu Coal, mills/kWh	32.79
Annual Cost @ $2/MM Btu Coal, $1000/yr	308,666
Per Unit @ $2/MM Btu Coal, mills/kWh	41.57

NOTES:

(1) At 100% Operating Load Factor.
(2) Mid-1976 Dollars and 70% Operating Load Factor.

Tables MS-7 and MS-8 give detailed breakdowns of plant investment, capital charges and working capital at 70% operating load factor for the two coal costs. Plant investment is independent of coal costs. Capital charges, however, are about 1 percent higher for the \$2.00/MM Btu coal and working capital 34 percent higher.

The accuracy of plant investment estimates is judged to be ±25%. Since other capital charges and working capital are keyed to elements of plant investment, this accuracy is reflected in other capital charges as well. Therefore, caution must be exercised in comparing Case MXSC with cases representing other gasification technologies.

The major element of plant investment is the Combined Cycle System. The Unit represents over 56 percent of the total plant investment. Process Condensate treating costs are lower for this case as compared to MACW. The steam to coal ratio is lower for Case MXSC when compared to MACW, resulting in less process condensate requiring waste water treatment.

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

Table MS-9 summarizes cost of services based on coal charged at $1.00/MM Btu and $2.00/MM Btu HHV. Costs are computed in accordance with criteria given by EPRI (Criteria section). They are presented as averages for the plant.

Operating labor requirement is a function of the number of units and trains. Requirements on a per shift basis are:

	CASE MXSC
"A" Operators	5
"B" Operators	19
Foremen	2
Lab and Instrument Technicians	4

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

Operating charges constitute nearly 45 percent of cost of services with coal at $1.00/MM Btu and about 56 percent at a coal cost of $2.00/MM Btu. Coal is the largest single operating charge, representing about 59 percent with coal at $1.00/MM Btu and 72 percent at the higher coal cost. The relationships as percentages are summarized below:

	CASE MXSC	
Coal Cost, $/MM Btu, HHV	1.00	2.00
Coal as % of Operating Charges	57.3	72.8
Coal as % of Total Cost of Services	25.7	40.5
Operating Charges as % of Total Cost of Services	44.8	55.6
Capital Charges as % of Total Cost of Services	55.2	44.4

TABLE MS-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR
AND $1.00/MM BTU COAL

CASE MXSC

	$1000 (1)	$/kW (2)	Percent
PLANT INVESTMENT			
Coal Handling	14,905	12.31	2.63
Oxidant Feed	57,938	47.85	10.23
Gasification and Ash Handling	49,628	40.99	8.76
Gas Cooling	8,936	7.38	1.58
Acid Gas Removal and Sulfur Recovery	23,498	19.41	4.15
Process Condensate Treating	31,169	25.74	5.50
Steam, Condensate and BFW	2,057	1.70	0.36
Support Facilities	58,099	47.98	10.25
Combined Cycle	320,400	264.60	56.54
Subtotal	566,630	467.96	100.00
Contingency	103,403	85.40	
TOTAL PLANT INVESTMENT	670,033	553.36	
ILLINOIS SALES TAX	15,641	12.92	
CAPITAL CHARGES			
Preproduction Costs	42,597	35.18	
Paid-up Royalties	3,350	2.77	
Initial Catalyst and Chemical Charges	800	0.66	
Construction Loan Interest	83,687	69.11	
TOTAL CAPITAL CHARGES	130,434	107.72	
DEPRECIABLE CHARGES	816,108	674.00	
WORKING CAPITAL	44,944	37.12	
TOTAL CAPITAL	861,052	711.12	

NOTES
(1) Mid-1976 dollars.
(2) Based on 100% Operating Load Factor.

TABLE MS-8

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR
AND $2.00/MM BTU COAL

CASE MXSC

	$1000 (1)	$/kW (2)	Percent
PLANT INVESTMENT			
Coal Handling	14,905	12.31	2.63
Oxidant Feed	57,938	47.85	10.23
Gasification and Ash Handling	49,628	40.99	8.76
Gas Cooling	8,936	7.38	1.58
Acid Gas Removal and Sulfur Recovery	23,498	19.41	4.15
Process Condensate Treating	31,169	25.74	5.50
Steam, Condensate and BFW	2,057	1.70	0.36
Support Facilities	58,099	47.98	10.25
Combined Cycle	320,400	264.60	56.54
Subtotal	566,630	467.96	100.00
Contingency	103,403	85.40	
TOTAL PLANT INVESTMENT	670,033	553.36	
ILLINOIS SALES TAX	15,641	12.92	
CAPITAL CHARGES			
Preproduction Costs	44,457	36.72	
Paid-up Royalties	3,350	2.77	
Initial Catalyst and Chemical Charges	800	0.66	
Construction Loan Interest	83,687	69.11	
TOTAL CAPITAL CHARGES	132,294	109.26	
DEPRECIABLE CAPITAL	817,968	675.54	
WORKING CAPITAL	60,199	49.72	
TOTAL CAPITAL	878,167	725.26	

NOTES
(1) Mid-1976 dollars.
(2) Based on 100% Operating Load Factor.

TABLE MS-9

COST OF SERVICES AT 70% OPERATING LOAD FACTOR

CASE MXSC

COAL COST, HHV	$1.00/MM Btu	$2.00/MM Btu
NET PRODUCTION (1)		
Net Power, MW	1,211.86	1,211.86
By-product Ammonia ST/SD	123.5	123.5
By-product Sulfur ST/SD	316	316
OPERATING CHARGES, $1000/YR		
Coal	62,522	125,044
Operating Labor	2,884	2,884
Catalyst and Chemicals	334	334
Utilities	1,197	1,197
Maintenance Labor	7,580	7,580
Maintenance Materials	11,370	11,370
Administration and Support Labor	3,139	3,139
General and Administration Expense	6,278	6,278
Ash Disposal	250	250
Property Tax and Insurance	16,751	16,751
By-product Ammonia	(3,155)	(3,155)
By-product Sulfur	(0)	(0)
TOTAL OPERATING CHARGES, $1000/yr	109,150	171,672
CAPITAL CHARGES, $1000/YR		
Total Capital Charges	134,324	136,994
COST OF SERVICES		
Total, $1000/yr	243,474	308,666
Per Unit Production, mills/kWh	32.79	41.57

NOTES

(1) At 100% Operating Load Factor.

CASES EAHC AND EXHC - FOSTER WHEELER

PLANT DESCRIPTION - CASES EAHC & EXHC

GENERAL

Grass roots plants for power generation using entrained bed gasifiers are shown schematically on Block Flow Diagrams EAHC-1-1 and EXHC-1-1. These plants are based on gasifying 10,000 ST/day of Illinois No. 6 coal. The first diagram (EAHC-1-1) represents a case where air is used as the oxidant in gasification. This case is referred to as Case EAHC. The second diagram (EXHC-1-1), identified as Case EXHC, represents the use of oxygen as the oxidant. The block flow diagrams show the major units in the plant, the number of operating units, major stream flows at 100 percent capacity operation and stream heat contents.

With the exception of the oxidant feed system, the overall processing schemes are similar for both air and oxygen. The main processing units are in four parallel operating trains in Case EAHC and three parallel operating trains in Case EXHC. One spare train is provided for the gasification and high temperature gas cooling areas in both cases. The parallel trains are largely independent. Each train consists of oxidant feed, gasification, particulate removal, gas cooling, and acid gas removal units. Integration between processing trains is minimum. Complete trains may be shut down in order to maintain efficiency during reduced capacity operation. The impact of an upset condition is limited to the train in which the upset occurs.

In addition to the main processing trains, each plant includes necessary offsite, utility and environmental facilities. Coal receiving, storage, grinding, drying and conveying is done in a single train to minimize space and operating labor requirements. Hydrogen sulfide removed from gasified coal is processed through sulfur recovery facilities which produce elemental sulfur. Other operating facilities in the plant are raw water treating, steam generation, cooling water, process condensate treating with ammonia by-product and effluent water treating. Support facilities to sustain an independent plant operation are provided as well. The Table EH-1 summarizes major equipment sections in the plant and shows the number of operating and spare sections.

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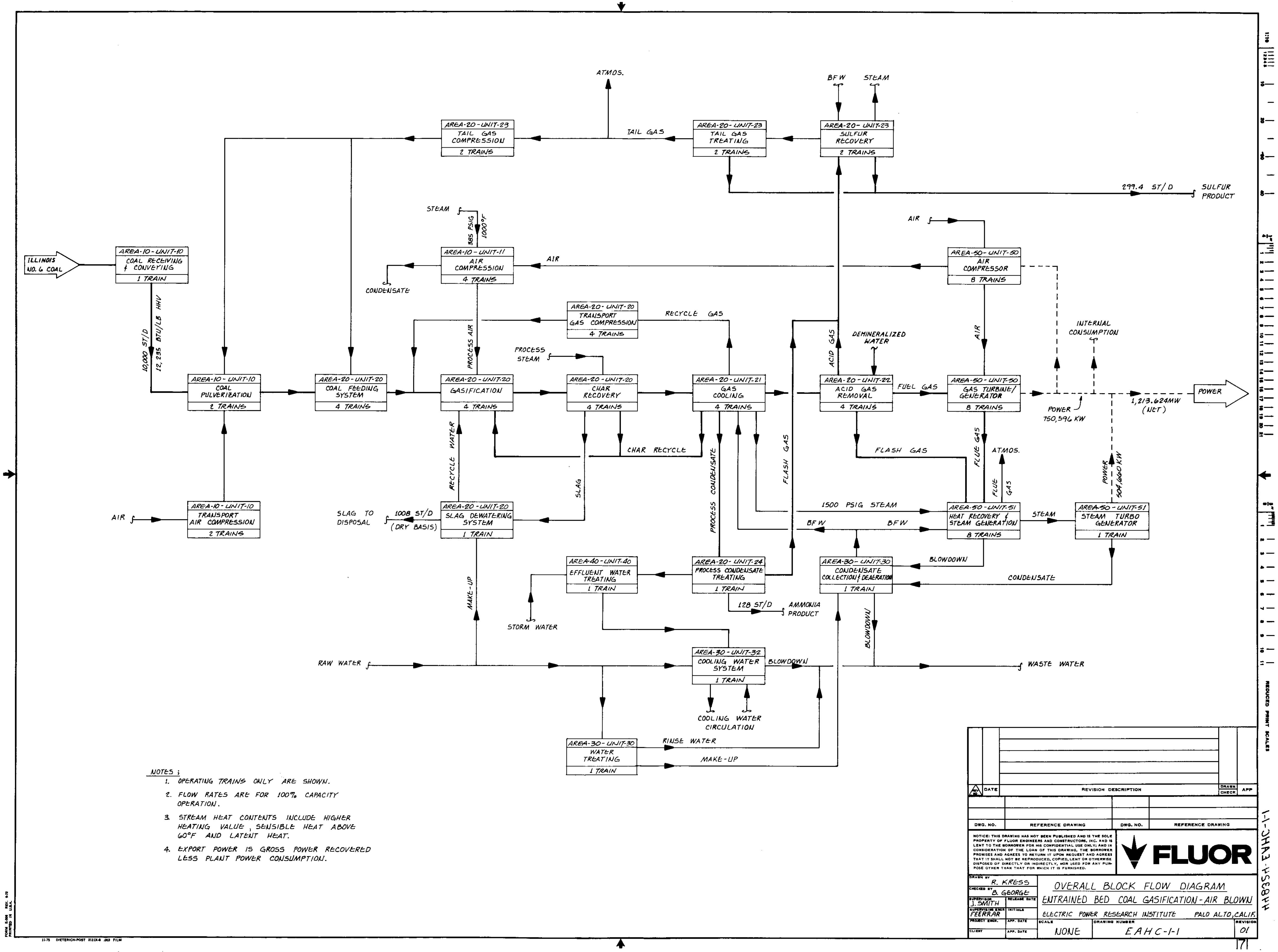

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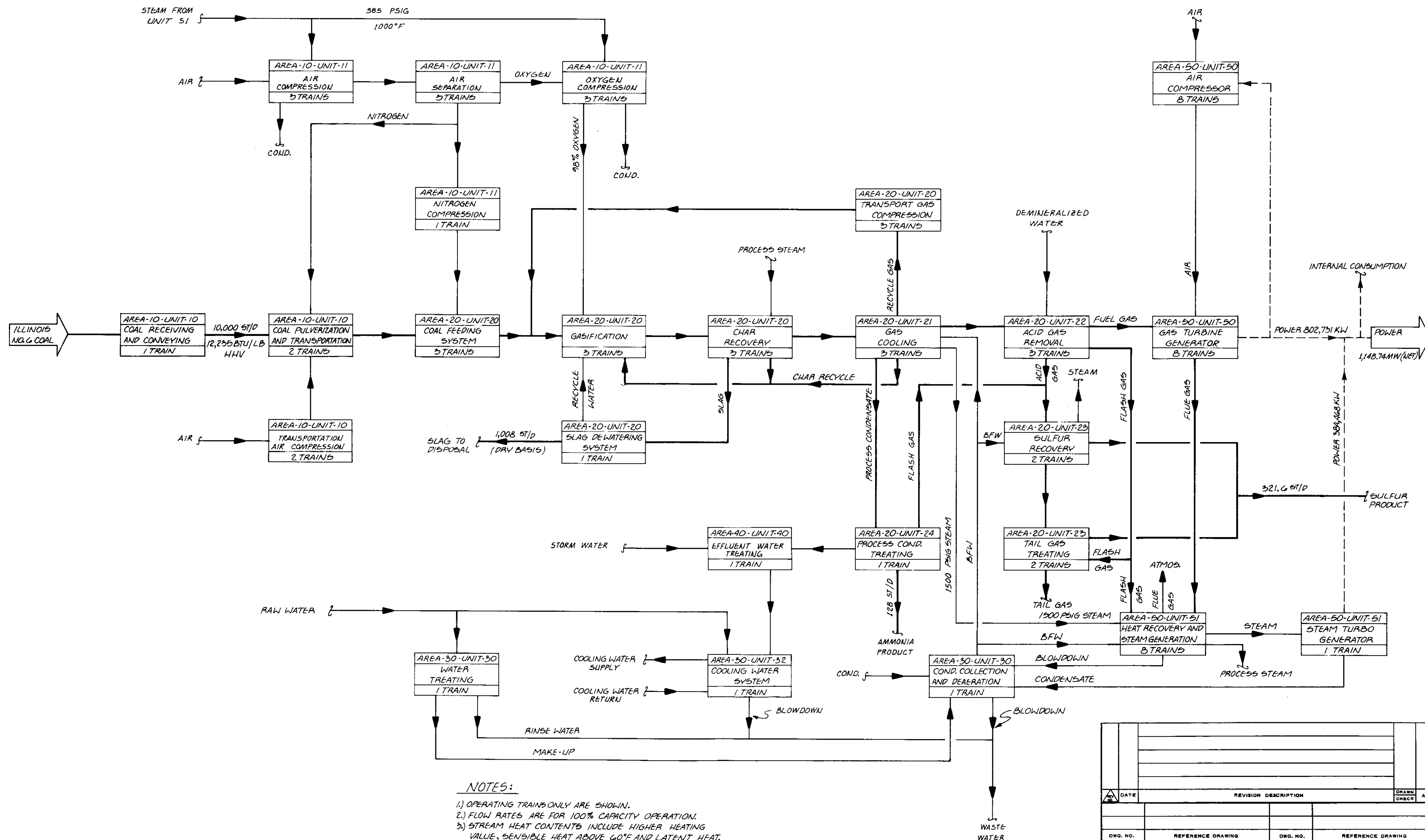

NOTES:

1.) OPERATING TRAINS ONLY ARE SHOWN.
2.) FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3.) STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE, SENSIBLE HEAT ABOVE 60°F AND LATENT HEAT.
4.) EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.

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CHECKED BY R. KRESS
SUPERVISOR J. SMITH
SUPERVISING ENGR. FEERRAR

OVERALL BLOCK FLOW DIAGRAM
ENTRAINED BED COAL GASIFICATION-OXYGEN BLOWN
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE | DRAWING NUMBER EXHC-1-1 | REVISION 01

44835A-EXHC-1-1

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TABLE EH-1

MAJOR EQUIPMENT SECTIONS - CASES EAHC AND EXHC

Unit No.	Unit Name	Case EAHC Operating	Case EAHC Spare	Case EXHC Operating	Case EXHC Spare
10	Coal Handling	1	0	1	0
10	Pulverization	2	0	2	0
11	Oxidant Feed System	4	0	3	0
20	Gasification	4	1	3	1
20	Ash Handling	1	0	1	0
21	Gas Cooling				
	. High Temperature	4	1	3	1
	. Particulate Scrubbing and Cooling	4	0	3	0
22	Acid Gas Removal	4	0	3	0
23	Sulfur Recovery and Tail Gas Treating	2	1	2	1
24	Process Condensate Treating	1	0	1	0
30	Steam, BFW and Condensate System				
	. Condensate Collection and Deaeration	1	0	1	0
	. Water Treating	1	0	1	0
32	Cooling Water System	1	0	1	0
40	Effluent Water Treating	1	0	1	0
50	Gas Turbine/Generator	8	0	8	0
51	Heat Recovery Steam Generator	8	0	8	0
51	Steam Turbine/ Generator	1	0	1	0

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COAL PREPARATION

Process Flow Diagrams EAHC-10-1 and EXHC-10-1 depict the process arrangement of coal preparation equipment in Case EAHC and EXHC, respectively. There is one coal handling train and two parallel coal pulverization trains.

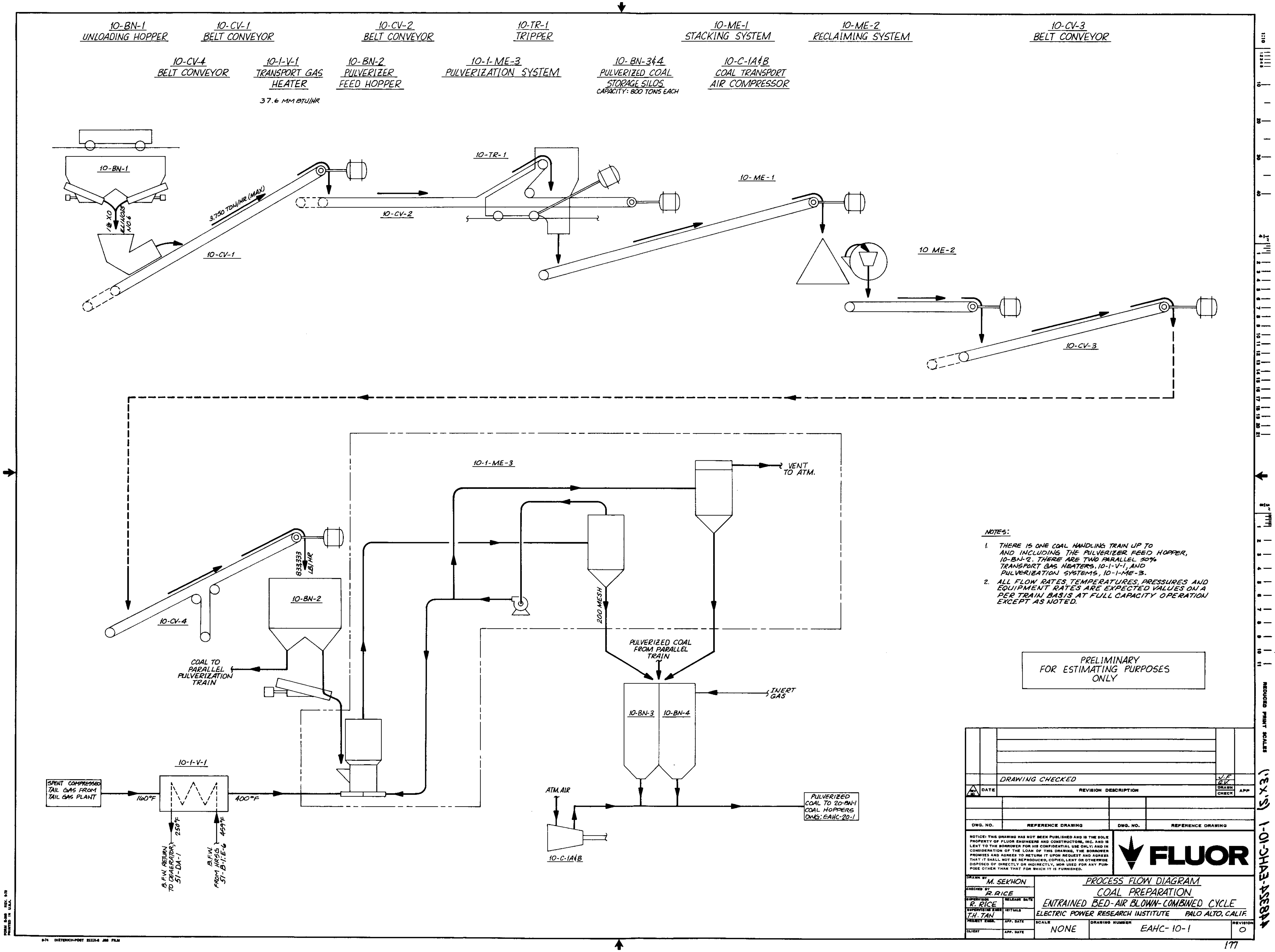
The basic process scheme for both the cases in this area is similar with the exception of the transport gas used for conveying coal through the pulverization system. Compressed tail gas from tail gas unit (Unit 23) is used for this purpose in Case EAHC. In Case EXHC, nitrogen from the air separation plants (Unit 11) is used as transport gas.

Washed, sized 1-1/2" by 0 coal is received at the plant site by unit train. The coal feed stock is unloaded from 100 ton bottom dump cars into an unloading hopper, withdrawn from the hopper by two vibrating feeders and transported by belt conveyors to a tripper conveyor. The tripper is attached to a traveling belt stacker. The stacker travels on tracks forming storage piles on either side. The unloading and stacking system is designed to handle a three-day supply in an eight-hour shift. No breaking and refuse disposal systems are included.

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

Coal is withdrawn from the pulverizer feed hopper by two vibrating feeders, each supplying coal to one package pulverization system, 10-1-ME-1. The coal is carried through 10-1-ME-1 by hot transport gas at 400°F. Transport gas also acts as a drying agent for the coal.

Compressed tail gas from sulfur recovery (Unit 10), or nitrogen from the air separation plants (Unit 11), respectively, is used as transport gas in Case EAHC or EXHC. The transport gas is preheated to 400°F by hot boiler feedwater (BFW) from the heat recovery steam generators (HRSG) in a transport gas heater, 10-1-V-1. BFW from the heater flows to the deaerator 51-DA-1 (Unit 51).


10-BN-1
UNLOADING HOPPER
10-CV-1
BELT CONVEYOR
10-CV-2
BELT CONVEYOR
10-TR-1
TRIPPER
10-ME-1
STACKING SYSTEM
10-ME-2
RECLAIMING SYSTEM
10-CV-3
BELT CONVEYOR
10-CV-4
BELT CONVEYOR
10-1-V-1
TRANSPORT GAS HEATER
37.6 MM BTU/HR
10-BN-2
PULVERIZER FEED HOPPER
10-1-ME-3
PULVERIZATION SYSTEM
10-BN-3&4
PULVERIZED COAL STORAGE SILOS
CAPACITY: 800 TONS EACH
10-C-1A&B
COAL TRANSPORT AIR COMPRESSOR
3,750 TON/HR (MAX)
1/8 XO ILLINOIS NO.6
833333 LB/HR
COAL TO PARALLEL PULVERIZATION TRAIN
VENT TO ATM.
200 MESH
PULVERIZED COAL FROM PARALLEL TRAIN
INERT GAS
10-BN-3
10-BN-4
ATM. AIR
PULVERIZED COAL TO 20-BN-1 COAL HOPPERS DWG: EAHC-20-1
SPENT COMPRESSED TAIL GAS FROM TAIL GAS PLANT
160°F
400°F
250°F
459°F
B.F.W. RETURN TO DEAERATOR, 51-DA-1
B.F.W. FROM HRSG 51-B-1:E-6
NOTES:
1. THERE IS ONE COAL HANDLING TRAIN UP TO AND INCLUDING THE PULVERIZER FEED HOPPER, 10-BN-2. THERE ARE TWO PARALLEL 50% TRANSPORT GAS HEATERS, 10-1-V-1, AND PULVERIZATION SYSTEMS, 10-1-ME-3.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
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G.V.
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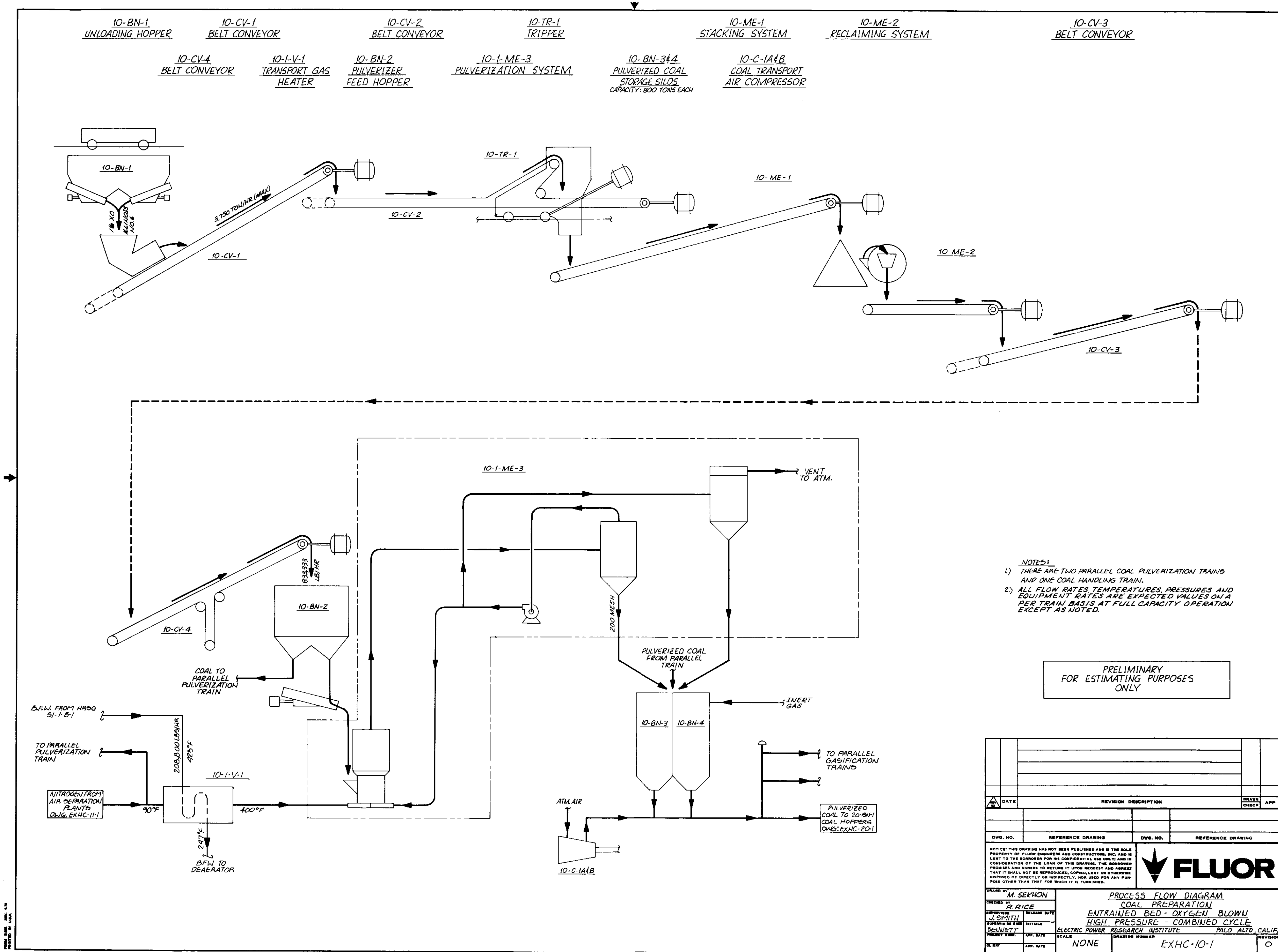
DRAWN BY M. SEKHON
CHECKED BY R. RICE
SUPERVISOR R. RICE
T.H. TAN
PROCESS FLOW DIAGRAM
COAL PREPARATION
ENTRAINED BED-AIR BLOWN-COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE
PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER EAHC-10-1
REVISION 0
REDUCED PRINT SCALES
4482324-EAHC-10-1 ('S',X3)



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The coal size is reduced in 10-1-ME-3 to at least 70 percent passing through 200 mesh screen. The pulverized coal is collected in the inert gas blanketed silos, 10-BN-3 and 4. Coal is withdrawn from the silos and pneumatically conveyed to the parallel gasifiers by compressed air supplied by coal transport air compressor.

Equipment Notes

All equipment is commercially available.

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OXIDANT FEED

Process Flow Diagrams EAHC-11-1 and EXHC-11-1 depict the oxidant feed system for Case EAHC and EXHC respectively. There are a total of four parallel trains of oxidant feed systems in Case EAHC. A total of three parallel oxidant feed trains are provided in Case EXHC. No spares are provided for the major equipment in either case.

Case EAHC

Bleed air from gas turbine air compressors (Unit 50) at 225 psig, 857°F is cooled to 544°F by intermediate pressure (IP) steam generation in a kettle boiler, 11-1-E-1. Approximately 43,400 lbs/hr of IP steam is generated and flows to the IP steam header.

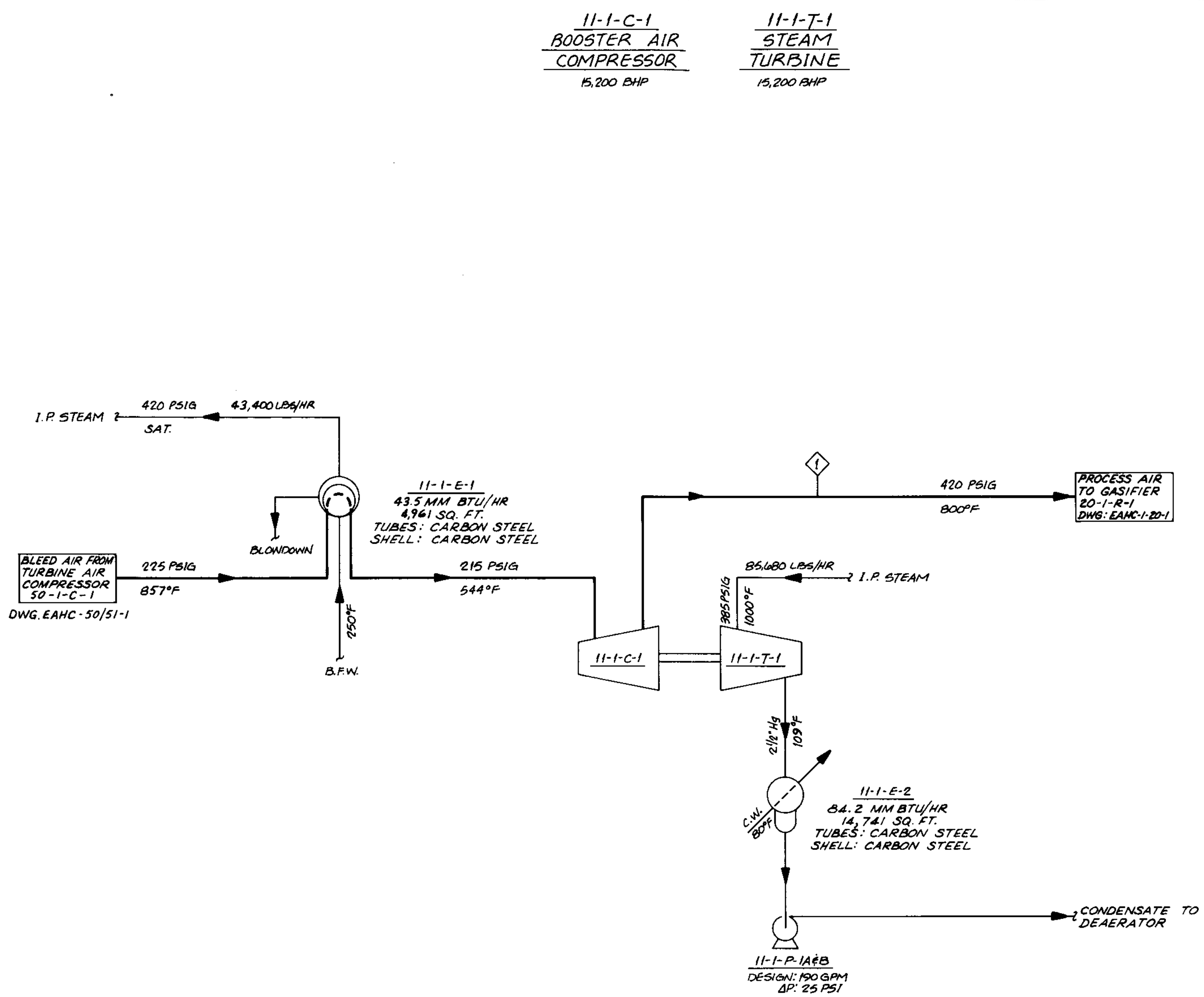
Bleed air is then compressed to 420 psig in a single stage booster air compressor, 11-1-C-1, and flows to the gasifiers. The 15,200 hp required by each compressor, 11-1-C-1, is supplied by condensing turbine drivers, 11-1-T-1. Steam to 11-1-T-1 is supplied by the HRSGs at 385 psig, 1000°F. The turbine exhausts at 2-1/2" Hg absolute and the vacuum condensate produced is pumped to the deaerator, 51-DA-1 (Unit 51).

Case EXHC

Each train provides one air compressor, one air separation plant and one oxygen compressor. Two reciprocating type nitrogen compressors (one operating and one spare) are provided for all three parallel trains.

Atmospheric air is compressed to 95 psig in a two-stage centrifugal machine, 11-1-C-1. Heat of compression is rejected to air in the interstage airfan cooler, 11-1-E-1.

The 39,500 hp required by each air compressor is supplied by a steam turbine, 11-1-T-1. The steam turbine driver is a condensing type machine operating at inlet steam conditions of 385 psig, 1000°F, with exhaust pressure at 2-1/2" Hg abs. The steam turbine is designed with excess capacity to provide response capabilities during turndown or upset conditions. No spare machine is provided for this service.



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ONLY

NOTES:

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

MATERIAL BALANCE

STREAM NUMBER	1	
	PROCESS AIR	
COMPONENTS	MPH	MOL %
O_2	4,150.0	20.53
N_2	15,611.8	77.00
H_2O	500.7	2.47
TOTAL MPH	20,262.5	100.00
TOTAL LB/HR	579,140	
MOL. WT.	28.58	

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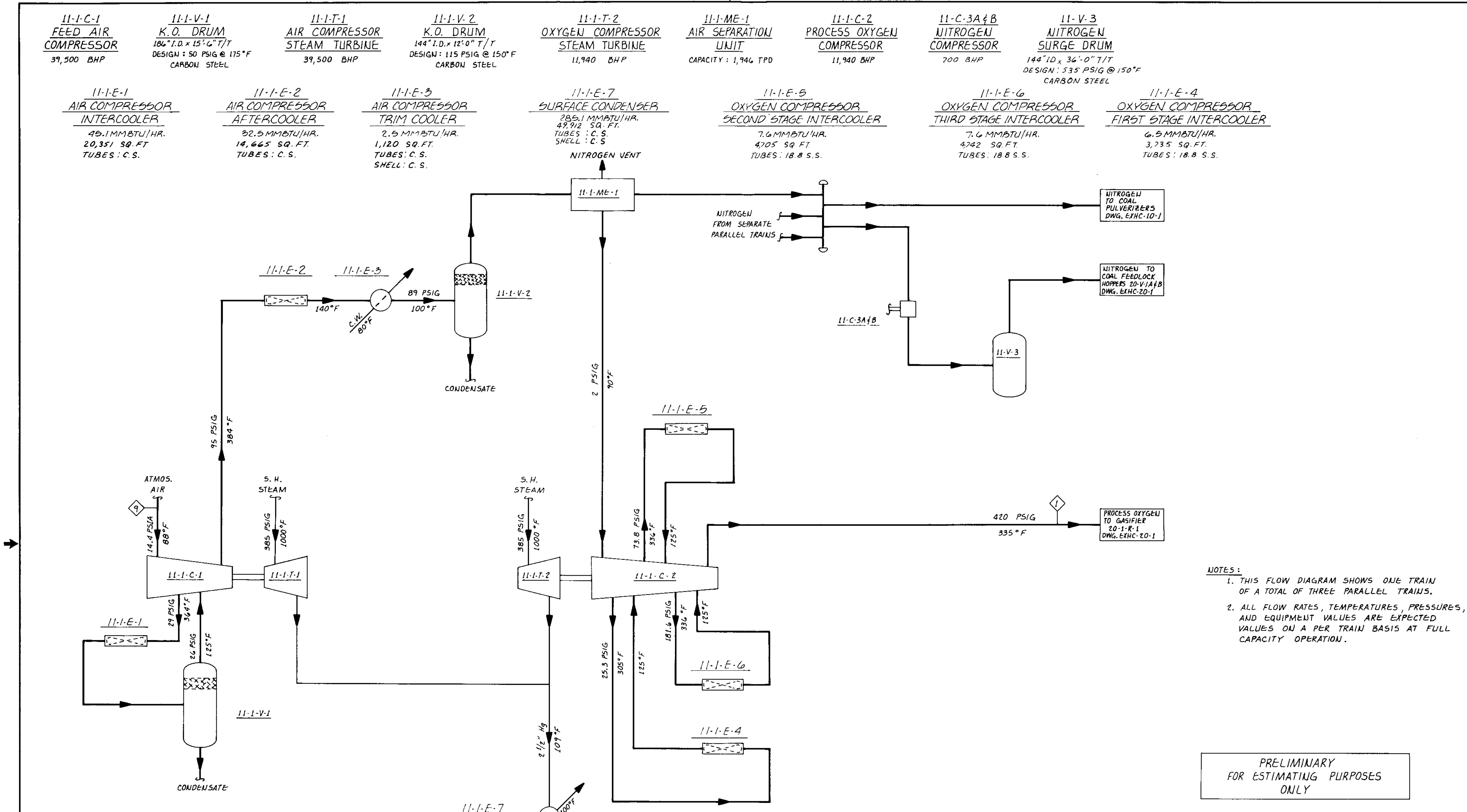
SUPERVISOR J. SMITH
RELEASE DATE
SUPERVISING ENGR. T.H. TAN
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM - ENTRAINED BED
AIR BLOWN - COMBINED CYCLE

ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: EAHC-11-1
REVISION: 02

4483SA-EAHC-11-1



MATERIAL BALANCE

STREAM NUMBER	9 FEED AIR		1 PROCESS OXYGEN	
COMPONENT	MPH	MOL %	MPH	MOL %
CH_4	—	—	—	—
H_2	—	—	—	—
CO	—	—	—	—
CO_2	—	—	—	—
H_2S	—	—	—	—
COS	—	—	—	—
NH_3	—	—	—	—
N_2	20,639.3	77.00	103.4	2.00
O_2	5,486.4	20.47	5,067.3	98.00
H_2O	678.1	2.53	—	—
TOTAL MPH	26,803.8	100.00	5,170.7	100.00
TOTAL LB/HR	766,013		165,051	
MOL. WT.	28.58		31.92	

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FLUOR

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CHECKED BY
SUPERVISOR J. SMITH — RELEASE DATE
SUPERVISING ENGR R. BENNETT — INITIALS
PROJECT ENGR. — APP. DATE
CLIENT — APP. DATE

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
ENTRAINED BED - OXYGEN BLOWN
HIGH PRESSURE - COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CA.

SCALE NONE — DRAWING NUMBER EXHC-11-1 — REVISION 01

4483A-EXHC-11-1

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The compressed air is processed in an air separation unit, 11-1-ME-1, which produces 1946 tons per day (100% basis) of 98% oxygen. Liquid oxygen storage of 5888 tons is provided, with attendant cryogenic pumps and vaporizer. Storage is equivalent to approximately three days of rated capacity operation of a single train. The three days of storage is anticipated to adequately cover any outage of the cryogenic unit.

The air separation plant produces oxygen at 2 psig and 90°F. The oxygen is compressed to 420 psig in four stages. As in the case of the air feed compressor, interstage heat of compression is rejected to the interstage air coolers, 11-1-E-4 through 6. The 11,940 hp compression requirement is supplied by a condensing type turbine. The inlet steam condition is 385 psig, 1000°F with backpressure at 2-1/2" Hg abs.

Nitrogen from the parallel air separation plants flows to a common nitrogen header. A portion of the nitrogen is compressed in a single stage reciprocating nitrogen compressor (11-C-3A or B) and transported to coal feed lock hoppers (Flow Diagram: EXHC-20-1). The balance flows as transport gas to the upstream coal preparation area (Flow Diagram: EXHC-10-1).

Equipment Notes

The air booster compressor (11-1-C-1) in Case EAHC is a nonlubricated machine and represents an extension of the present state of art for these compressors. The current experience for such machines is limited to the units with approximately 5000 bhp and a discharge temperature of (maximum) 650°F. The major areas for the development effort will be metallurgy and the shaft and sealing arrangements. Considerable metallurgical information is available for gas expanders operating with an inlet temperature up to 1450°F. This knowledge will need to be applied to these compressors.

The air and oxygen compressors and cryogenic air separation plant required for Case EXHC are commercially available.

The condensing turbines with 1000°F inlet temperature represent an extension of the present state of art for turbines. It is expected that such turbines would be available in the next few years.

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GASIFICATION AND ASH HANDLING

Process Flow Diagrams EAHC-20-1 and EXHC-20-1 depict the process arrangement of equipment in this area in Cases EAHC and EXHC, respectively.

There are five parallel trains of gasifiers in Case EAHC (four operating and one spare). Case EXHC consists of three operating and one spare train. There is one ash handling train in both cases.

Case EAHC

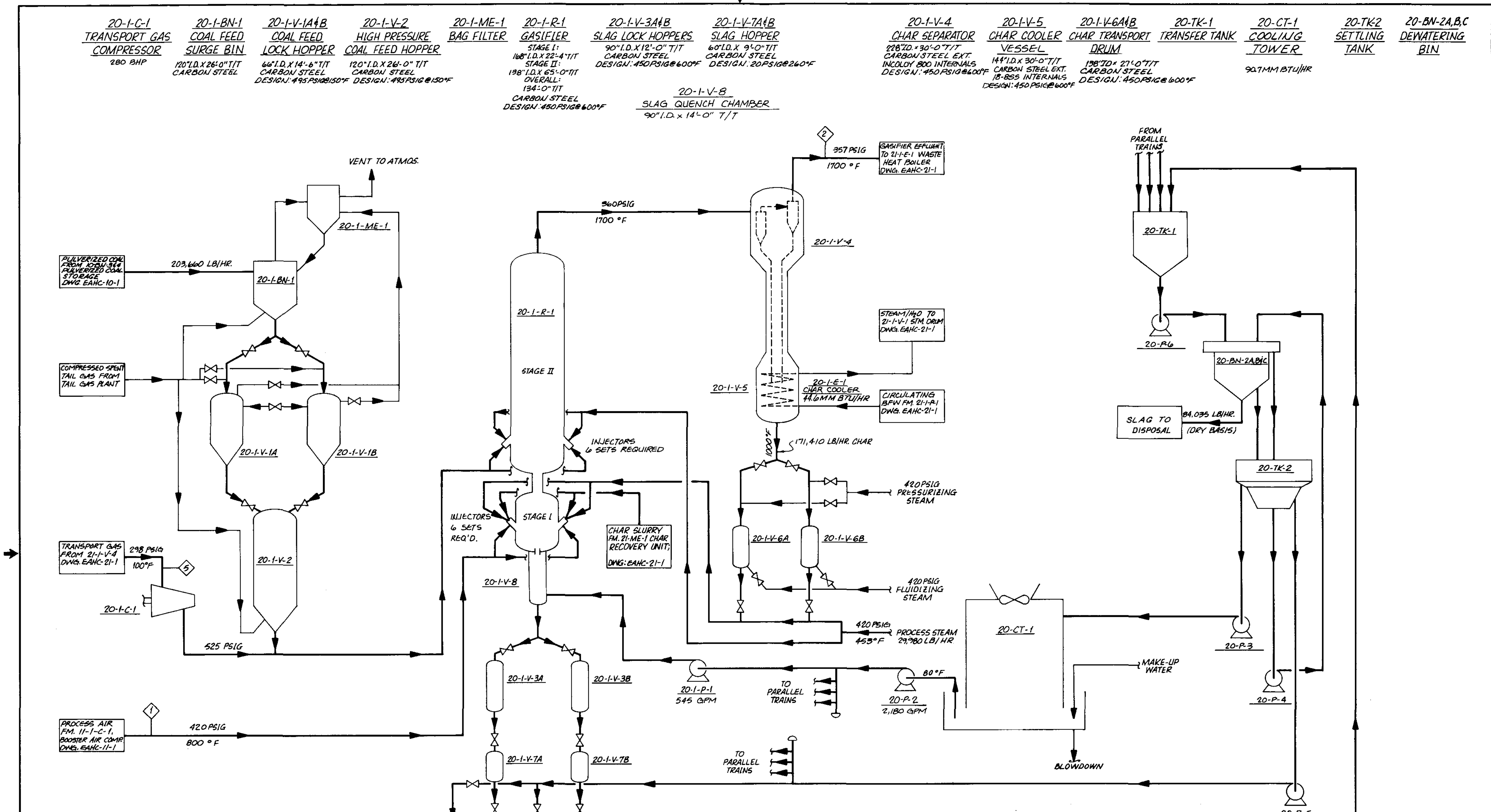
Pulverized coal is pneumatically conveyed from the coal preparation area to a coal feed bin, 20-1-BN-1, from which it is fed by gravity to one of the depressurized coal feed lock hoppers, 20-1-V-1A or B, through hydraulically operated valves. Both the coal feed lock hoppers feed a common high pressure coal feed hopper, 20-1-V-2.

Compressed desulfurized tail gas from sulfur recovery unit (Unit 23) is used to pressurize 20-1-V-1A and B and continuously maintain the pressure of 20-1-V-2 at about 525 psig.

One coal feed lock hopper (20-1-V-1A) at a time is depressurized and receives coal from 20-1-BN-1. The other coal feed lock hopper (20-1-V-1B) is under pressure and feeds coal to 20-1-V-2. After 20-1-V-1B is emptied it is depressurized by venting gases through a common bag filter assembly, 20-1-ME-1, to atmosphere. At this time the cycle is switched, 20-1-V-1B starts receiving the coal from 20-1-BN-1 and 20-1-V-1A is pressurized and starts feeding coal to 20-1-V-2. Necessary valving is provided to perform this operation.

Pulverized coal from 20-1-V-2 is transported to the second stage of the gasifier by compressed crude gas from the transport gas compressor, 20-1-C-1. Crude overhead gas from the ammonia scrubber, 21-1-V-4 (Flow Diagram: EAHC-21-1), is used as transport gas.

The gasifier is a two stage cylindrical vessel with the stages joined by a connecting pipe. Partial gasification of incoming coal is carried out in the second (upper) stage. Unreacted char leaving the top of the second stage is gasified in the first (lower) stage. Each stage has a set of injection nozzles for feeding coal or char and steam and air.



MATERIAL BALANCE

STREAM NUMBER	1 PROCESS AIR		2 GASIFIER EFFLUENT		5 TRANSPORT GAS	
COMPONENT	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	—	—	1,123.5	3.37	25.8	3.44
H_2	—	—	4,558.0	13.68	104.9	13.98
CO	—	—	9,580.3	28.75	220.5	29.37
CO_2	—	—	1,010.7	3.03	23.3	3.10
H_2S	—	—	223.6	0.67	5.2	0.69
COS	—	—	22.4	0.07	0.5	0.07
N_2	15,611.8	77.00	15,998.0	48.02	368.3	49.05
NH_3	—	—	142.6	0.43	—	—
O_2	4,150.0	20.53			—	—
H_2O	500.7	2.47	658.5	1.98	2.3	0.30
TOTAL MPH	20,262.5	100.00	33,317.5	100.00	750.8	100.00
TOTAL VAPOR LB/HR	579,140		811,440		18,390	
CHAR CARRYOVER LB/HR	—		860		—	
TOTAL LB/HR	579,140		812,300		18,390	
MOL. WT. VAPOR	28.58		24.35		24.50	

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ONLY

NOTES
1. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
2. THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FIVE PARALLEL TRAINS UP TO THE SLAG SLURRY DEWATERING SYSTEM, ONE TRAIN BEING A SPARE. THERE IS ONE SLURRY DEWATERING SYSTEM.

REV NO	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP
		DRAWING CHECKED	J.F. / R.R.	

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

FLUOR

DRAWN BY M. SEKHON
CHECKED BY R. RICE
SUPERVISOR R. RICE
SUPERVISING ENGR T.H. TAN

PROCESS FLOW DIAGRAM
GASIFICATION AND ASH HANDLING
ENTRAINED BED-AIR BLOWN-COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE | DRAWING NUMBER: EAHC-20-1 | REVISION: 02

44854-EAHC-20-1 (15-0X3-0)

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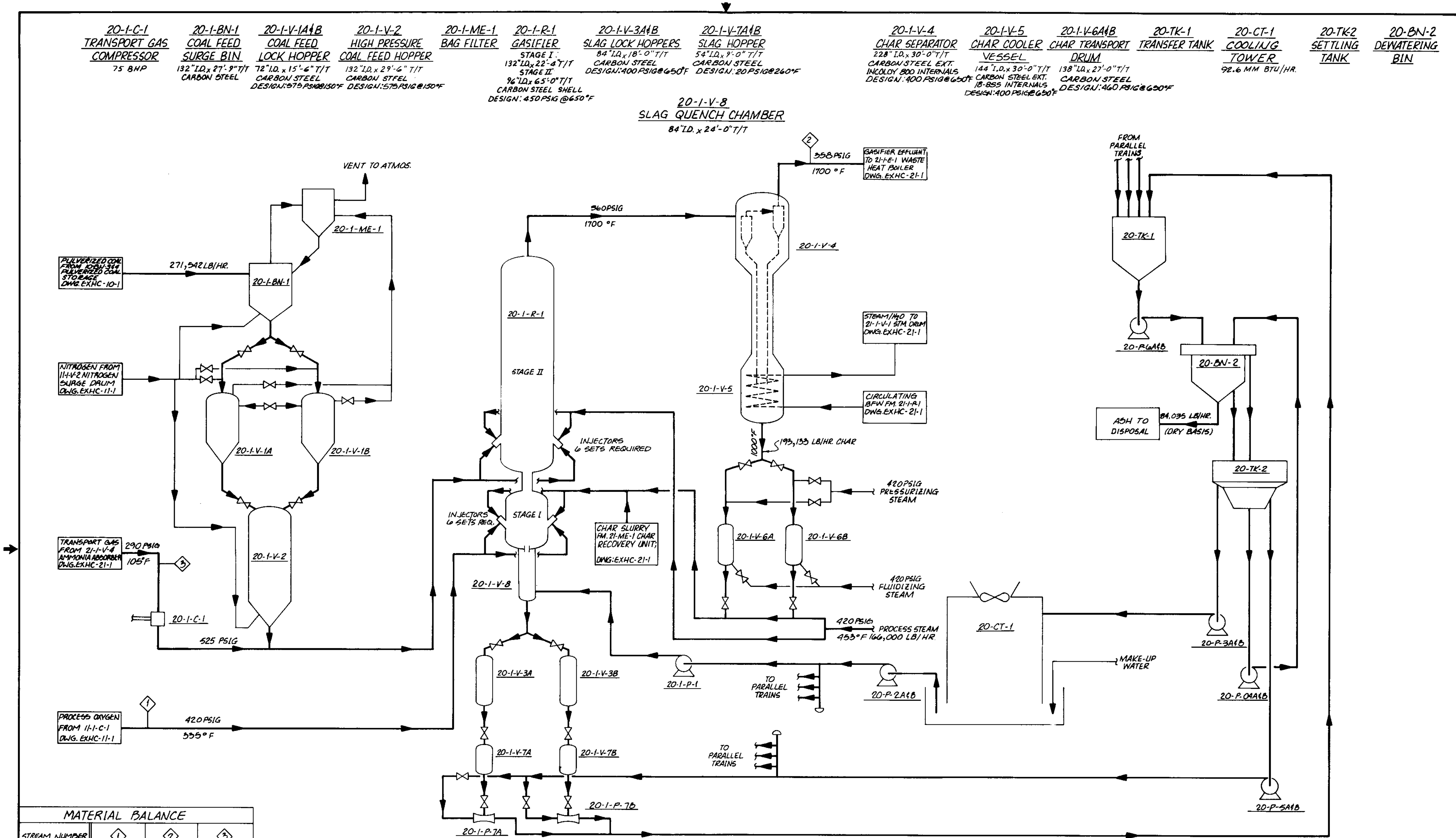

MATERIAL BALANCE						
STREAM NUMBER	1		2		3	
	PROCESS OXYGEN		GASIFIER EFFLUENT		TRANSPORT GAS	
COMPONENT	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	—	—	1,794.4	6.03	84.3	7.08
H_2	—	—	8,768.9	29.50	412.0	34.61
CO	—	—	10499.5	35.34	493.2	41.43
CO_2	—	—	3,718.1	12.51	174.6	14.67
H_2S	—	—	305.7	1.03	14.4	1.21
COS	—	—	30.5	0.10	1.4	0.12
N_2	103.4	2.00	133.5	0.45	6.3	0.53
NH_3	—	—	190.1	0.64	—	—
O_2	5,067.3	98.00	—	—	—	—
H_2O	—	—	4,280.5	14.40	4.2	0.35
TOTAL VAPOR	5,170.7	100.00	29,721.2	100.00	1,190.4	100.00
TOTAL VAPOR LB/HR	165,051		600,551		24,510	
CHAR CARRYOVER LB/HR	—		970		—	
TOTAL LB/HR	165,051		601,521		24,510	
MOL. WT.	31.92		20.21		20.59	

PRELIMINARY
FOR ESTIMATING PURPOSES
ONLY

NOTES:

1.) THERE ARE THREE PARALLEL AND ONE SPARE GASIFICATION TRAINS AND ONE ASH HANDLING UNIT.

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY M. SEKHON
CHECKED BY R. RICE
SUPERVISOR J. SMITH
SUPERVISING ENGR. BENNETT

PROCESS FLOW DIAGRAM
GASIFICATION AND ASH HANDLING
ENTRAINED BED - OXYGEN BLOWN
HIGH PRESSURE - COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE | DRAWING NUMBER EXHC-20-1 | REVISION 01

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The coal is introduced into the second stage through a set of injection nozzles. Atomizing steam is also fed to these nozzles. The hot gases rising from the first stage entrain and partially gasify the incoming coal. The resulting product gas, containing char, exits the top of the gasifier at approximately 1700°F and flows through a char separator, 20-1-V-4. The char separator vessel is equipped with internal cyclones which remove the char. The hot gasifier effluents at 357 psig and 1700°F flow to the waste heat boiler, 21-1-E-1 (Flow Diagram: EAHC-21-1), for heat recovery. The system designed for this study anticipated that oils and tars would not be produced in the gasifier. This assumption must be checked when operating data become available.

The hot char flows from the cyclones in 20-1-V-4, by gravity, to the char cooler vessel, 20-1-V-5. The char separator and char cooler are provided in the same vessel. Char is cooled to about 1000°F in 20-1-V-5 by heat exchange against saturated high pressure steam generation in 20-1-E-1. The cooled char then flows to either of the char transport drums, 20-1-V-6A and B.

Char is kept in a continuous state of fluidization in 20-1-V-6A, B by the injection of 420 psig intermediate pressure (IP) steam.

One drum at a time (20-1-V-6A) receives char from 20-1-V-5 while the other (20-1-V-6B) is under pressure and feeds the char transport header to the first stage of the gasifier. IP steam is used for pressurizing the drums.

When 20-1-V-6B is emptied, the cycle is switched and it starts receiving char from 20-1-V-5 while 20-1-V-6A is pressurized by IP steam to feed the char transport header. Necessary valving is provided to perform this operation.

A stream of char slurry from the char slurry recovery unit, 21-ME-1 (Flow Diagram: EAHC-21-1), is also fed continuously to the first stage through the injection nozzles.

Char is fed to the first stage of the gasifier through a set of injection nozzles by IP steam which also supplies the process steam needed for the gasification reactions. Process air from the booster air compressor, 11-1-C-1 (Flow Diagram: EAHC-11-1), is also supplied to the first stage of the gasifier through the char injection nozzles. Char is combusted and gasified in the first stage and the hot product gases and unconsumed char flow upward to the second stage to react with the incoming coal.

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The first stage operates above the ash fusion temperature. The ash is melted to form slag. The circular motion induced in the first stage chamber forces the slag radially outward against the vessel walls. The slag forms two layers on the vessel walls. The inner layer against the refractory lining is solidified. The outer layer is molten at a temperature high enough to maintain a viscosity of approximately 20 poises. This allows it to flow by gravity to the taphole.

Slag flows from the taphole to a quench chamber, 20-1-V-8, where it is quenched by water to form fine granular particles. The slag slurry from 20-1-V-8 then flows through two parallel slag lock hoppers, 20-1-V-3A, B, to the parallel slag hoppers, 20-1-V-7A and B, and is educted to a common transfer tank, 20-TK-1, using water as motive fluid.

The slag is pumped from 20-TK-1 to the bins, 20-1-BN-2A, B, C where it is dewatered to produce slag ready for disposal. Final cleaning of the water overflowing the dewatering bins is accomplished in a settling tank, 20-TK-2, where ash fines settle and are pumped back to the dewatering bins. A portion of the clarified water is recycled to the slag quench chambers after it is cooled in an induced draft-type cooling tower (20-CT-1). The balance of the water provides the motive fluid for the slag slurry transfer eductors.

<u>Case EXHC</u>

The basic process scheme for the gasification and ash handling area for Case EXHC is similar to Case EAHC with the following exceptions.

- 98% oxygen is used as oxidant instead of air.
- Compressed nitrogen from the nitrogen compressors, 11-1-C-3A and B (Flow Diagram: EXHC-11-1), is used for coal feed lock hopper operation instead of tail gas from the sulfur recovery unit in Case EAHC.
- Water cooling of the second stage is required. Cooling coils are located between the vessel walls and refractory lining with metal studs projecting through the refractory.

The ash handling system is similar to the one described for Case EAHC.

Equipment Notes

The Foster Wheeler gasifier is based on the work performed by Bituminous Coal Research, Inc. Construction of an 80 ton/day BIGAS unit at Homer City, Pennsylvania was completed early in 1977. This SNG pilot plant is being funded by the Energy Research and Development Administration and American Gas Association. Although the oxygen blown BIGAS reactor is designed for higher pressure operation than the Foster Wheeler gasifier, data from this pilot plant will be invaluable in evaluating the operability of the Foster Wheeler concept.

As of summer 1977, the BIGAS unit has not been operated completely on coal. It has been operated with coal feed to the second stage and natural gas fired in the bottom stage. Therefore, it must be realized that there are no data yet available to confirm the operating and design estimates used for this study.

GAS COOLING

Process Flow Diagrams EAHC-21-1 and EXHC-21-1 depict the process arrangement in this section for Cases EAHC and EXHC, respectively.

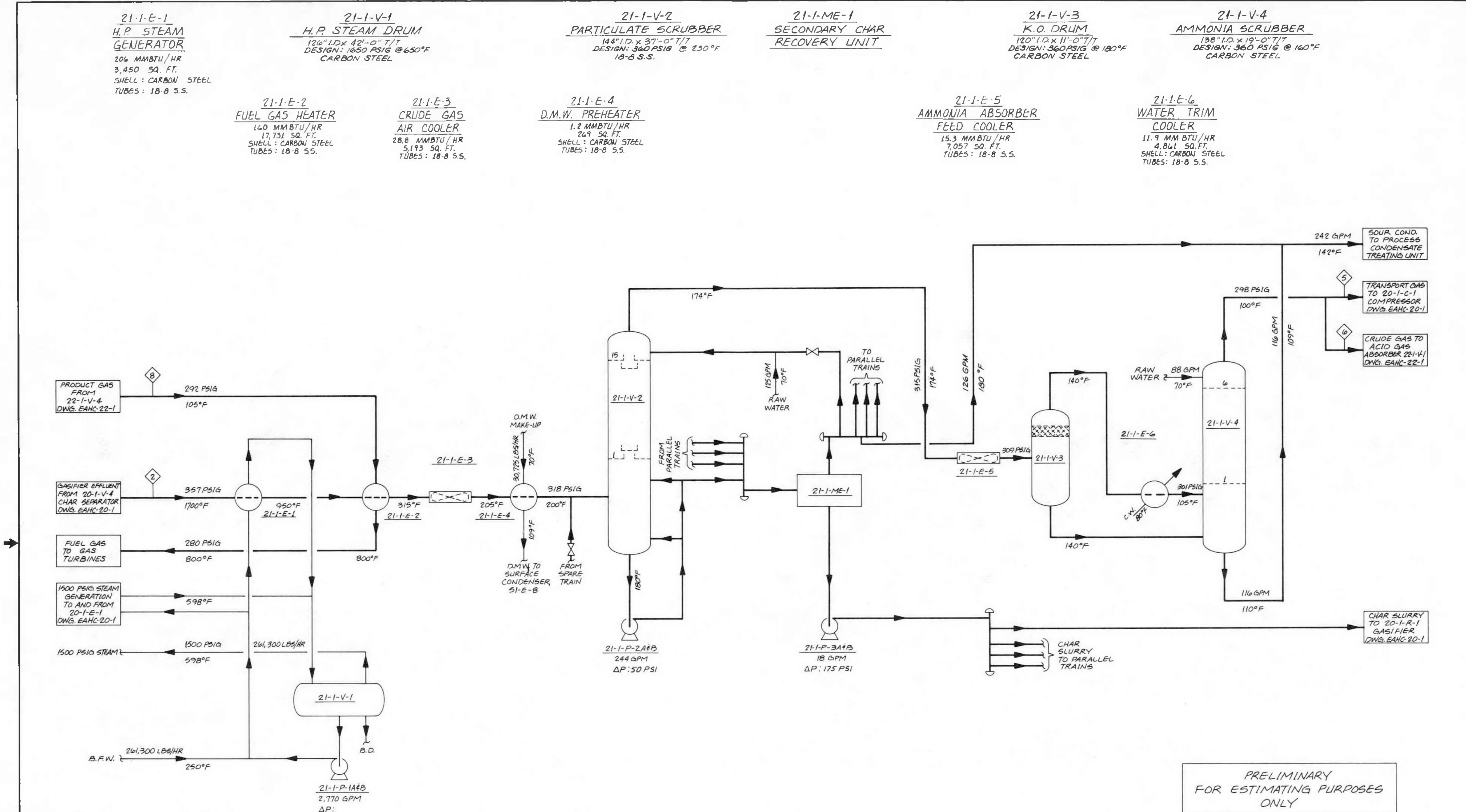
There are a total of five (four operating, one spare) high temperature gas cooling trains and four parallel particulate scrubbing and cooling trains in Case EAHC. For the Case EXHC, there are four parallel trains (three operating, one spare) for high temperature gas cooling and three parallel operating trains for the particulate scrubbing and low temperature gas cooling sections.

Case EAHC

The hot crude gas from the char separator, 20-1-V-4, at 357 psig and 1700°F is cooled to 950°F on the tube side of a shell and tube waste heat boiler, 21-1-E-1. High pressure (HP) saturated steam is generated on the shell side of 21-1-E-1. The HP steam flows to the heat recovery steam generators (HRSGs) located in Unit 51.

The crude gas then exchanges heat with cold product gas in a shell and tube exchanger, 21-1-E-2, and is cooled down to about 315°F. The product gas is heated to 800°F and flows to the gas turbine, 50-1-GT-1. Further cooling of the crude gas to approximately 200°F is obtained by a combination of air (21-1-E-3) and water (21-1-E-4) cooling. Makeup demineralized water is heated in 21-1-E-4 and flows to the surface condenser, 51-E-8. The cooled crude gas at 318 psig and 200°F flows to a particulate scrubber, 21-1-V-2, where it is scrubbed countercurrently with cold (70°F) raw water on trays.

The char slurry produced in the scrubber is continuously recirculated back to the scrubber by the pumps, 21-1-P-2A or B. A slipstream of the char slurry is withdrawn from the discharge line of 21-1-P-2A and B and combined with the char slurry streams from the other trains. Further concentration of the combined char slurry is accomplished in a secondary char recovery unit, 21-1-ME-1. The concentrated char slurry from 21-1-ME-1 is pumped back to the gasifiers, 20-1-R-1, through 21-1-P-3A or B. A portion of the effluent water from 21-1-ME-1 flows to the top section of the particulate scrubbers, 20-1-V-2, in the parallel trains. The balance flows to the sour water header. The clean crude gas from 21-1-V-2 is cooled to approximately 140°F in an air cooler, 21-1-E-5. The condensate produced



MATERIAL BALANCE

STREAM NUMBER	2 GASIFIER EFFLUENT		5 TRANSPORT GAS		6 CRUDE GAS		8 PRODUCT GAS	
COMPONENT	MPH	MOL%	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	1,123.5	3.37	25.8	3.44	1,097.7	3.44	1,078.1	3.43
H_2	4,558.0	13.68	104.9	13.98	4,453.0	13.98	4,442.1	14.15
CO	9,580.3	28.75	220.5	29.37	9,359.8	29.37	9,310.2	29.66
CO_2	1,010.7	3.03	23.3	3.10	987.4	3.10	840.1	2.68
H_2S	223.6	0.67	5.2	0.69	218.5	0.69	16.4	0.05
COS	22.4	0.07	0.5	0.07	21.9	0.07	14.8	0.05
N_2	15,998.0	48.02	368.3	49.05	15,629.7	49.05	15,575.0	49.62
NH_3	142.6	0.43	—	—	—	—	—	—
H_2O	658.5	1.98	23	0.30	96.9	0.30	113.0	0.36
TOTAL MPH	33,317.5	100.00	750.8	100.00	31,864.7	100.00	31,389.7	100.00
TOTAL VAPOR LB/HR	811,440		18,390		780,650		763,700	
CHAR CARRYOVER LB/HR	860		—		—		—	
TOTAL LB/HR	812,300		18,390		780,650		763,700	
MOL. WT. VAPOR	24.35		24.50		24.50		24.33	

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NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FOUR PARALLEL TRAINS. EQUIPMENT UP TO AND INCLUDING 21-1-E-4 IS SPARED IN ONE ADDITIONAL PARALLEL TRAIN.
2. ALL FLOW RATES, TEMPERATURES AND PRESSURES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY B. GUEVARA
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
SUPERVISING ENGR. T.H. TAN

PROCESS FLOW DIAGRAM
GAS COOLING & PARTICULATE REMOVAL
ENTRAINED BED-AIR BLOWN-
COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE | DRAWING NUMBER: EAHC-21-1 | REVISION: 02

AAAEBDA-EAHC-21-1 (S) (XE')

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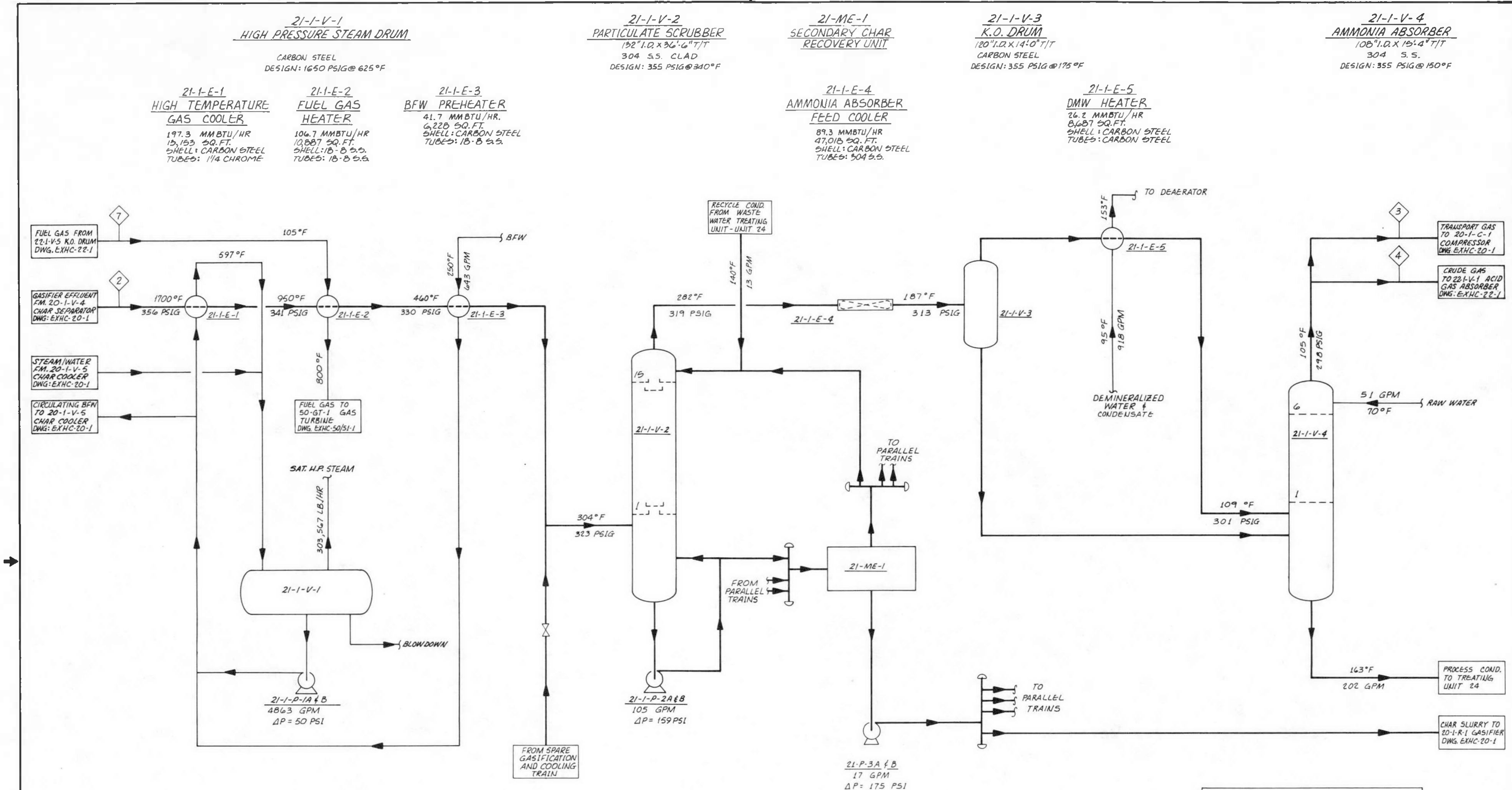

PRELIMINARY
FOR ESTIMATING PURPOSES
ONLY

MATERIAL BALANCE								
STREAM NUMBER	2 GASIFIER EFFLUENT		3 TRANSPORT GAS		4 CRUDE GAS		7 FUEL GAS	
COMPONENT	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	1,794.4	6.03	84.3	7.08	1,710.0	7.08	1,679.8	7.29
H_2	8,768.9	29.50	412.0	34.61	8,356.9	34.60	8,331.5	36.14
CO	10,499.5	35.34	493.2	41.43	10,006.3	41.43	9,935.7	43.09
CO_2	3,718.1	12.51	174.6	14.67	3,543.5	14.67	2,863.8	12.42
H_2O	4,280.5	14.40	4.2	0.35	85.8	0.36	83.5	0.36
H_2S	305.7	1.03	14.4	1.21	291.3	1.21	17.3	0.08
COS	30.5	0.10	1.4	0.12	29.1	0.12	16.9	0.07
NH_3	190.1	0.64	—	—	—	—	—	—
N_2	133.5	0.45	6.3	0.53	127.2	0.53	126.9	0.55
O_2	—	—	—	—	—	—	—	—
TOTAL VAPOR	29,721.2	100.00	1,190.4	100.00	24,150.1	100.00	23,055.4	100.00
MOL. WT.	20.21		20.59		20.59		19.72	
LB/HR VAPOR	600,551		24,510		497,307		454,757	
LB/HR CHAR	970		—		—		—	
TOTAL LB/HR	601,521		24,510		497,307		454,757	

NOTE:
MATERIAL BALANCE IS ON A PER TRAIN BASIS AT 100% OF CAPACITY OPERATION.

NOTES:
1. THIS DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE PARALLEL TRAINS. EQUIPMENT UP TO & INCLUDING 21-1-E-3 IS SPARED IN ONE ADDITIONAL PARALLEL TRAIN.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

REV NO	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY C. VALDEZ
CHECKED BY R. RICE
SUPERVISOR J. BARGER
SUPERVISING ENGR R. BENNETT
RELEASE DATE
INITIALS
PROJECT ENGR. APP. DATE
CLIENT APP. DATE

PROCESS FLOW DIAGRAM
GAS COOLING & PARTICULATE REMOVAL
ENTRAINED BED - OXYGEN BLOWN
HIGH PRESSURE - COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE | DRAWING NUMBER EXHC-21-1 | REVISION 01

4883A-EXHC-21-1

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in gas cooling is separated in a knockout drum, 21-1-V-3, and flows to the bottom section of the ammonia absorber, 21-1-V-4. The overhead gas from 21-1-V-3 is further cooled to approximately 105°F in the water cooler, 21-1-E-6, and flows to the bottom of the ammonia scrubber, 21-1-V-4.

The ammonia scrubber is a trayed column. Final removal of ammonia from the crude gas is obtained by scrubbing the crude gas countercurrently with cold (70°F) raw water over trays. A portion of the ammonia free crude gas from the top of the ammonia scrubber flows to the transport gas compressor, 20-1-C-1, located in the gasification unit (Flow Diagram: EAHC-20-1). The balance of the crude gas from 21-1-V-4 flows to the downstream acid gas removal unit (Flow Diagram: EAHC-22-1) for further processing. The scrubber bottoms after combining with the sour water from 21-1-ME-1 flows to the process condensate treating unit.

Case EXHC

The hot crude gas from the char separator, 20-1-V-4 at 356 psig and 1700°F is cooled to 950°F on the tube side of a shell and tube waste heat boiler, 21-1-E-1. High pressure (HP) steam is generated on the shell side of 21-1-E-1. The HP steam flows to the heat recovery steam generators (HRSG) units located in Unit 51.

The crude gas then exchanges heat with the cold fuel gas in a shell and tube exchanger, 21-1-E-2, and is cooled down to about 330°F. The fuel gas is heated to 800°F and flows to the gas turbines, 50-1-GT-1, located in Unit 50. Further cooling of the crude gas to approximately 304°F is obtained in 21-1-E-3 against high pressure boiler feedwater heating. The crude gas then flows to a particulate scrubber, 21-1-V-2, where it is scrubbed countercurrently with water over trays.

The char slurry produced in the scrubber is continuously recirculated back to the scrubber by the pumps, 21-1-P-2A or B. A slip stream of the char slurry is withdrawn from the discharge line of 21-1-P-2A and B and combined with the char slurry streams from the parallel trains. Further concentration of the combined char slurry is accomplished in a secondary char recovery unit, 21-1-ME-1. The concentrated char slurry from 21-1-ME-1 is pumped back to the gasifiers, 20-1-R-1 through 21-1-P-3A or B. The effluent water from 21-1-ME-1 flows back to the top section of the particulate scrubbers, 20-1-V-2.

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The clean crude gas from 21-1-V-2 is cooled to approximately 187°F in an air cooler, 21-1-E-4. The condensate produced in gas cooling is separated in the knockout drum, 21-1-V-3, and flows to the bottom section of the ammonia absorber, 21-1-V-4. The overhead gas from 21-1-V-3 is further cooled to approximately 109°F by exchanging heat against a demineralized water and condensate stream in 21-1-E-5 and flows to an ammonia absorber, 21-1-V-4.

The ammonia absorber is a trayed column. Final removal of ammonia from the crude gas is obtained by scrubbing the crude gas countercurrently with cold (70°F) raw water over trays. A portion of the ammonia free crude gas from the top of the ammonia absorber flows to the transport gas compressor, 20-1-C-1, located in the gasification unit (Flow Diagram: EXHC-20-1). The balance of the crude gas from 21-1-V-4 flows to the downstream acid gas removal unit (Flow Diagram: EXHC-22-1). The scrubber bottoms flow to the process condensate unit for further processing.

Equipment Notes

The shell and tube waste heat boiler (21-1-E-1), while not an "off the shelf" item, is commercially feasible. Successful design for such units has been developed by two West German firms, Steinmuller and Siegerner. Another West German firm, BORSIG, also offers their patented design for this service. Several of these units have seen extended commercial service in similar services and appear applicable in coal gasification plants.

All other equipment in this section is commercially available.

ACID GAS REMOVAL

Process Flow Diagrams EAHC-22-1 and EXHC-22-1 depict the acid gas removal system for Cases EAHC and EXHC, respectively.

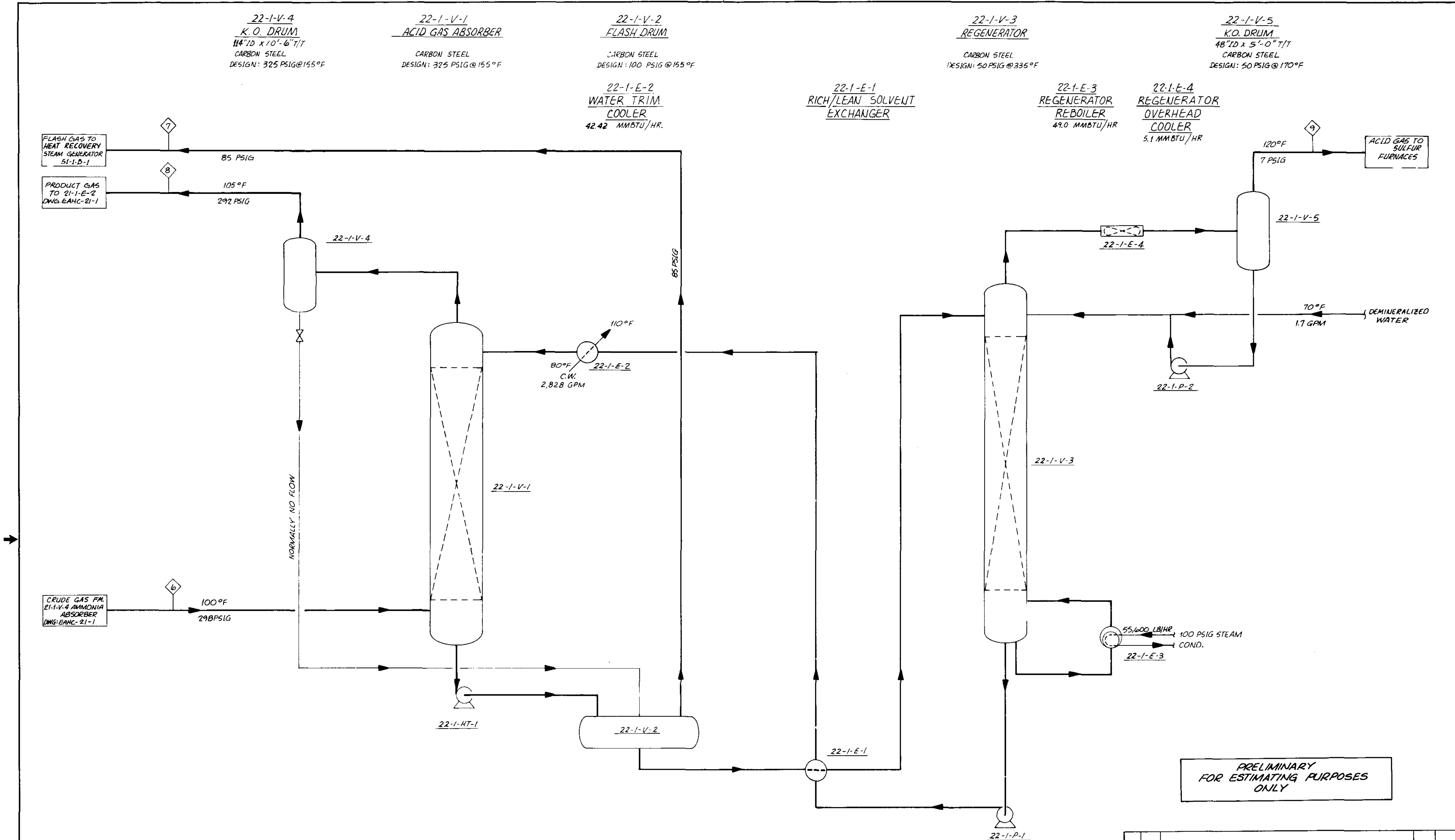
This system employs Allied Chemical Corporation's Selexol® process for selective removal of hydrogen sulfide. Hydrogen sulfide in the crude gas is absorbed in Selexol® solvent in order to reduce sulfur in the treated gas to 1.0 pound sulfur dioxide (SO_2) equivalent per million Btu (HHV) coal charged to the plant.

The process scheme for this section for both case EAHC and EXHC are similar. There are four parallel trains in Case EAHC and three parallel trains in Case EXHC. No spare trains are provided for either case.

The cooled ammonia-free gas flows through an acid gas absorber where it contacts Selexol® solvent countercurrently over a packed bed. The treated gas from the top of the absorber flows through a knockout drum to recover entrained solvent. The cleaned fuel gas is heated in the upstream units (Unit 21) and then flows to the gas turbines located in Unit 50.

The rich solvent from the bottom of the absorber is let down through a hydraulic turbine, 22-1-HT-1, which supplies a portion of the power required by the lean solution pump, 22-1-P-1. It then flows to a flash drum, 22-1-V-2, where most of the dissolved hydrocarbon gases and some of the CO_2 flash off. Most of the dissolved H_2S and COS are retained in the solvent because of their selective absorption in the Selexol® solvent. The flash gases from 22-1-V-2 flow to the heat recovery steam generators (HRSGs) in both the Case EAHC and EXHC. Part of the flash gas in Case EXHC is also supplied to the tail gas unit (Unit 23).

The rich solvent solution from the flash drum exchanges heat with hot regenerated solution in 22-1-E-1 and flows on to the top of the regenerator, 22-1-V-3. In the regenerator the absorbed H_2S and CO_2 are stripped from the solution. Reboil heat is supplied by 100 psig steam in a thermosyphon reboiler, 22-1-E-3. Hot regenerated solvent is pumped back to the absorber, 22-1-V-1 through exchangers 22-1-E-2. Heat is first exchanged with rich solution in 22-1-E-1 in order to reduce reboiler duty. Then the lean solution is cooled down to operating temperature with cooling water in the exchanger, 22-1-E-2.



MATERIAL BALANCE

STREAM Nº	6 CRUDE GAS		7 FLASH GAS		8 PRODUCT GAS		9 ACID GAS	
COMPONENT	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	1,097.7	3.44	11.8	9.41	1,078.1	3.43	7.8	1.95
H_2	4,453.0	13.98	10.0	7.97	4,442.1	14.15	0.9	0.23
CO	9,359.8	29.37	37.4	29.82	9,310.2	29.66	12.2	3.05
CO_2	987.4	3.10	15.0	11.96	840.1	2.68	132.3	33.09
H_2S	218.5	0.69	4.1	3.27	16.4	0.05	198.0	49.52
COS	21.9	0.07	0.3	0.24	14.8	0.05	6.8	1.95
N_2	15,629.7	49.05	45.4	36.21	15,575.0	49.62	9.2	2.30
H_2O	96.9	0.30	1.4	1.12	113.0	0.36	31.5	7.91
TOTAL MPH	31,864.7	100.00	125.4	100.00	31,389.7	100.00	398.7	100.00
TOTAL LB/HR	780,650		3,370		763,700		14,270	
MOL. WT.	24.50		26.89		24.33		35.79	

NOTES:

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY C. VALDEZ
CHECKED BY R. RICE
SUPERVISOR J. BARGER
RELEASE DATE
SUPERVISING ENGR T.H. TAN
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
ENTRAINED BED-AIR BLOWN-COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF

SCALE NONE
DRAWING NUMBER EAHC-22-1
REVISION 02

REDUCED PRINT SCALES (0.5" x 3.0")

AA-8384-EAHC-SS-1

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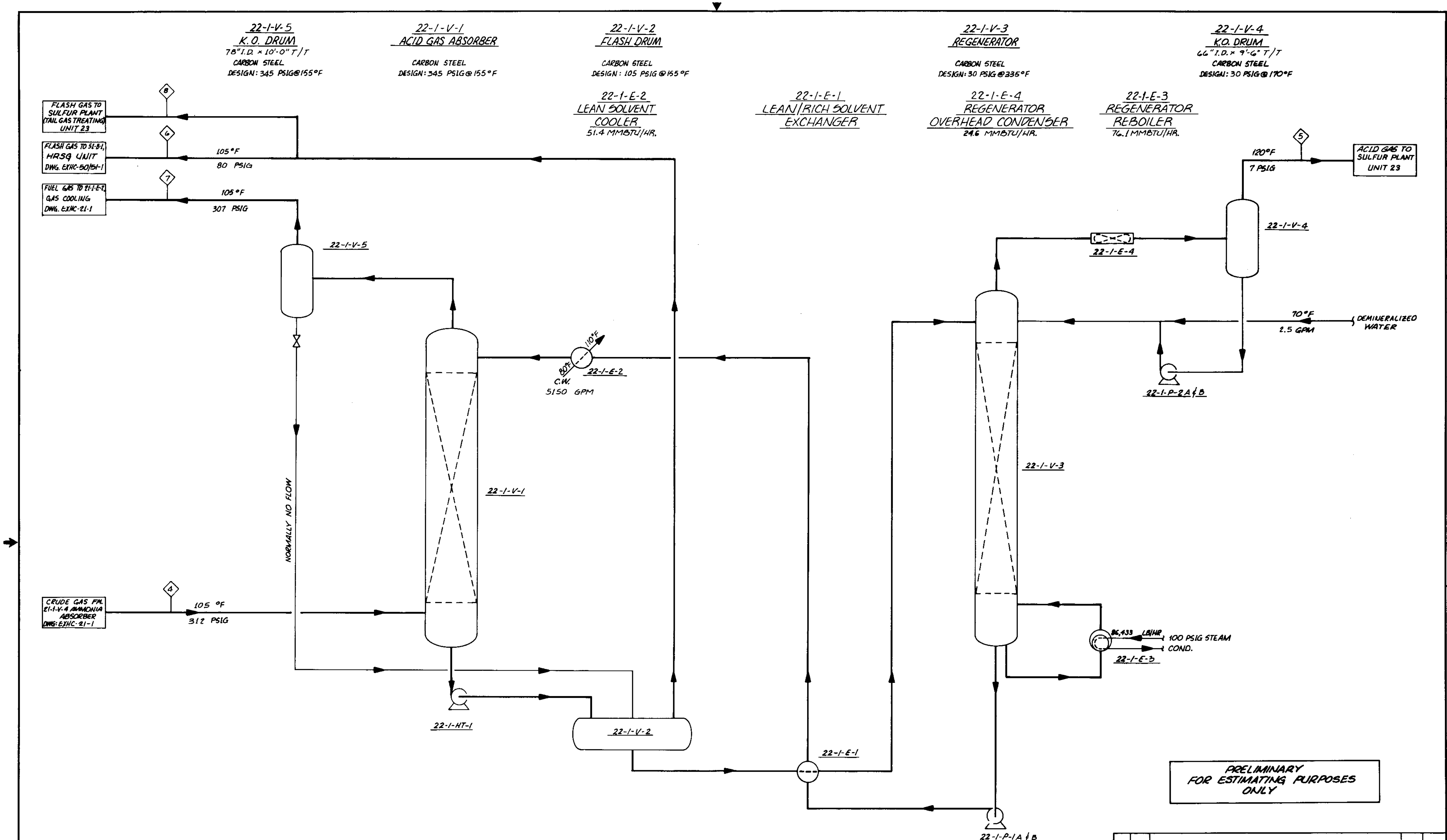

MATERIAL BALANCE

STREAM Nº	4 CRUDE GAS		5 ACID GAS		6 FLASH GAS		7 FUEL GAS		8 FLASH GAS	
COMPONENT	MPH	MOL%	MPH	MOL%	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	1,710.0	7.08	11.3	1.16	13.8	9.84	1,679.8	7.29	5.1	9.84
H_2	8,356.9	34.60	1.8	0.18	17.3	12.28	8,331.5	36.14	6.3	12.28
CO	10,006.3	41.43	12.7	1.30	42.4	30.15	9,935.7	43.09	15.5	30.15
CO_2	3,543.5	14.67	598.1	61.17	59.7	42.46	2,863.8	12.42	21.9	42.46
H_2S	291.3	1.21	267.1	27.32	5.1	3.60	17.3	0.08	1.8	3.60
COS	29.1	.12	11.5	1.18	.5	0.36	16.9	0.07	.2	0.36
N_2	127.2	0.53	—	—	.2	0.17	126.9	0.55	.1	0.17
H_2O	85.8	0.36	75.2	7.69	1.6	1.14	83.5	0.36	.6	1.14
NH_3	—	—	—	—	—	—	—	—	—	—
O_2	—	—	—	—	—	—	—	—	—	—
TOTAL MPH	24,150.1	100.00	977.7	100.00	140.6	100.00	23,055.4	100.00	51.5	100.00
TOTAL LB/HR	497,307		38,013		4,311		454,757		1,579	
MOL. WT.	20.59		38.88		30.66		19.72		30.66	

NOTES;

1. THERE ARE THREE PARALLEL ACID GAS REMOVAL TRAINS.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES, AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

DATE	REVISION DESCRIPTION	DRAWN CHECK	APP

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

FLUOR

DRAWN BY C. VALDEZ
CHECKED BY R. RICE
SUPERVISOR J. SMITH
SUPERVISING ENGR. BENNETT

PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
ENTRAINED BED - OXYGEN BLOWN
HIGH PRESSURE - COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CA.

SCALE NONE | DRAWING NUMBER EXHC-22-1 | REVISION 1

Acid gas from the regenerator overhead is cooled to 120°F in the air fan cooler, 22-1-E-4. The condensate produced in cooling is separated in the knockout drum, 22-1-V-5, and then pumped back to the regenerator through 22-1-P-2. A small stream of demineralized water is added to the condensate at the discharge of 22-1-P-2 to maintain the water balance in the absorption system. The cooled acid gas from 22-1-V-5 flows to the sulfur recovery unit for further processing.

Equipment Notes

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

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SULFUR RECOVERY AND TAIL GAS TREATING

The processes used in these units for both Case EAHC and EXHC are the same as for Case MACW. Refer to Case MACW and the Process Flow Diagrams MACW-23-1, MACW-23-2 and MACW-23-3 for the detailed process descriptions of these units.

There are two 50 percent parallel operating sulfur recovery trains each followed by a tail gas treating unit. Sulfur recovery per train is 149.7 short tons per day for Case EAHC and 160.8 short tons per day for Case EXHC. There is a third (spare) train because of the important environmental requirements these units fulfill.

PROCESS CONDENSATE TREATING

The process condensate treating unit employs U.S. Steel's proprietary PHOSAM-W process for both Case EAHC and EXHC.

The PHOSAM-W is a widely accepted process to reclaim anhydrous ammonia from the sour process condensate. About 128 short tons per day of ammonia is recovered in both the Cases.

The flash gas stream, containing H_2S and COS, from the process condensate unit is sent to tail gas unit (Unit 23) for further processing.

STEAM, BOILER FEEDWATER AND CONDENSATE

Process Flow Diagrams EAHC-30-1 and EXHC-30-1 schematically represent steam, boiler feedwater and condensate systems.

Case EAHC

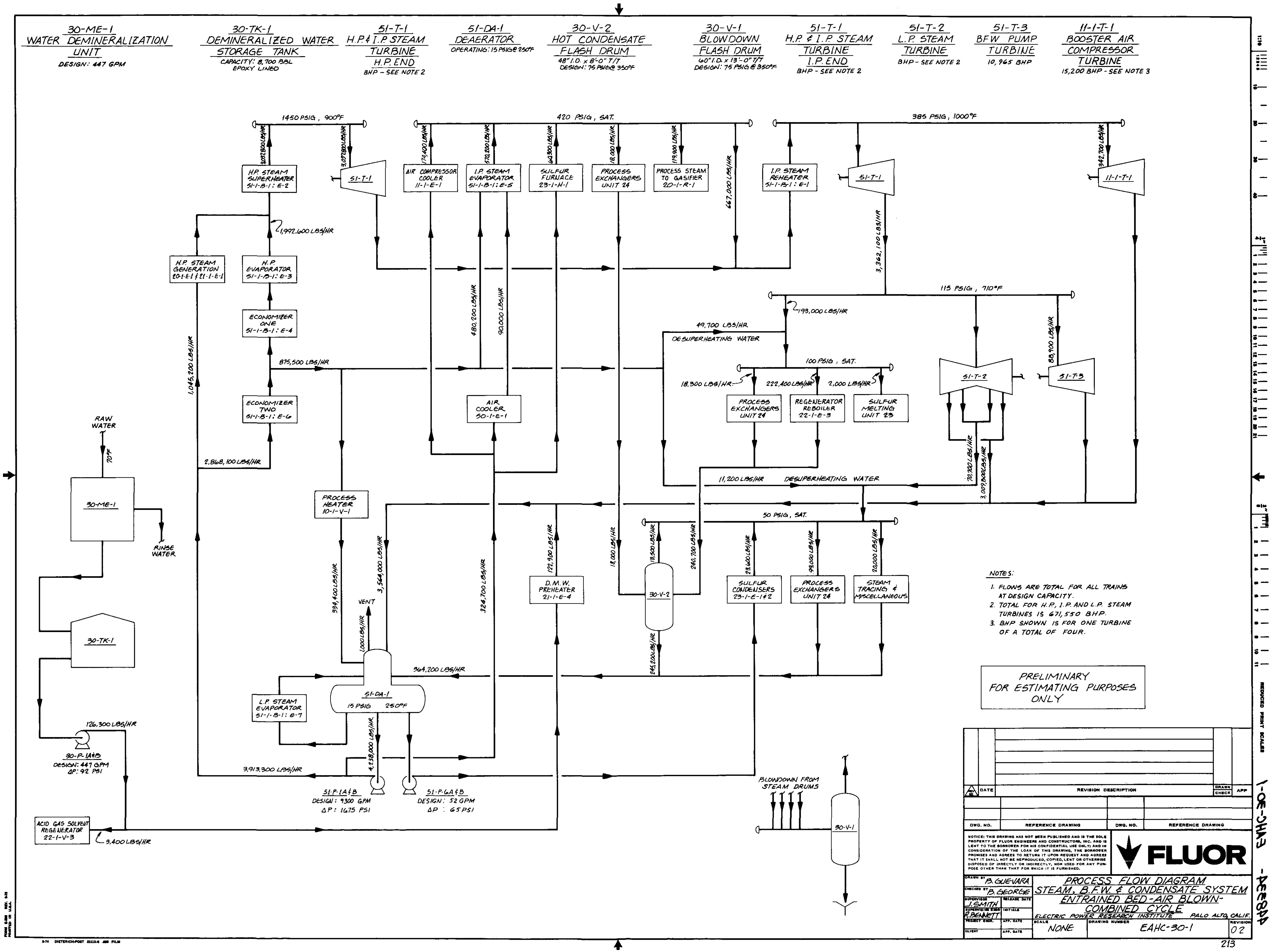
The process plant steam generation is integrated with the combined cycle system. The steam system operates at four levels:

High Pressure	-	1450 psig, 900°F
Intermediate Pressure	-	400 psig
Low Pressure	-	100 psig
Low Pressure	-	50 psig

Major high pressure steam (HP) generation is carried out in heat recovery steam generators (HRSGs) 51-B-1, of gas turbines, 50-1-GT-1. There are eight gas turbines and each gas turbine has its own HRSG. The HRSG is described in detail in Appendix A. Additional HP steam generation is obtained in the process coolers, 20-1-E-1 and 21-1-E-1. The saturated HP steam from the process units combines with the saturated steam produced in the HP evaporators (51-B-1:E-3) and superheated to 900°F in the HRSG's superheaters, 51-1-B-1:E-2. All the superheated HP steam is used to drive the single back pressure type turbine, 51-T-1.

The HP end of the turbine, 51-T-1, takes steam at 1450 psig, 900°F and exhausts at 400 psig.

Saturated intermediate pressure (IP) steam generation at 420 psig is also obtained in the IP steam generators located in the sulfur plant (23-1-H-1), oxidant feed area (11-1-E-1) and the HRSGs (51-B-1:E-5). Part of the saturated IP steam flows to the gasification area, to meet the process steam demand of the gasifiers, and to the process exchangers located in Unit 24. The balance of the saturated IP steam together with the exhaust steam from 51-T-1 is superheated to 1000°F in the HRSG's reheaters (51-1-B1:E-1). The superheated IP steam at 385 psig, 1000°F is then used in the IP end of 51-T-1 and the condensing turbine, 11-1-T-1. The IP end of 51-T-1 exhausts steam to the 100 psig steam header.


30-ME-1
WATER DEMINERALIZATION UNIT
DESIGN: 447 GPM
30-TK-1
DEMINERALIZED WATER STORAGE TANK
CAPACITY: 8,700 BBL
EPOXY LINED
51-T-1
H.P. & I.P. STEAM TURBINE
H.P. END
BHP - SEE NOTE 2
51-DA-1
DEAERATOR
OPERATING: 15 PSIG @ 250°F
30-V-2
HOT CONDENSATE FLASH DRUM
48" I.D. x 8'-0" T/T
DESIGN: 75 PSIG @ 350°F
30-V-1
BLOWDOWN FLASH DRUM
60" I.D. x 13'-0" T/T
DESIGN: 75 PSIG @ 350°F
51-T-1
H.P. & I.P. STEAM TURBINE
I.P. END
BHP - SEE NOTE 2
51-T-2
L.P. STEAM TURBINE
BHP - SEE NOTE 2
51-T-3
BFW PUMP TURBINE
10,965 BHP
11-1-T-1
BOOSTER AIR COMPRESSOR TURBINE
15,200 BHP - SEE NOTE 3
1450 PSIG, 900°F
420 PSIG, SAT.
385 PSIG, 1000°F
115 PSIG, 710°F
100 PSIG, SAT.
50 PSIG, SAT.
H.P. STEAM SUPERHEATER 51-1-B-1: E-2
AIR COMPRESSOR COOLER 11-1-E-1
I.P. STEAM EVAPORATOR 51-1-B-1: E-5
SULFUR FURNACE 23-1-H-1
PROCESS EXCHANGERS UNIT 24
PROCESS STEAM TO GASIFIER 20-1-R-1
I.P. STEAM REHEATER 51-1-B-1: E-1
H.P. STEAM GENERATION 20-1-E-1 / 21-1-E-1
H.P. EVAPORATOR 51-1-B-1: E-3
ECONOMIZER ONE 51-1-B-1: E-4
ECONOMIZER TWO 51-1-B-1: E-6
AIR COOLER 50-1-E-1
PROCESS EXCHANGERS UNIT 24
REGENERATOR REBOILER 22-1-E-3
SULFUR MELTING UNIT 23
PROCESS HEATER 10-1-V-1
D.M.W. PREHEATER 21-1-E-4
SULFUR CONDENSERS 23-1-E-1 & 2
PROCESS EXCHANGERS UNIT 24
STEAM TRACING & MISCELLANEOUS
L.P. STEAM EVAPORATOR 51-1-B-1: E-7
51-DA-1
15 PSIG 250°F
VENT
RAW WATER
70°F
RINSE WATER
30-P-1A & B
DESIGN: 447 GPM
ΔP: 92 PSI
51-P-1A & B
DESIGN: 9300 GPM
ΔP: 1675 PSI
51-P-6A & B
DESIGN: 52 GPM
ΔP: 65 PSI
ACID GAS SOLVENT REGENERATOR 22-1-V-3
BLOWDOWN FROM STEAM DRUMS
DESUPERHEATING WATER
1,992,600 LBS/HR
875,500 LBS/HR
2,868,100 LBS/HR
1,045,200 LBS/HR
3,913,300 LBS/HR
126,300 LBS/HR
3,400 LBS/HR
4,238,000 LBS/HR
364,200 LBS/HR
1,000 LBS/HR
334,400 LBS/HR
3,564,000 LBS/HR
324,700 LBS/HR
480,200 LBS/HR
90,000 LBS/HR
122,900 LBS/HR
18,000 LBS/HR
13,500 LBS/HR
240,700 LBS/HR
245,200 LBS/HR
667,000 LBS/HR
3,362,100 LBS/HR
342,700 LBS/HR
193,000 LBS/HR
49,700 LBS/HR
18,300 LBS/HR
222,400 LBS/HR
2,000 LBS/HR
88,900 LBS/HR
11,200 LBS/HR
70,700 LBS/HR
3,009,800 LBS/HR
23,600 LBS/HR
99,000 LBS/HR
20,000 LBS/HR
NOTES:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. TOTAL FOR H.P., I.P. AND L.P. STEAM TURBINES IS 671,550 BHP.
3. BHP SHOWN IS FOR ONE TURBINE OF A TOTAL OF FOUR.
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
REVISION DESCRIPTION
REFERENCE DRAWING
FLUOR
DRAWN BY B. GUEVARA
CHECKED BY B. GEORGE
J. SMITH
R. BENNETT
PROCESS FLOW DIAGRAM
STEAM, B.F.W. & CONDENSATE SYSTEM
ENTRAINED BED-AIR BLOWN-COMBINED CYCLE
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER EAHC-30-1
REVISION 02
213
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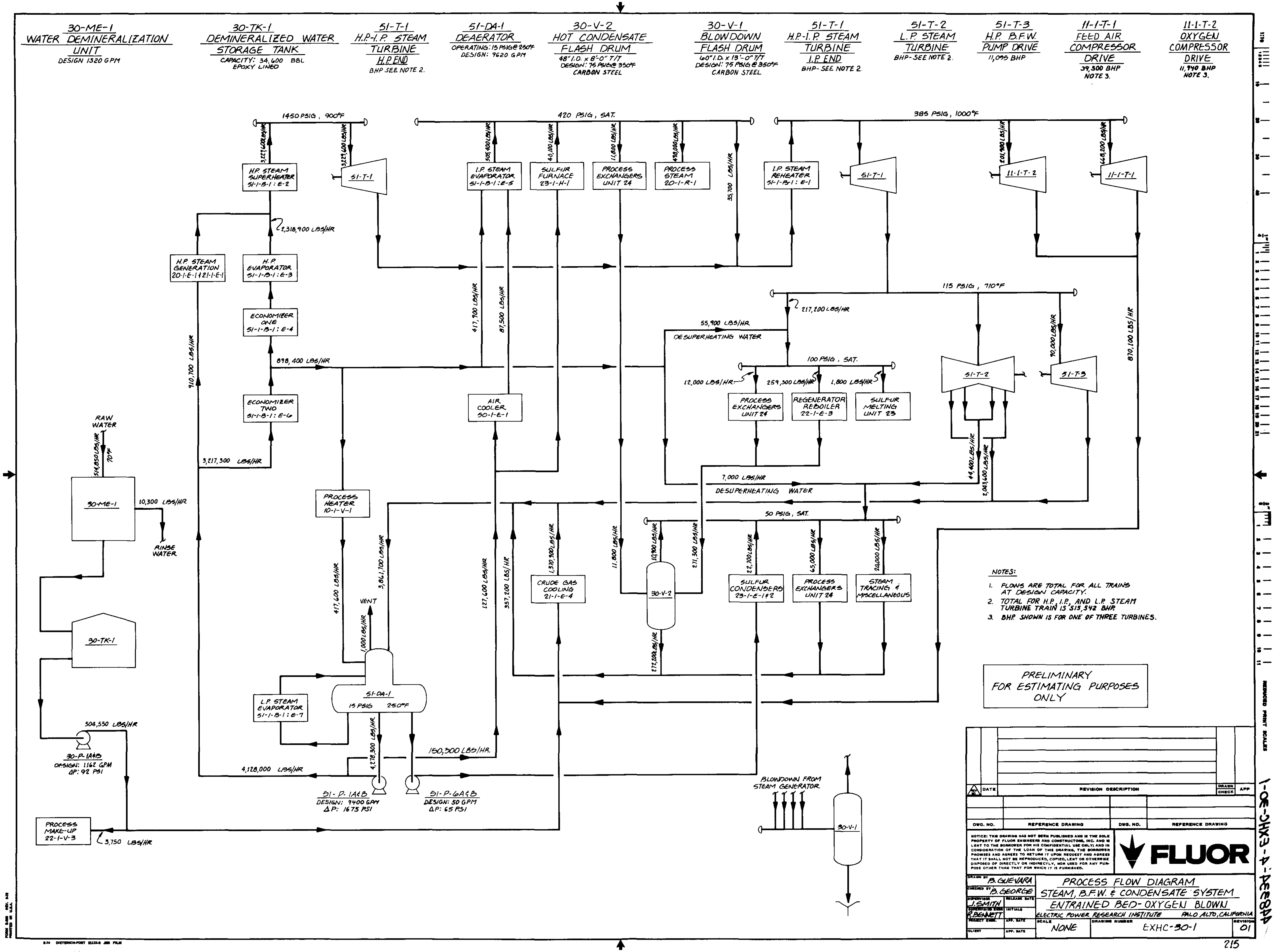
EAHC-30-1



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A portion of the 100 psig steam is desuperheated and supplied to the process exchangers in the acid gas removal and process condensate units. A small quantity of the desuperheated IP steam is also used for sulfur melting in Unit 23. The balance of the 100 psig exhaust steam is used in the LP turbines, 51-T-2 and 51-T-3. The LP turbine, 51-T-2, is a condensing turbine exhausting at 2-1/2" Hg abs. The turbine is provided with the capability of steam extraction at 50 psig. The turbine, 51-T-3, is a condensing turbine exhausting at 2-1/2" Hg abs.

The 50 psig steam header is supplied by steam generation in the process exchangers, 23-1-E-1 and E-2, and extraction from Turbine 51-T-2. Hot condensate from the 100 psig steam users in Unit 24 is flashed in 30-V-2 at 50 psig and the flashed steam flows to the 50 psig header. The 50 psig steam is mainly used for the process exchangers in Unit 24, and steam tracing and other miscellaneous users.

Raw water is treated in a semiautomatic resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for 1500 psig boiler. Demineralized water is stored in a tank, 30-TK-1. Storage equivalent to 24 hours of demineralized water production is provided. The demineralized water requirement is estimated at approximately 253 gpm. Demineralized water from the storage tank is transported to the deaerator through Pumps 30-1-P-1A and B. A small quantity of the makeup water is withdrawn from the discharge of 30-1-P-1A and B and transported to Unit 22. The balance of the demineralized water after recovering heat in 21-1-E-4 combines with the vacuum condensate from the turbines, 11-1-T-1, 51-T-2 and 51-T-3. The combined stream of demineralized water and vacuum condensate then flows to the deaerator, 51-DA-1. The condensate from the 50 psig steam users, 30-V-2, and the cold BFW from the process heater, 10-1-V-1, also flows to the deaerator.

The deaerator is a tray type unit operating at 28 psia. The deaerator provides for 10 minute storage. The deaerating steam is generated in the HRSG's LP steam evaporators 51-1-B-1:E-7.

Boiler feedwater (BFW) from the deaerator is pumped through high pressure boiler feedwater pumps (51-P-1A and B) to the HRSGs, process high pressure steam generators (20-1-E01 and 21-1-E-1) and the IP steam generators, 11-1-E-1, 23-1-H-1 and 51-1-B-1:E-5.

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BFW to the HRSGs is first heated to the IP steam saturation temperature (459°F) in Economizers 51-1-B-1:E-6. Part of the BFW is withdrawn downstream of 51-1-B-1:E-6 and supplied to the process heaters, 10-1-V-1, HRSG's IP evaporators 51-1-B-1:E-5 and the desuperheating stations for the 50 psig and 100 psig steam headers. The balance of the BFW is heated to the HP steam saturation temperature (598°F) in the HRSG's economizers, 51-1-B-1:E-4 and flows to the HP evaporators, 51-1-B-1:E-3, where saturated HP steam is generated.

The 50 psig steam generators are supplied boiler feedwater by a separate set of pumps, 51-P-6A and B.

Case EXHC

The steam boiler feedwater system for Case EXHC is similar to Case EAHC with the exception that additional IP steam turbines, 11-1-T-2, are provided for driving the oxygen compressors. Turbines 11-1-T-2 are condensing type with steam intake at 385 psig, 1000°F and steam exhaust at 2-1/2" Hg abs.

There are eight gas turbines, 50-1-GT-1, and each gas turbine has its own HRSG. There is a single steam turbine generator unit driven by the HP and IP turbine, 51-T-1 and LP turbine, 51-T-2.

COMBINED CYCLE SYSTEM

Process Flow Diagrams EAHC-50/51-1 and EXHC-50/51-1 schematically represent the combined cycle system for Case EAHC and EXHC, respectively. These diagrams depict the total power block flows.

In both cases there are eight parallel trains of gas turbines, 50-1-GT-1, generators, 50-1-G-1, and heat recovery steam generators (HRSG), 51-1-B-1 and one 100 percent steam turbine, 51-T-1&2, and generator unit (51-G-1). Refer to Appendix A for a detailed description of the combined cycle system. Detailed performance information of the power block components, i.e., gas turbines, HRSGs and the steam turbine is also summarized in Appendix A.

Approximately 34 percent of the total high pressure (HP) steam generated in the plant is supplied by process steam generators in Case EAHC. About 28 percent of the total HP steam produced in the plant is supplied by the process steam generators in Case EXHC. In both cases process HP steam is combined with the HP steam produced in the HRSG and superheated in 51-1-B-1:E-2. The superheated HP steam then flows to the HP end of the turbine, 51-T-1.

Steam for the boiler feedwater pump drives is taken from steam crossover from IP to LP steam turbine cylinders in both cases. Steam for the air compressor drive in Case EAHC and the air and oxygen compressors' drives in Case EXHC is taken from the hot reheat line.

Hot feedwater at 459°F from the outlet of economizer two (51-1-B-1:E-6) in the HRSG is supplied to the process plants to meet the process heating requirements in both cases.

In both cases, where possible, process cooling loads are integrated into the condensate and makeup system. Approximately 4.8 MM Btu/hr and 78.7 MM Btu/hr of low level process cooling loads are recovered respectively in Cases EAHC and EXHC.

Equipment Notes

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

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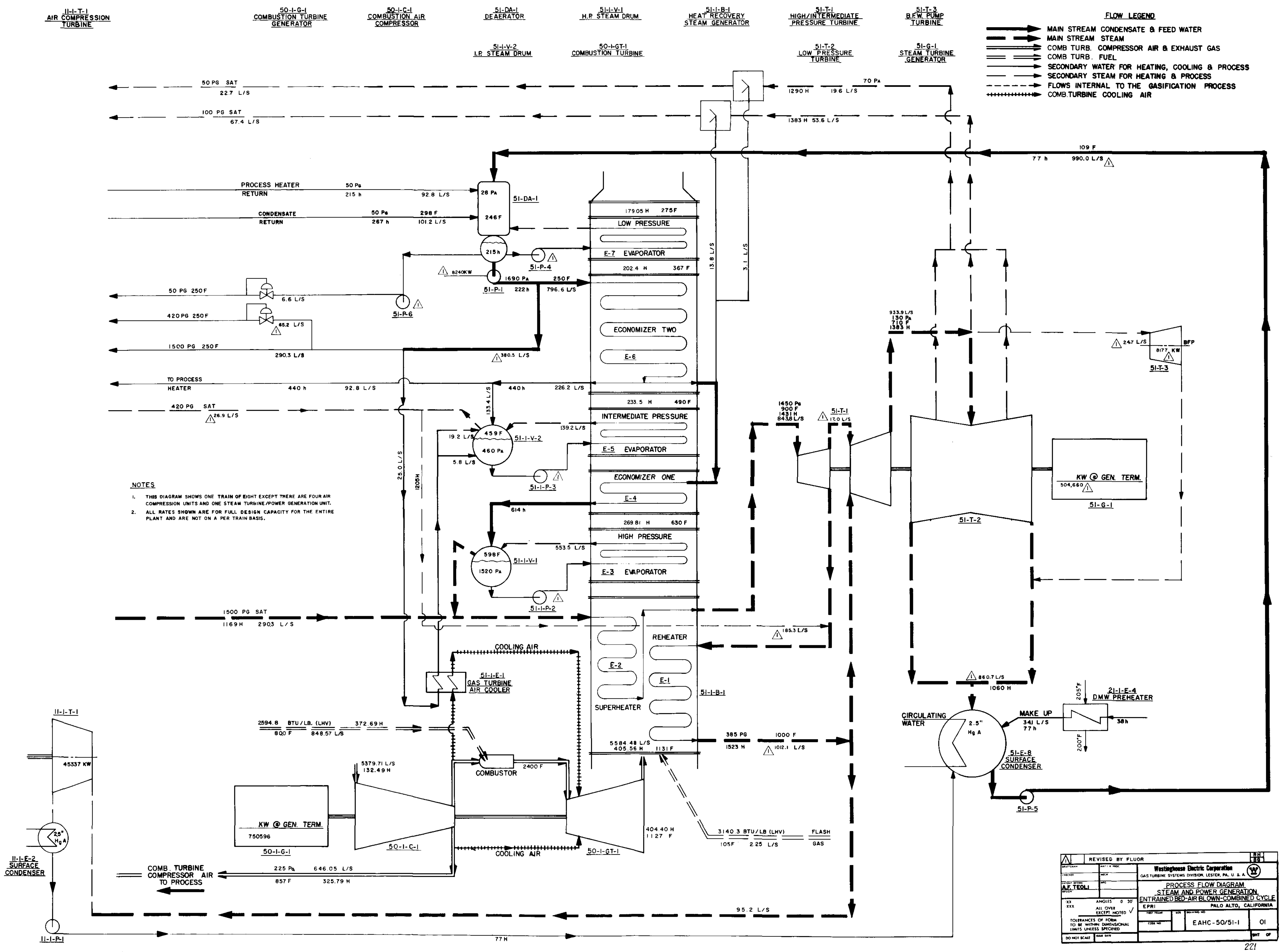

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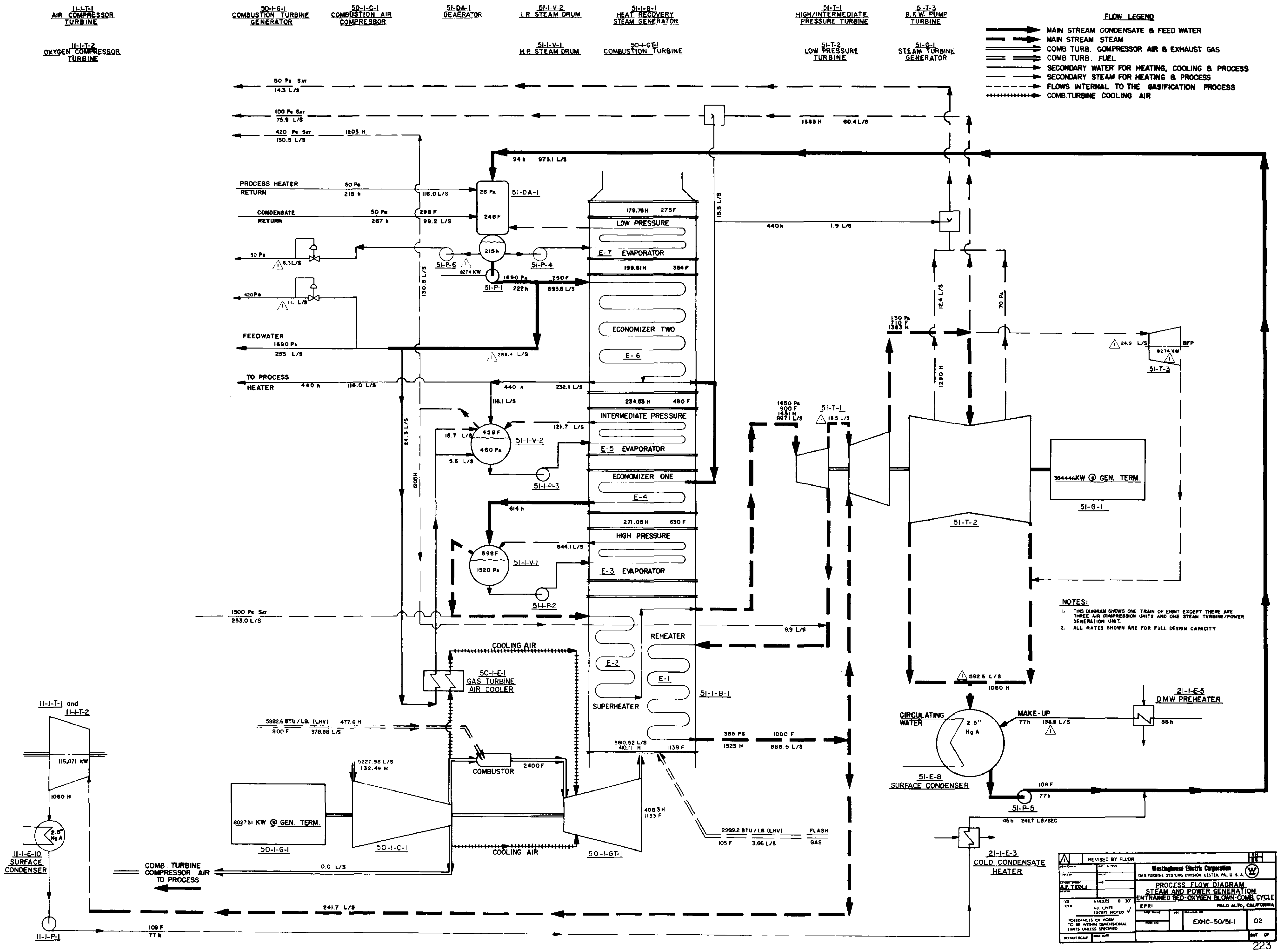

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PROCESS DISCUSSION

The table below summarizes pertinent heat and material balance results.

TABLE EH-2

SUMMARY OF SYSTEM PERFORMANCE - CASES EAHC AND EXHC

	Case EAHC	Case EXHC
GASIFICATION AND GAS CLEANING SYSTEM		
Coal Feed Rate, lbs/hr (m.f.)	798,333	798,333
Oxygen or Air (1)/Coal Ratio, lbs/lb m.f.	2.857	0.609
Oxidant Temperature, °F	800	335
Steam/Coal Ratio, lbs/lb m.f.	0.150	0.624
Gasifier Exist Pressure, psig	360	360
Crude Gas Temperature, °F	1,700	1,700
Crude Gas HHV (dry basis), Btu/SCF (2)	174.1	315.4
Temperature of Fuel Gas to Gas Turbine, °F	800	800
POWER SYSTEM		
Gas Turbine Inlet Temperature, °F	2,400	2,400
Pressure Ratio	17:1	17:1
Turbine Exhaust Temperature, °F	1,127	1,133
Steam Conditions, psig/°F/°F	1,450/900/1,000	1,450/900/1,000
Condenser Pressure, Inches Hg abs	2.5	2.5
Stack Temperature, °F	275	275
Gas Turbine Power MW, (3)	751	803
Steam Turbine Power MW, (3)	504	384
Power Consumed, MW	42	38
Net System Power, MW	1,213	1,149
OVERALL SYSTEM		
Process and Deaerator Makeup Water, gpm/1000 MW	497	1,031
Cooling Tower Makeup Water, gpm/1000 MW	6,125	6,003
Cooling Water Circulation Rate, gpm/MW	341	321
Cooling Tower Heat Rejection, % of Coal HHV	36.8	33.2
Air Cooler Heat Rejection, % of Coal HHV	3.2	7.2
Net Heat Rate, Btu/kWh	8,428	8,876
Overall System Efficiency (Coal → Power), % of Coal HHV	40.5	38.5

(1) Dry Basis, 100% O_2 for oxygen blown
(2) Excluding the HHV of H_2S, COS and NH_3
(3) At Generator Terminals

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Gasifier Material Balances

Tables EH-3A and EH-3X present gasifier material balances for the air and oxygen cases. The gasification units presented in this section represent those proposed by the Foster Wheeler Corporation and operate at relatively high pressure, around 400 psig. They are two-stage entrained flow devices, similar in many respects to the BIGAS system. Due to the two-stage nature of this device as well as the operating pressure, significant quantities of methane should be produced.

Significant quantities of ammonia are produced and recovered as byproduct, using the PHOSAM-W process.

As in Case EXTC, little or no hydrocarbons heavier than methane are expected.

Acid Gas Removal

Selexol® process is used in both EAHC and EXHC for acid gas removal. Case EXHC has three parallel trains and Case EAHC, four parallel acid gas removal trains. Case EXHC has a much higher acid gas loading, 11,503 mph (CO_2 + H_2S), compared to 4823 mph for EAHC, requiring higher solvent circulation. The total gas flow to the acid gas absorber is higher for Case EAHC because of the nitrogen in the air blown case. Because of this, the number of parallel trains increase from three for EXHC to four for EAHC.

Process Energy Balances

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

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

TABLE EH-3A

MATERIAL BALANCE - CASE EAHC

FEEDS

	T (°F)	lb/hr	lb mol/hr
Coal	200		
Moisture		16,293	904.4
Ash		79,833	
MAF Coal			
Carbon		555,113	46,216.7
Hydrogen		42,535	21,099.4
Oxygen		80,041	2,501.4
Nitrogen		9,987	356.5
Sulfur		30,824	961.3
TOTAL COAL		814,626	
Steam	453	119,896	6,655.0
Oxidant	800		
Oxygen		531,164	16,599.9
Nitrogen		1,749,376	62,447.3
Moisture		36,066	2,001.9
TOTAL OXIDANT		2,316,606	81,049.1
Transport Gas	228		
CH_4		1,649	103.4
H_2		846	419.7
CO		24,709	882.1
CO_2		4,093	93.0
H_2O		191	10.6
H_2S		702	20.6
COS		126	2.1
N_2		41,272	1,473.3
TOTAL TRANSPORT GAS		73,598	3,004.8
Char Slurry	180		
Char		3,428	
Water		5,142	
TOTAL CHAR SLURRY		8,570	
TOTAL FEEDS		3,333,296	

EFFLUENTS

	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	1,700			
CH_4		72,097	4,494.0	3.37
H_2		36,755	18,231.8	13.68
CO		1,073,432	38,321.3	28.75
CO_2		177,915	4,042.6	3.03
H_2O		47,453	2,634.1	1.98
H_2S		30,485	894.5	0.67
COS		5,377	89.5	0.07
N_2		1,792,605	63,991.9	48.02
NH_3		9,714	570.4	.43
TOTAL GASIFIER EFFLUENT		3,245,833	133,270.1	100.00
Ash	2,700	84,035		
Char	1,700	3,428		
TOTAL EFFLUENTS		3,333,296		

TABLE EH-3X

MATERIAL BALANCE - CASE EXHC

FEEDS

	T (°F)	lb/hr	lb mol/hr
Coal	200		
Moisture		16,293	904.4
Ash		79,833	
MAF Coal			
Carbon		555,113	46,216.4
Hydrogen		42,535	21,099.6
Oxygen		80,041	2,501.4
Nitrogen		9,987	356.5
Sulfur		30,824	961.3
TOTAL COAL		814,626	
Steam	453	498,000	
Oxidant	335		
Oxygen		486,458	15,201.9
Nitrogen		8,693	310.2
Moisture			
TOTAL OXIDANT		495,151	15,512.1
Transport Gas	320		
CH_4		4,059	252.9
H_2		2,492	1,236.0
CO		41,448	1,479.6
CO_2		23,049	523.8
H_2S		1,472	43.2
COS		252	4.2
N_2		530	18.9
H_2O		227	12.6
TOTAL TRANSPORT GAS		73,529	3,571.2
Char	290		
Char		2,911	
Water		4,326	
TOTAL CHAR		7,237	
TOTAL FEEDS		1,888,543	

EFFLUENTS

	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	1,700			
CH_4		86,357	5,383.2	6.04
H_2		53,034	26,306.7	29.50
CO		822,273	31,498.5	35.33
CO_2		490,901	11,154.3	12.51
H_2O		231,352	12,841.5	14.40
H_2S		31,251	917.1	1.03
COS		5,496	91.5	0.10
N_2		11,220	400.5	0.45
NH_3		9,713	570.3	0.64
TOTAL GASIFIER EFFLUENT		1,801,597	89,163.6	100.00
Ash	2,700	84,035		
Char	1,700	2,911		
TOTAL EFFLUENTS		1,888,543		

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The tables show that the air blown case results in somewhat more of the coal energy charged to the plant being converted to power than does the oxygen blown case. Coal charged at 10,000 ton/day is equivalent to 10,196 MM Btu/hr HHV. Case EAHC produces 4129 MM Btu/hr power. Case EXHC produces 3921 MM Btu/hr.

With all the power consumed in the process unit included, the system cold efficiency, net power at 3413 Btu/kWh as a percent of the coal HHV, is 40.5 percent for Case EAHC and 38.5 for Case EXHC. The heat rate based on net power produced is 8428 Btu/kWh for the air blown case, and 8876 Btu/kWh for the oxygen blown case.

Comparisons drawn from the tables illustrate some of the differences between the air and oxygen blown cases. The air blown case exhibits a significantly higher overall thermal efficiency than the oxygen blown system. The oxidant feed unit in the oxygen blown case requires 93,520 BHP more compression power than in the air blown case. For both cases, this power is supplied by the expansion of 385 psig, 1000°F steam. As a result, the air blown case shows higher power generation from the steam cycle when compared to the oxygen blown case, 501 MW versus 384 MW. The oxidant compression in the oxygen blown case requires inter-stage cooling to keep within acceptable operating temperature limitations. This results in a significant heat loss which cannot be recovered in a manner that increases the plant's energy output. This loss is 2.99 percent for the oxygen blown case and zero for the air blown case.

There are major differences in the acid gas removal unit design between the air and oxygen cases although the resulting heat rejections are not far different. Heat rejection is 39 MM Btu/hr (0.38 percent of coal HHV) higher in the oxygen case. Even though there is a dilution effect of nitrogen in the gasifier effluent for the air blown case, this is offset by the higher CO_2 content of the crude gas for the oxygen blown case.

The oxygen blown case gives higher power generation from the gas turbine, 803 MW compared to 751 MW for the air blown case. This also results in greater stack gas losses from the HRSG. The oxygen blown case has 231 MM Btu/hr (2.28 percent of coal HHV) higher HRSG stack losses than the air blown case.

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TABLE EH-4A

ENERGY BALANCE - CASE EAHC

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

	MM Btu/hr					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Turbine Compressor Suction Air		117	281			398
Demineralized Water		1				1
Auxiliary Power Inputs					142	142
TOTAL	10,196	123	281	0	142	10,742
HEAT OUT						
Ash/Slag	58	99				157
Gasifier Heat Loss				29		29
Gasifier Cooling		3				3
Sulfur Product	106	1				107
Ammonia Product	100					100
Generated Power					4,271	4,271
Power Block Losses(1)				46	107	153
Oxidant Compressor Surface Condenser			337			337
Power Surface Condenser			3,099			3,099
HRSG Stack Losses		1,062	791			1,853
Gasifier Effluent Cooling		182	42			224
Selexol Solvent Cooling		192				192
Selexol Regeneration Condenser		1	19			20
Transport Gas Vent		14	3			17
Process Condensate Treating		140				140
Steam Heat Losses		18	3			21
Waste Water Effluent		19				19
Pulverizer Vent Gas		48	21			69
TOTAL	264	1,779	4,315	75	4,378	10,811

$$\frac{\text{Output - Input}}{\text{Input}} = 0.64\%$$

(1) Includes mechanical and electrical losses.

TABLE EH-4X

ENERGY BALANCE - CASE EXHC

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

	MM Btu/hr					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Air Compressor Suction Air		16	39			55
Turbine Compressor Suction Air		129	318			447
Transport Air Compressor Suction Air		1	3			4
Demineralized Water		5				5
Auxiliary Power Inputs					131	131
TOTAL	10,196	156	360	0	131	10,843
HEAT OUT						
Ash/Slag	58	101				159
Gasifier Heat Loss				24		24
Gasifier Cooling		34				34
Sulfur Product	106	1				107
Ammonia Product	100					100
Generated Power					4,052	4,052
Power Block Losses(1)				47	133	180
Oxidant Compressor Surface Condensers			855			855
Power Surface Condensers			2,180			2,180
HRSG Stack Gas		1,089	995			2,084
Gasifier Effluent Cooling		63	205			268
Oxidant Compressors Interstage Cooling		304	1			305
Selexol Solvent Cooling		155				155
Selexol Regeneration Condenser		5	91			96
Process Condensate Treating		50	50			100
Transport Air Vent		14	3			17
Steam Heat Losses		18	3			21
Waste Water Effluent		12				12
Pulverizer Vent Gas		48	21			69
Air Separation Plant Waste Gas		14				14
TOTAL	264	1,909	4,403	71	4,185	10,832

$$\frac{\text{Input - Output}}{\text{Input}} = 0.10\%$$

(1) Includes electrical and mechanical losses.

TABLE EH-5

ENERGY BALANCE AS PERCENT COAL HHV - CASES EAHC AND EXHC

	Case EAHC		Case EXHC	
	MM Btu/hr	Percent	MM Btu/hr	Percent
IN				
Coal HHV	10,196	100.0	10,196	100.0
OUT				
Net Power	4,129	40.50	3,921	38.45
Sulfur Product, HHV	107	1.05	107	1.05
Ammonia Product, HHV	100	0.98	100	0.98
Selexol Sensible and Latent	212	2.08	251	2.46
Oxidant Interstage Cooling	0	0.00	305	2.99
Ash/Slag	157	1.54	159	1.56
HRSG Stack Gases	1,853	18.17	2,084	20.45
Rejected at Condensers	3,436	33.70	3,035	29.77
Other Sensible Losses	301	2.95	97	0.95
Other Latent Losses	(212)	(2.08)	(78)	(0.77)
Gasifier Heat Losses	29	0.28	24	0.24
Power Block Losses	153	1.50	180	1.77
	10,265	100.67	10,185	99.90

ECONOMICS

Important economic results are summarized below.

TABLE EH-6

SUMMARY OF ECONOMICS - CASE EAHC & EXHC

	Case EAHC	Case EXHC
PRODUCTION AT DESIGN CAPACITY		
Net Power, MW (1)	1,213.6	1,148.7
Overall Plant Heat Rate, Btu/kWh	8,428	8,876
TOTAL CAPITAL (2)		
Total Capital @ $1/MM Btu Coal, $1,000	853,119	848,392
Total Capital @ $1/MM Btu, $/kW	705	739
Total Capital @ $2/MM Btu Coal, $1,000	870,237	865,510
Total Capital @ $2/MM Btu, $/kW	719	753
AVERAGE COSTS OF SERVICES (2)		
Annual Cost @ $1/MM Btu Coal, $1000/yr	241,328	239,851
Per Unit @ $1/MM Btu Coal, mills/kWh	32.53	34.05
Annual Cost @ $2/MM Btu Coal, $1000/yr	306,520	305,043
Per Unit @ $2/MM Btu Coal, mills/kWh	41.32	43.30

NOTES

(1) At 100% Operating Factor
(2) Mid-1976 Dollars and 70% Operating Factor

Even though the total capital requirements and annual cost of services are slightly higher for the air blown case, when viewed on a net power produced basis, the air blown case is about 5 percent lower in cost than the oxygen blown case.

Tables EH-7 and EH-8 give detailed breakdowns of plant investment, capital charges and working capital for both cases at 70 percent operating factor and $1.00/MM Btu and $2.00/MM Btu coal respectively. The accuracy of plant investment estimates is judged to be ±25%. Since other capital charges and working capital are keyed to elements of plant investment, this accuracy is reflected in other capital figures as well. This should be kept in mind when comparing different gasifier cases. The accuracy of comparisons of the same gasifier types should be somewhat more accurate since the same estimate accuracies will occur in both cases.

Although the total plant investment for the air and oxygen blown cases are nearly the same, there are substantial differences in costs of various units. As would be expected, the cost of the oxidant feed system in the oxygen blown case is much higher. The steam, condensate and BFW unit cost is also higher in the oxygen blown case. Higher process steam requirements for gasification in the oxygen blown case lead to higher boiler feedwater requirements and about 6 percent more high pressure steam production. For the oxygen blown case, the higher cost in the oxidant feed system and steam, condensate, and boiler feedwater unit is partially offset by the lower costs in the remaining process units and in the utilities and offsites area. Higher costs of these facilities in the air blown case are principally due to higher mass throughput resulting from nitrogen dilution.

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

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

Operating labor requirements are functions of the number of units and trains. Requirements on a per shift basis are:

	Case EAHC	Case EXHC
"A" Operators	5	5
"B" Operators	16	17
Foremen	2	2
Lab and Instrument Technicians	4	4

Case EXHC has one more operator per shift because of the air separation unit and the additional compressors in the oxidant feed system.

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

The operating charges are nearly the same for both cases even though there are some differences in the operating charge items.

Operating charges constitute about 45 percent of cost of services with coal at \$1.00/MM Btu and nearly 56 percent at a coal cost of \$2.00/MM Btu. For all cases, coal is the largest single operating charge. The relationship as percentages are summarized below:

	Case EAHC		Case EXHC	
Coal Cost, $1/MM Btu, HHV	1.00	2.00	1.00	2.00
Coal as % of Operating Charges	57.8	73.2	58.1	73.5
Coal as % of Total Cost of Services	25.9	40.8	26.1	41.0
Operating Charges as % of Total Cost of Services	44.9	55.7	44.8	55.7
Capital Charges as % of Total Cost of Services	55.1	44.3	55.2	44.3

TABLE EH-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $1.00/MM BTU COAL

	Case EAHC $1,000[1]	Case EAHC $/kW[2]	Percent	Case EXHC $1,000[1]	Case EXHC $/kW[2]	Percent
PLANT INVESTMENT						
Coal Handling	29,320	24.24	5.41	29,320	25.52	5.39
Oxidant Feed	21,456	17.74	3.96	66,891	58.23	12.30
Gasification and Ash Handling	42,710	35.30	7.88	35,603	30.99	6.55
Gas Cooling	32,155	26.58	5.93	19,388	16.88	3.57
Acid Gas Removal and Sulfur Recovery	36,408	30.10	6.72	27,268	23.74	5.01
Process Condensate Treating	9,830	8.13	1.81	7,648	6.66	1.41
Steam, Condensate and BFW	1,650	1.36	0.30	2,758	2.40	0.51
Support Facilities	61,916	51.13	11.42	57,551	50.10	10.58
Combined Cycle	306,677	253.50	56.57	297,373	258.88	54.68
Subtotal	542,122	448.08	100.00	543,800	473.40	100.00
Contingency	121,769	100.65		116,767	101.65	
Total Plant Investment	663,891	548.73		660,567	575.05	
ILLINOIS SALES TAX	14,845	12.27		14,845	12.92	
CAPITAL CHARGES						
Preproduction Costs	42,157	34.85		41,900	36.47	
Paid-up Royalties	3,319	2.74		3,303	2.88	
Initial Catalyst and Chemical Charge	1,351	1.12		846	0.74	
Construction Loan Interest	82,920	68.54		82,505	71.82	
Total Capital Charges	129,747	107.25		128,554	111.91	
DEPRECIABLE CAPITAL	808,483	668.25		803,966	699.88	
WORKING CAPITAL	44,636	36.90		44,426	38.67	
TOTAL CAPITAL	853,119	705.15		848,392	738.55	

NOTE

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

TABLE EH-8

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $2.00/MM BTU COAL

	Case EAHC			Case EXHC		
	$1,000 (1)	$/kW (2)	Percent	$1,000 (1)	$/kW (2)	Percent
PLANT INVESTMENT						
Coal Handling	29,320	24.24	5.41	29,320	25.52	5.39
Oxidant Feed	21,456	17.74	3.96	66,891	58.23	12.30
Gasification and Ash Handling	42,710	35.30	7.88	35,603	30.99	6.55
Gas Cooling	32,155	26.58	5.93	19,388	16.88	3.57
Acid Gas Removal and Sulfur Recovery	36,408	30.10	6.72	27,268	23.74	5.01
Process Condensate Treating	9,830	8.13	1.81	7,648	6.66	1.41
Steam Condensate & BFW	1,650	1.36	0.30	2,758	2.40	0.51
Support Facilities	61,916	51.18	11.42	57,551	50.10	10.58
Combined Cycle	306,677	253.50	56.57	297,373	258.88	54.68
Subtotal	542,122	448.13	100.00	543,800	473.40	100.00
Contingency	121,769	100.65		116,767	101.65	
Total Plant Investment	663,891	548.78		660,567	575.05	
ILLINOIS SALES TAX	14,845	12.27		14,845	12.92	
CAPITAL CHARGES						
Preproduction Cost	44,017	36.39		43,760	38.09	
Paid-up Royalties	3,319	2.74		3,303	2.88	
Initial Catalyst and Chemical Charge	1,351	1.12		846	0.74	
Construction Loan Interest	82,920	68.54		82,505	71.82	
Total Capital Charges	131,607	108.79		130,414	113.53	
DEPRECIABLE CAPITAL	810,343	669.84		805,825	701.50	
WORKING CAPITAL	59,894	49.51		59,684	51.95	
TOTAL CAPITAL	870,237	719.35		865,510	753.45	

NOTE

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

TABLE EH-9

COST OF SERVICES AT 70% OPERATING LOAD FACTOR

COAL COST, HHV	Case EAHC $1/MM Btu	Case EAHC $2/MM Btu	Case EXHC $1/MM Btu	Case EXHC $2/MM Btu
NET PRODUCTION (1)				
Net Power, MW	1,213.6	1,213.6	1,148.7	1,148.7
By-product Ammonia ST/SD	128	128	128	128
By-product Sulfur ST/SD				
OPERATING CHARGES, $1000/YEAR				
Coal	62,522	125,044	62,522	125,044
Operating Labor	2,590	2,590	2,692	2,692
Catalyst and Chemicals	429	429	391	391
Utilities	1,179	1,179	1,189	1,189
Maintenance, Labor	7,531	7,531	7,289	7,289
Maintenance, Materials	11,296	11,296	10,934	10,934
Administrative and Support Labor	3,036	3,036	2,994	2,994
General and Administrative Expenses	6,073	6,073	5,989	5,989
Ash Disposal	258	258	258	258
Property Tax/Insurance	16,597	16,597	16,514	16,514
By-product, Ammonia	(3,270)	(3,270)	(3,270)	(3,270)
By-product, Sulfur	(0)	(0)	(0)	(0)
TOTAL OPERATING CHARGES, $1000/Yr	108,241	170,763	107,502	170,024
CAPITAL CHARGES, $1,000/YEAR				
Total Capital Charges	133,087	135,757	132,349	135,019
COST OF SERVICES				
Total, $1,000/year	241,328	306,520	239,851	305,043
Per Unit Production, mills/kWh	32.53	41.32	34.05	43.30

NOTES
(1) At 100% Operating Load Factor.

CASE EALC AND SUBCASE EALC-LP

COMBUSTION ENGINEERING

PLANT DESCRIPTION - CASE EALC

GENERAL

A grass roots plant for power generation using entrained bed low pressure air-blown gasifiers is schematically shown on the block flow diagram, EALC-1-1. This plant is based on gasifying 10,000 ST/day of Illinois No. 6 coal.

This case is an extension of Case EAL, described in the Fuel Gas Report[1], to generate electric power. Refer to this report for the pertinent description and development status of the processes used, and other technical comments on the equipment. The major differences between Cases EALC and EAL occur because of the heat integration of the combined cycle system with the processing units.

The gasification process used for Case EALC operates at atmospheric pressure. Compression of the cleaned product gas to approximately 300 psig is required to supply product gas to the gas turbine combustor chamber operating at 225 psig in Case EALC. An alternate subcase designated as EALC-LP is also considered to evaluate the impact of product gas compression on the overall economics of Case EALC. Gas turbines with lower combustor chamber pressure (126 psig) are considered for this subcase requiring product gas compression to 200 psig. The processing plants for subcase EALC-LP are identical to Case EALC.

The main processing units are in four parallel and largely independent trains. Each process train consists of oxidant feed, gasification, gas-cooling and acid gas removal unit. Oxidant feed, gasification and high temperature gas cooling units are provided with a spare train on standby basis. Integration between processing trains is minimized. Complete trains may be shut down in order to maintain efficiency during reduced capacity operation. The impact of upset conditions is limited to the train in which the upset occurs. In addition to the main processing trains, the complete plant includes necessary offsite, utility and environmental facilities. Coal receiving, storage and conveying is done in a single train to minimize space and operating labor requirements. Coal pulverization and transportation is done in five trains, four operating and one spare, each integrated with the corresponding gasifiers. Hydrogen sulfide from the

1. Economics of Current and Advanced Gasification Processes for Fuel Gas Production, EPRI AF 244, Project 239, Final Report, July 1976.

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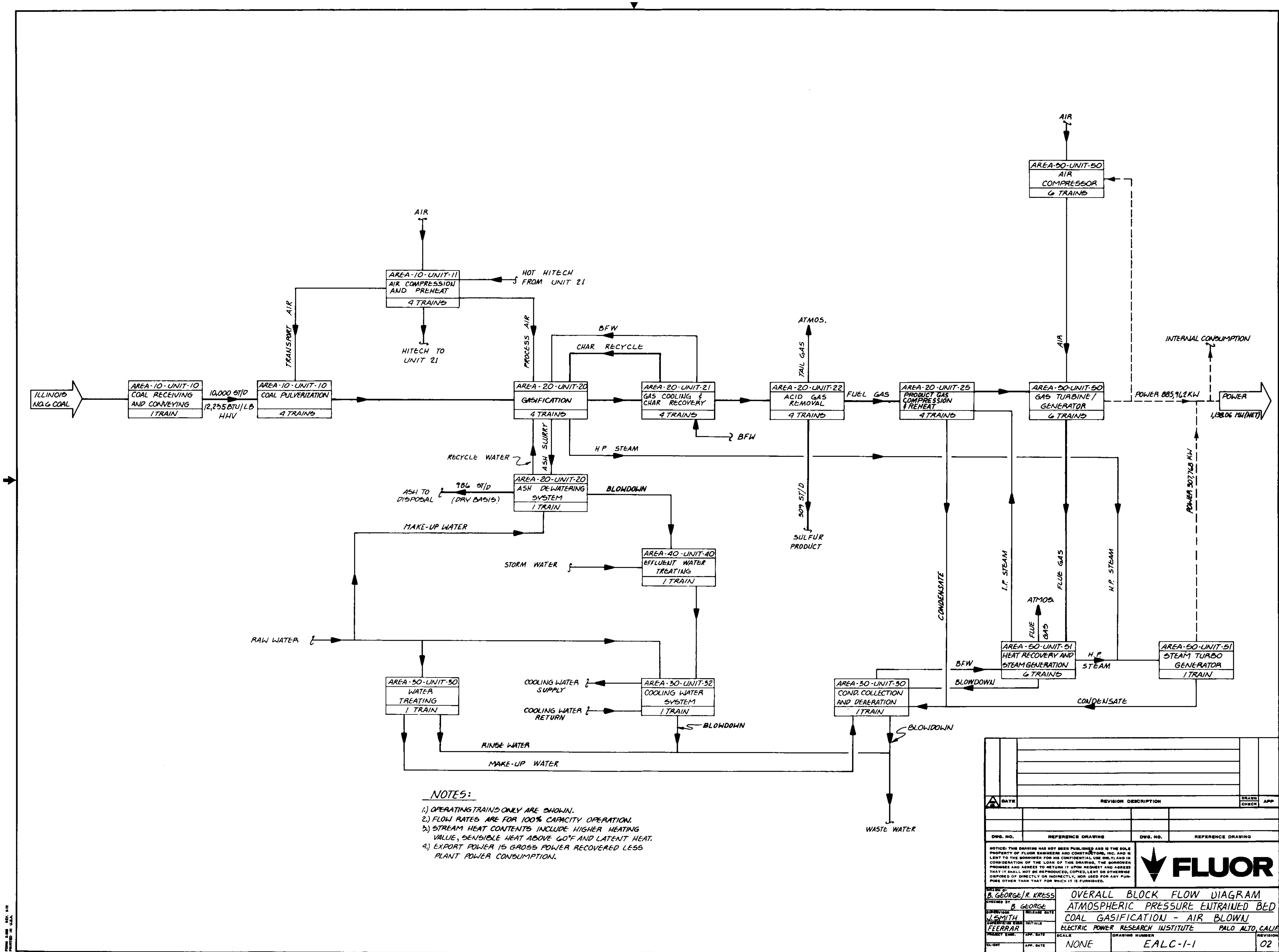

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crude gasifier effuent is removed in the acid gas removal unit and converted to elemental sulfur.

Other facilities in the plant are raw water treating, cooling water and effluent water treating.

Table EL-1 summarizes major equipment sections in the plant and shows the numbers of operating and spare sections.

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TABLE EL-1

MAJOR EQUIPMENT SECTIONS - CASE EALC

Unit	Name	Operating	Spare
10	Coal Handling	1	0
10	Coal Pulverizing and Drying	4	1
11	Air Compression	4	1
11	Air Preheat	4	1
20	Gasification	4	1
20	Ash Handling	1	0
20	High Temperature Gas Cooling	4	1
21	Gas Cooling and Char Recovery	4	0*
22	Acid Gas Removal	4	0
25	Product Gas Compression	4	0
25	Product Gas Heating	4	1
30	Water Treating	1	0
30	Condensate Collection and Deaeration	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0
50	Gas Turbine/Generator	6	0
51	Heat Recovery Steam Generator	6	0
51	Steam Turbine/Generator	1	0

*A spare train is provided for equipment integrated with high temperature gas cooling.

COAL PREPARATION

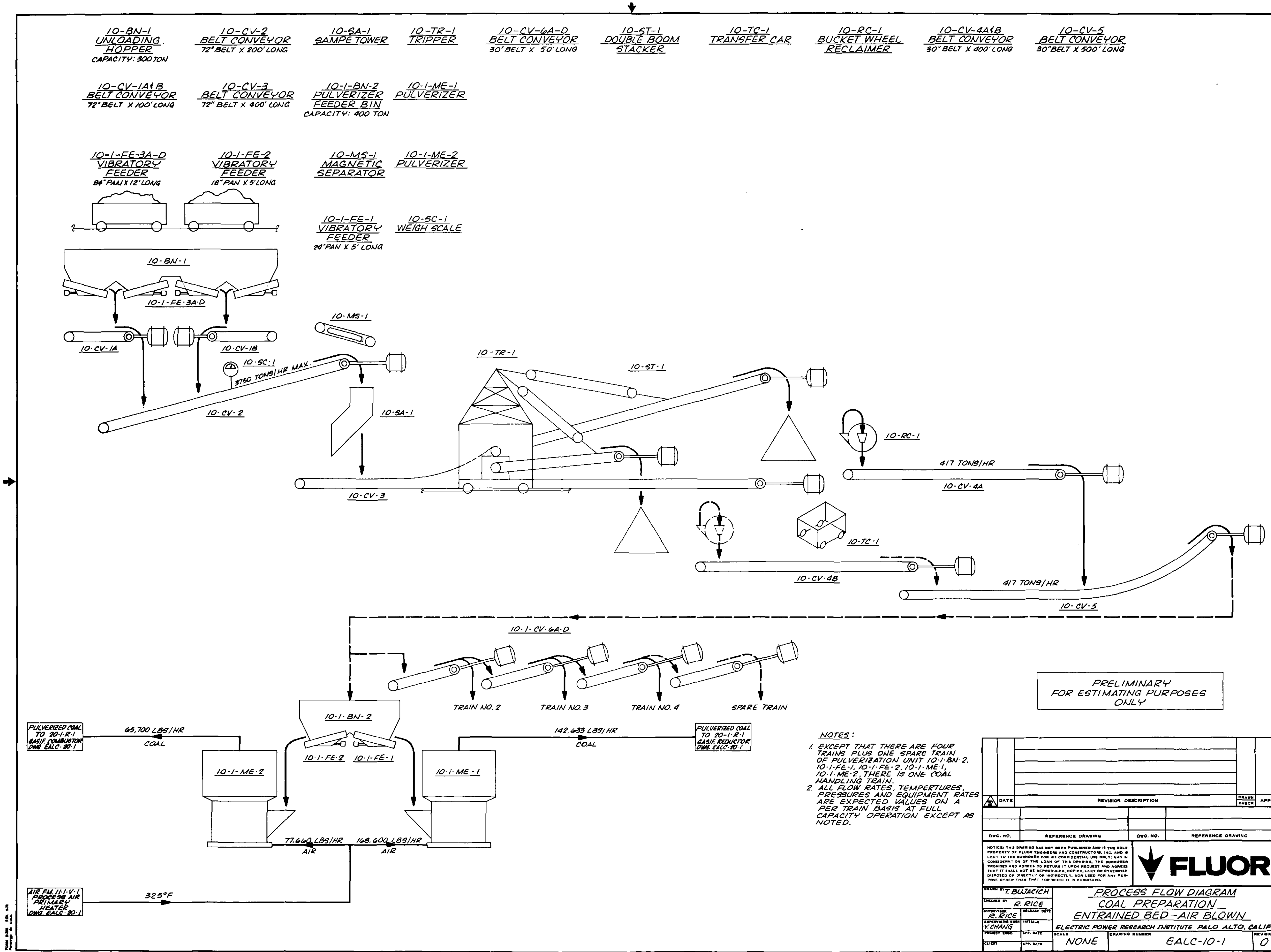
Process Flow Diagram EALC-10-1 depicts the process arrangement of equipment in this section. There is one coal handling train, with the exception of the pulverization unit, which has five parallel trains, four operating and one spare.

Washed, 1-1/2" by 0 coal is received at the plant site by unit train. No breaking and refuse disposal systems are included. The coal feedstock is unloaded from 100 ton bottom dump cars into an unloading hopper, 10-BN-1, withdrawn from the hopper by four vibrating feeders and transported by belt conveyors to a tripper conveyor. The tripper is attached to a double boom traveling belt stacker. The stacker travels on tracks and forms storage piles on both sides of the tracks. The unloading and stacking system is designed to handle a three day supply in eight hours.

Coal is reclaimed from storage piles by a bridge type bucket wheel reclaimer (10-RC-1) rated at 500 tons per hour. This machine is a rail mounted bridge which supports a rotating bucket wheel and belt conveyor. The wheel moves across the face of the pile, making a vertical cut across the many layers of coal. At the end of one cut, the reclaimer moves ahead a small, predetermined distance and the wheel makes another cut in the opposite direction. The excavated coal is carried by a series of conveyors to five hoppers (10-1-BN-2) which feed coal to five pulverization units. Four pulverization units are operating at a time. One pulverization unit is spare. Each pulverizer unit is integrated with one gasifier train in the gasification area (Flow Diagram: EALC-20-1).

Coal is ground to 70 percent passing a 200 mesh screen, classified and dried simultaneously in the pulverizers. The drying gas is 325°F process air from the primary air heater, 11-1-V-1. The pulverized coal is pneumatically conveyed to the gasifiers by the drying gas.

Each pulverizer train consists of two pulverizers, 10-1-ME-1 and 10-1-ME-2. Pulverizer 10-1-ME-1 has a larger capacity and supplies approximately 142,633 lbs/hr of pulverized coal at 100% capacity to the reductor chamber of the gasifier, 20-1-R-1. Pulverizer 10-1-ME-2 supplies approximately 65,700 lbs/hr of pulverized coal to the combustor chamber of the gasifier.



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OXIDANT FEED, GASIFICATION AND ASH HANDLING

Process Flow Diagram EALC-20-1 shows oxidant feed, gasification and ash handling systems. There are five parallel air compression, air preheat and gasification trains, four operating and one spare. There is one ash handling train.

Oxidant Feed

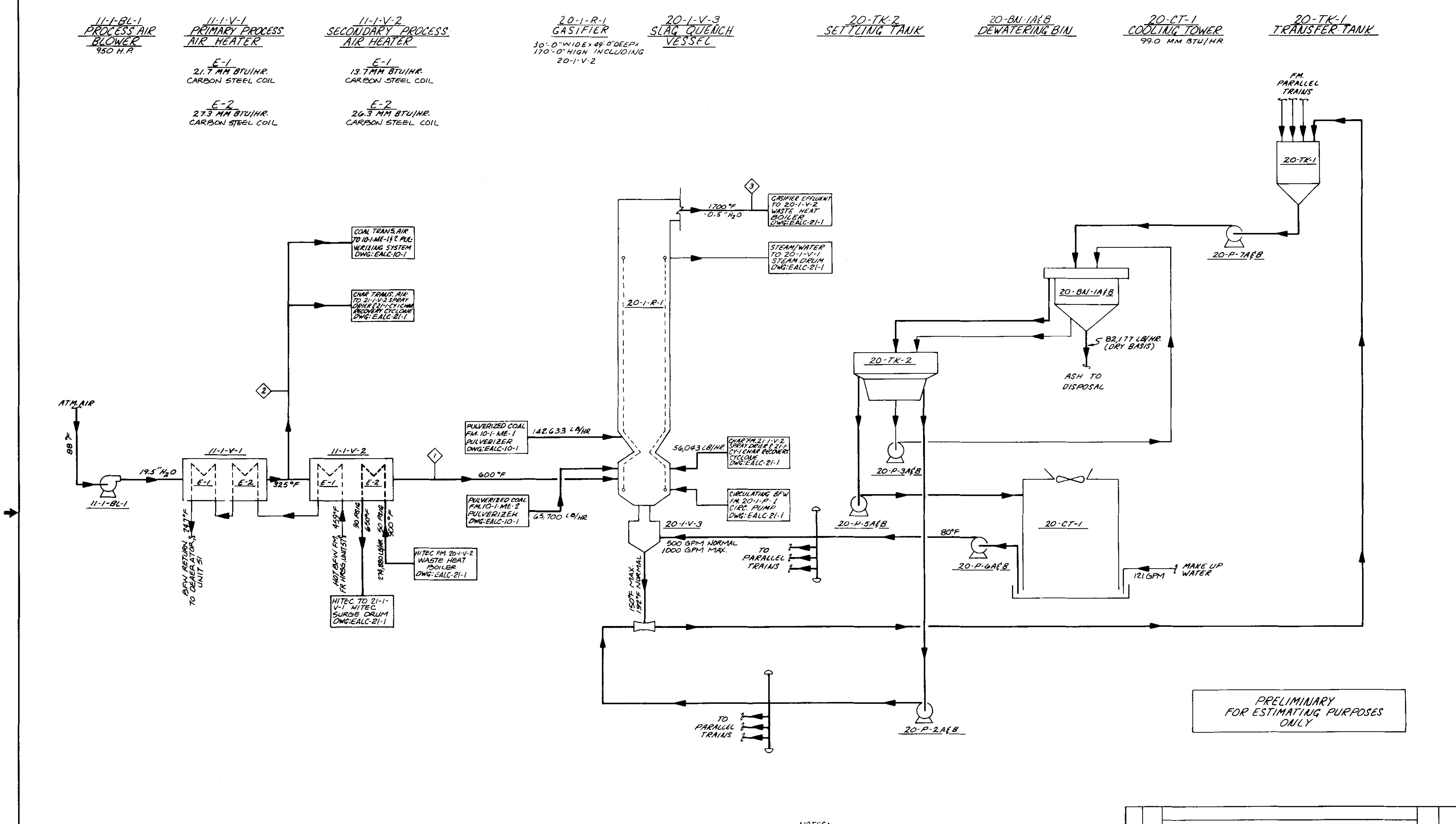
Atmospheric air is compressed to 19.5 inches water gauge pressure in a single stage blower, 11-1-BL-1. Each blower is motor driven and requires about 950 hp.

The compressed air from the blower is preheated to 325°F by heat exchange against the hot boiler feedwater (BFW) in the primary process air heater, 11-1-V-1. Approximately 54% of the preheated process air is used to transport the coal and char to the gasifier and to dry the pulverized coal. The balance of the process air is heated to 600°F in the secondary process air heater, 11-1-V-2, before being fed to the gasifier. The heater, 11-1-V-2, consists of two heating coils, E-1 and E-2. Process air is first heated in heating coil E-1 by hot BFW from the heat recovery steam generator (HRSG). The boiler feedwater then flows back to the HRSGs through 11-1-V-1. Final heating of the process air to 600°F is accomplished in the second heating coil of 11-1-V-2, where heat is exchanged against hot Hitec salt at 900°F from the waste heat boiler, 20-1-V-2:E-2. The Hitec salt from 11-1-V-2 then flows back to the Hitec surge drum, 21-1-V-1, to complete the circuit (Flow Diagram: EALC-21-1).

Gasification and Ash Handling

Pulverized coal, oxidant and recycle char tangentially enter an atmospheric pressure entrained bed gasifier, 20-1-R-1. Flow is through water cooled nozzles at two levels near the bottom of the gasifier. The lower level is a combustor section where all the recycle char and a portion of the coal are reacted with the oxidant. The coal split is controlled to maintain 3200°F in the combustor section. About 32 percent of the pulverized coal is consumed in the combustor section.

The balance of the coal is injected at the higher level (the reductor section) into hot gases leaving the combustor section. Devolatilization of the reductor coal and cracking of the volatile matter occurs in the lower level of the reductor. The gases cool to 1700°F as they flow up through the reductor section as a consequence of endothermic reactions that occur in this section.



MATERIAL BALANCE						
STREAM NUMBER	1 PROCESS AIR		2 TRANSPORT AIR		3 GASIFIER EFFLUENT	
COMPONENT	MPH	MOL.%	MPH	MOL.%	MPH	MOL.%
CH_4	—	—	—	—	0.2	0.00
H_2	—	—	—	—	4,718.0	11.22
CO	—	—	—	—	9,350.3	22.24
CO_2	—	—	—	—	2,138.5	5.08
H_2O	509.08	2.53	276.31	2.53	1,612.19	3.85
H_2S	—	—	—	—	226.0	0.54
COS	—	—	—	—	14.4	0.03
N_2	15,493.66	77.00	8,409.43	77.00	23,978.89	57.04
NH_3	—	—	—	—	0.4	0.00
O_2	4,118.9	20.47	2,235.6	20.47	—	—
TOTAL VAPOR	20,121.64	100.00	10,921.34	100.00	42,038.88	100.00
MOL. WT.	28.58		28.58		25.57	
LB/HR. VAPOR	575,047		312,116		1,074,944	
LB/HR. CHAR	—		—		56,043	
TOTAL LB/HR.	575,047		312,116		1,130,987	

NOTES:

1. THERE ARE FOUR PARALLEL AND ONE SPARE AIR COMPRESSION, AIR PREHEAT AND GASIFICATION TRAINS AND ONE ASH HANDLING UNIT.
2. ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

REV NO	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY K. WILSON
CHECKED BY R. RICE
SUPERVISOR R. RICE
SUPERVISING ENGR Y. CHANG

PROCESS FLOW DIAGRAM
OXIDANT FEED, GASIFICATION & ASH HANDLING
ENTRAINED BED-AIR BLOWN
ELECTRIC POWER RESEARCH INSTITUTE—PALO ALTO, CALIF.

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

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255

Complete gasification of the coal is not obtained in one pass through the gasifier. Unreacted char is swept out of the gasifier in the hot crude gas and recovered in the gas cooling unit. Entrainment is about 27 pounds char per 100 pounds coal feed.

The gasifier is enclosed with water cooled, fin-welded, studded, refractory-covered walls. Boiler feedwater is pumped through the walls and high pressure steam is generated.

Molten slag collects on the combustor walls and drains from the gasifier into a water filled slag quench vessel, 20-1-V-3, where it is quenched and fragmented under controlled conditions. The resultant slurry is educted to a common transfer tank, 20-TK-1, using recycle process water as the motive fluid. Slag grinders prevent large chunks of slag from plugging transfer lines and a slag breaker disintegrates slag icicles at the gasifier tap hole. The slag slurry is dewatered in dewatering bins 20-BN-1A&B, producing an ash ready for disposal. Final cleaning of the water overflowing the dewatering bin is accomplished in a settling tank, 20-TK-2, where slag fines settle and are pumped back to a dewatering bin. A portion of the clarified water is recycled for slag quenching after cooling in an induced draft type cooling tower, 20-CT-1. The balance of the clarified water is pumped to the slag slurry transfer eductors to serve as motive fluid.

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GAS COOLING, CHAR RECOVERY AND PRODUCT GAS COMPRESSION AND REHEATING

Process Flow Diagram EALC-21-1 depicts the gas cooling, char recovery and product gas compression and reheating systems. Except for the high temperature gas cooling section and secondary char recovery unit (21-ME-1), there are a total of four parallel trains. There are five parallel trains of high temperature gas cooling section (four operating, one spare) and one secondary char recovery unit.

The gasifier effluent is cooled to 232°F in a waste heat boiler (20-1-V-2) by heat exchange with other fluids. The gasifier effluents flow downward from the top in 20-1-V-2. There are the following eight heat recovery coils (E-1 through 8) stacked from top to bottom in 20-1-V-2:

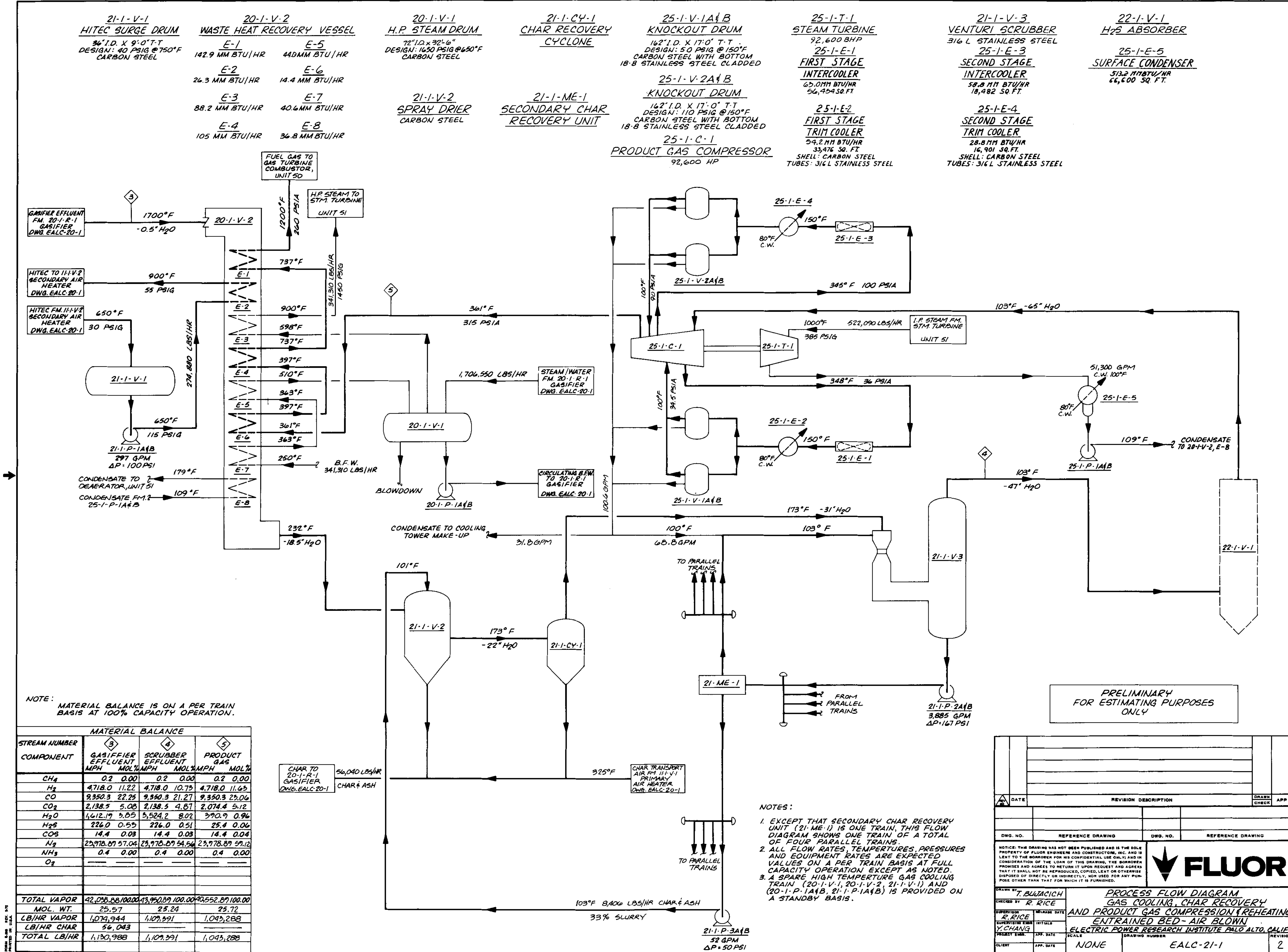
- Fuel gas heating coils (E-6, E-4, E-1)
- Hitech heating coil (E-2)
- High pressure steam superheating coil (E-3)
- Boiler feedwater heating coils (E-7, E-5)
- Condensate heating coil (E-8)

Fuel gas from product gas compressor, 25-1-C-1, at 300 psig and 361°F is heated to 1200°F in coils E-6, E-4 and E-1. Fuel gas flows through these coils in series, in the given order, to the gas turbines.

Hitec salt is circulated from the surge drum, 21-1-V-1, through coil E-2 where it is heated from 650°F to 900°F. The hot Hitec salt then flows to the upstream gasification unit (Flow Diagram: EALC-20-1) for process air heating. Boiler feedwater (BFW) from deaerator 51-DA-1 is heated to 510°F in Coils E-7 and E-5, and flows to 20-1-V-1. BFW is continuously circulated through the gasifiers' walls and high pressure steam is generated. High pressure (HP) steam from the HP steam drum (20-1-V-1) is superheated to 900°F in coil E-3 and flows to the combined cycle system (Unit 51).

The final gas cooling is obtained in coil E-8 by heat exchange against a cold condensate stream which flows to the deaerator, 51-DA-1.

The cooled crude effluent gas-char stream from 20-1-V-2 then flows through a spray dryer, 21-1-V-2, into which a 33 percent by weight char slurry from the secondary char recovery unit (21-ME-1) is sprayed. The purpose of the spray



NOTE:
MATERIAL BALANCE IS ON A PER TRAIN BASIS AT 100% CAPACITY OPERATION.

MATERIAL BALANCE						
STREAM NUMBER	3		4		5	
COMPONENT	GASIFFIER EFFLUENT		SCRUBBER EFFLUENT		PRODUCT GAS	
	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	0.2	0.00	0.2	0.00	0.2	0.00
H_2	4,718.0	11.22	4,718.0	10.73	4,718.0	11.63
CO	9,350.3	22.25	9,350.3	21.27	9,350.3	23.06
CO_2	2,138.5	5.08	2,138.5	4.87	2,074.4	5.12
H_2O	1,612.19	3.85	3,524.2	8.02	390.9	0.96
H_2S	226.0	0.53	226.0	0.51	25.4	0.06
COS	14.4	0.03	14.4	0.03	14.4	0.04
N_2	23,978.89	57.04	23,978.89	54.56	23,978.89	59.12
NH_3	0.4	0.00	0.4	0.00	0.4	0.00
O_2	—	—	—	—	—	—
TOTAL VAPOR	42,038.88	100.00	43,950.89	100.00	40,552.89	100.00
MOL. WT.	25.57		25.24		25.72	
LB/HR VAPOR	1,079,944		1,109,391		1,093,288	
LB/HR CHAR	56,043		—		—	
TOTAL LB/HR	1,130,988		1,109,391		1,093,288	

NOTES:

1. EXCEPT THAT SECONDARY CHAR RECOVERY UNIT (21-ME-1) IS ONE TRAIN, THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FOUR PARALLEL TRAINS.
2. ALL FLOW RATES, TEMPERTURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
3. A SPARE HIGH TEMPERTURE GAS COOLING TRAIN (20-1-V-1, 20-1-V-2, 21-1-V-1) AND (20-1-P-1A & B, 21-1-P-1A & B) IS PROVIDED ON A STANDBY BASIS.

DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY T. BUJACICH
CHECKED BY R. RICE
SUPERVISOR R. RICE
SUPERVISING ENGR Y. CHANG
RELEASE DATE
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
GAS COOLING, CHAR RECOVERY
AND PRODUCT GAS COMPRESSION & REHEATING
ENTRAINED BED - AIR BLOWN
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: EALC-21-1
REVISION: 2

dryer is to dry and agglomerate the solids which are collected in the wet secondary char recovery unit before they are recycled to the combustor section of the gasifier. Some of these solids are collected in the hopper bottom of the spray dryer and the rest are collected in a char recovery cyclone, 21-1-CY-1, through which the dryer effluent gas passes. Recovered char is pneumatically transported to the gasifier with process air. About 85 percent of the char entrained in the gasifier is recovered in the dryer and cyclone.

The cooled crude gas stream from the char recovery cyclone flows through a venturi scrubber, 21-1-V-3, where it is washed with water to remove the remaining particulate matter. Scrubber water, approximately 2 wt percent char, is pumped from the bottom of the scrubber knockout vessel, combined with slurry streams from parallel trains, and fed to a secondary char recovery unit. Hydroclones thicken the 2 wt percent bottoms to 33 wt percent slurry. The filtrate from the recovery unit is recycled back to the venturi scrubbers. The underflow from the recovery unit is sprayed into the top of the spray dryers, where the hot gas dries the atomized slurry to a dry, free flowing powder.

A portion of the condensate produced in interstage coolers of the downstream product gas compressors (25-1-C-1) is added to the scrubbers to maintain the water balance in the scrubbing system. The scrubbed char free crude gas from the scrubber knockout vessel flows to the H_2S absorber, 22-1-V-1, of the acid gas removal system.

The desulfurized product gas from the H_2S absorber, 22-1-V-1, is compressed to 300 psig in three stage product gas compressors (25-1-C-1). The compressors are driven by condensing turbines, 25-1-T-1. Steam at 385 psig, 1000°F is supplied to these turbines from the heat recovery steam generators (HRSG) located in combined cycle system (Unit 51). The exhaust steam from the turbines at 2-1/2" Hg abs. is condensed in 25-1-E-5 and pumped through 20-1-V-2:E-8 to deaerator 51-DA-1. The heat of compression is rejected to air and cooling water in the interstage coolers, 25-1-E-1 through 4.

The compressed product gas at 300 psig and 361°F is further heated in the waste heat recovery vessel, 20-1-V-2, to 1200°F and transported to the gas turbines located in Unit 50.

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Equipment Notes

Refer to Case EAL of the Fuel Gas Report[1], for the pertinent comments regarding the equipment state of art applicable to the gasification and gas cooling section.

The product gas compressor is commercially proven. The technology for double flow centrifugal compressors up to 400,000 ICFM is available. The operating pressure range is not a problem. However, only a limited number of machines with such large capacities (approximately 350,000 ICFM) have been built. Consequently, the available operating experience on these machines is limited.

1. Economics of Current and Advance Gasification Processes for Fuel Gas Production, EPRI-AF-244, Project 239, Final Report, July 1976.

ACID GAS REMOVAL

The acid gas removal unit employs the Stretford process for the selective removal of H_2S from the cooled crude gas. This unit is similar to the one described for Case EAL in the Fuel Gas Report[1]. Refer to this report for a process description and typical process flow diagram of the acid gas removal system. (Flow Diagrams MACW-23-2 and MACW-23-3, included in this report for Case MACW, excluding Beavon's front end modification, also represent a typical Stretford plant.)

There are four parallel operating trains. No spare train is provided.

The Stretford process is used because of the low pressures prevailing in this case. The Stretford process can efficiently absorb H_2S at low concentration from low pressure gas streams because it is a chemical absorption process rather than a physical one (like Selexol). Hydrogen sulfide (H_2S) is absorbed and oxidized directly to elemental sulfur using a chemically active solution of sodium meta vanadate and anthraquinone disulfonic acid. The sulfur content of the gas is reduced to an equivalent of 1.0 pound sulfur dioxide per million Btu (HHV) coal feed to the gasifiers. About 309 ST/day of total sulfur is produced in the plant. The sulfur in the product gas from the acid gas removal unit is 978 ppm on volume basis.

Equipment Notes

These Stretford units are two to three times larger than units currently in operation. The scaleup does not present any unusual problems. The Stretford process requires a somewhat larger plot area than the Selexol units used for this service in the other cases reported here.

1. Economics of Current and Advanced Gasification processes for Fuel Gas Production, EPRI AF 244, Project 239, Final Report, July 1976.

STEAM, BOILER FEEDWATER AND CONDENSATE

Process Flow Diagram EALC-30-1 schematically represents steam, boiler feedwater and condensate systems.

The steam system operates at four pressure levels:

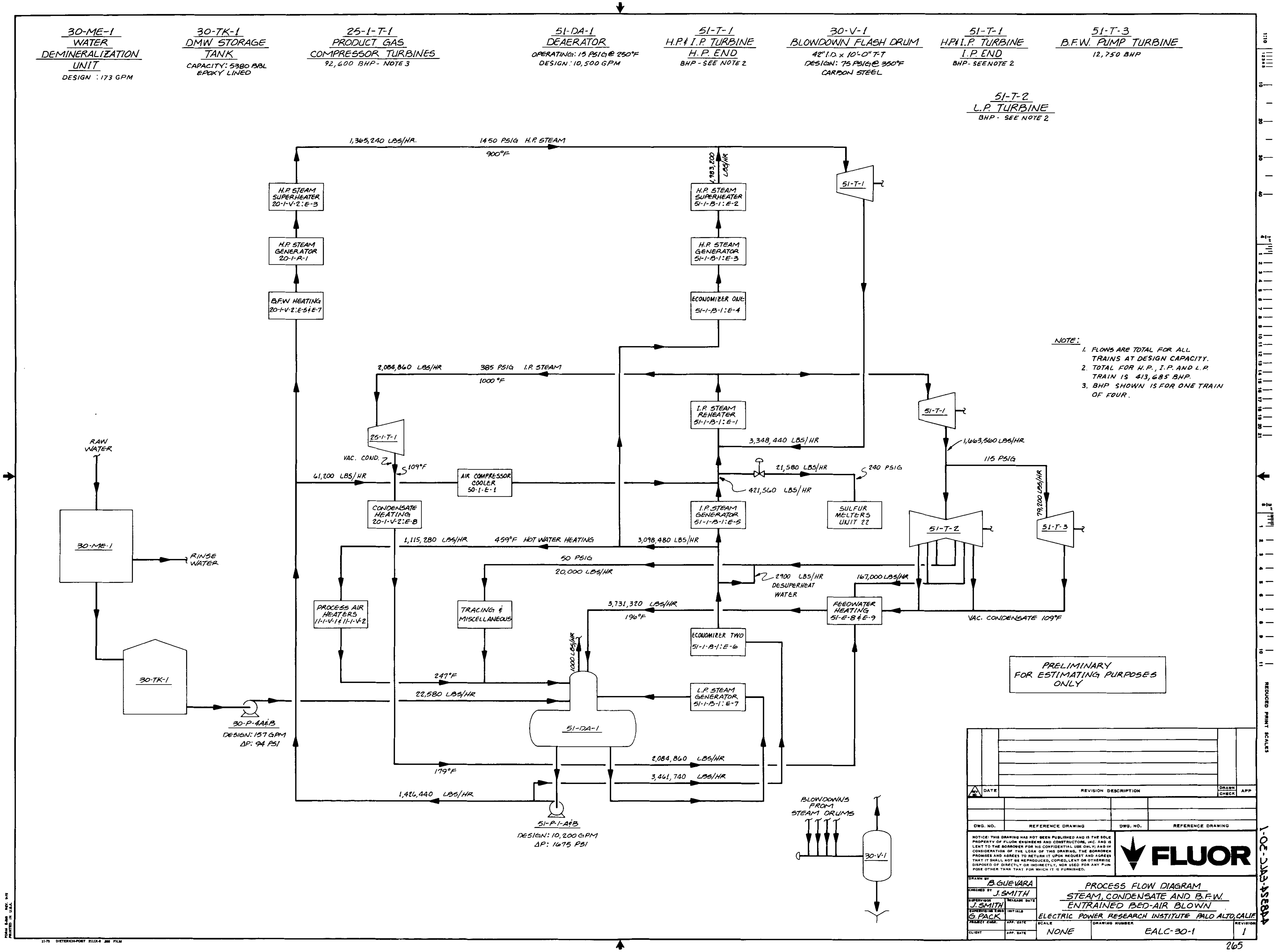
- High Pressure Superheated - 1450 psig, 900°F
- Intermediate Pressure - 400 psig
- Low Pressure - 100 psig
- Low Pressure - 500 psig

Major high pressure (HP) steam generation is carried out in four operating gasifiers, 20-1-R-1, in the gasification area and six heat recovery steam generators (HRSGs), 51-1-B-1, coupled with gas turbines, 50-1-GT-1, in the combined cycle systems. The HRSG is described in detail in Appendix A. All the HP steam is used to drive the single backpressure type turbine, 51-T-1.

The high pressure end of 51-T-1 takes steam at 1450 psig, 900°F and exhausts to a 400 psig steam header. Additional 400 psig steam generation is obtained in the HRSGs (51-1-B-1:E-5). Part of the 400 psig steam from the HRSGs is let down to 240 psig and supplied to sulfur melters in the acid gas removal unit. The balance combines with 51-T-1 exhaust steam and is superheated to 1000°F in the intermediate pressure steam reheaters (51-1-B-1:E-1) of the HRSGs. Part of the superheated intermediate pressure (IP) steam is supplied to the product gas compressors' steam turbine drives, 25-1-T-1, and the rest is used in the IP end of the turbine, 51-T-1, which exhausts to a 100 psig steam header.

The 100 psig steam is used in the low pressure (LP) turbine, 51-T-2, and a condensing type boiler feedwater (BFW) pump turbine, 51-T-3. The LP turbine is condensing type turbine exhausting at 2-1/2" Hg abs. The turbine is provided with the capability of steam extraction at 50 psig and 13 psia. The 13 psia steam is used in the feedwater heaters.

About 17,100 lbs/hr of 50 psig steam is extracted from the LP turbine and supplied to the 50 psig steam header. The 50 psig steam is desuperheated and used for steam tracing and miscellaneous small users.


30-ME-1
WATER DEMINERALIZATION UNIT
DESIGN : 173 GPM
30-TK-1
DMW STORAGE TANK
CAPACITY: 5380 BBL
EPOXY LINED
25-1-T-1
PRODUCT GAS COMPRESSOR TURBINES
92,600 BHP - NOTE 3
51-DA-1
DEAERATOR
OPERATING: 15 PSIG @ 250°F
DESIGN: 10,500 GPM
51-T-1
H.P. & I.P. TURBINE H.P. END
BHP - SEE NOTE 2
30-V-1
BLOWDOWN FLASH DRUM
42" I.D. x 10'-0" T-T
DESIGN: 75 PSIG @ 350°F
CARBON STEEL
51-T-1
H.P. & I.P. TURBINE I.P. END
BHP - SEE NOTE 2
51-T-3
B.F.W. PUMP TURBINE
12,750 BHP
51-T-2
L.P. TURBINE
BHP - SEE NOTE 2
1,365,240 LBS/HR.
1450 PSIG H.P. STEAM
900°F
1,983,200 LBS/HR
H.P. STEAM SUPERHEATER 20-1-V-2:E-3
H.P. STEAM GENERATOR 20-1-R-1
B.F.W. HEATING 20-1-V-2:E-5 & E-7
H.P. STEAM SUPERHEATER 51-1-B-1:E-2
H.P. STEAM GENERATOR 51-1-B-1:E-3
ECONOMIZER ONE 51-1-B-1:E-4
2,084,860 LBS/HR
385 PSIG I.P. STEAM
1000°F
I.P. STEAM REHEATER 51-1-B-1:E-1
3,348,440 LBS/HR
1,663,560 LBS/HR
115 PSIG
RAW WATER
25-1-T-1
VAC. COND.
109°F
61,200 LBS/HR
AIR COMPRESSOR COOLER 50-1-E-1
21,580 LBS/HR
240 PSIG
421,560 LBS/HR
SULFUR MELTERS UNIT 22
78,200 LBS/HR
51-T-2
51-T-3
30-ME-1
RINSE WATER
CONDENSATE HEATING 20-1-V-2:E-8
I.P. STEAM GENERATOR 51-1-B-1:E-5
1,115,280 LBS/HR
459°F HOT WATER HEATING
3,098,480 LBS/HR
50 PSIG
20,000 LBS/HR
2900 LBS/HR DESUPERHEAT WATER
167,000 LBS/HR
PROCESS AIR HEATERS 11-1-V-1 & 11-1-V-2
TRACING & MISCELLANEOUS
3,731,320 LBS/HR
190°F
FEEDWATER HEATING 51-E-8 & E-9
VAC. CONDENSATE 109°F
ECONOMIZER TWO 51-1-B-1:E-6
1000 LBS/HR
30-TK-1
247°F
22,580 LBS/HR
L.P. STEAM GENERATOR 51-1-B-1:E-7
30-P-4A&B
DESIGN: 157 GPM
ΔP: 94 PSI
51-DA-1
179°F
2,084,860 LBS/HR
3,461,740 LBS/HR
1,426,440 LBS/HR
51-P-1-A&B
DESIGN: 10,200 GPM
ΔP: 1675 PSI
BLOWDOWNS FROM STEAM DRUMS
30-V-1
NOTE:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. TOTAL FOR H.P., I.P. AND L.P. TRAIN IS 413,685 BHP.
3. BHP SHOWN IS FOR ONE TRAIN OF FOUR.
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
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REFERENCE DRAWING
FLUOR
DRAWN BY B. GUEVARA
CHECKED BY J. SMITH
SUPERVISOR J. SMITH
G. PACK
PROCESS FLOW DIAGRAM
STEAM, CONDENSATE AND B.F.W.
ENTRAINED BED-AIR BLOWN
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER EALC-30-1
REVISION 1
265
REDUCED PRINT SCALES

Raw water is treated in a semiautomatic, resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for a 1500 psig boiler system. Storage equivalent to 24 hours of demineralized water production is provided. The demineralized water is pumped from the storage tank, 30-TK-1, to the deaerator, 51-DA-1.

The deaerator is a tray type unit operating at 28 psia. The deaerating steam is supplied by the HRSGs (51-1-B-1:E-7). The deaerator provides for ten minutes storage capacity.

The vacuum condensate from the turbines, 25-1-T-1, is heated in the condensate heating coil (E-8) of 20-1-V-2 (Flow Diagram: EALC-20-1). The hot condensate then combines with the vacuum condensate from the LP turbine, 51-T-2. The combined condensate steam is heated in the feedwater heaters (51-E-8 and E-9) by 13 psia extraction steam from the LP turbine. The condensate stream flow is on the tube side of the feedwater heaters. Hot condensate then flows to the deaerator. The 13 psia steam condensate from the shell side flows to the surface condenser of 51-T-2 where it mixes with the vacuum condensate.

Boiler feedwater (BFW) makeup for steam generation is pumped from the deaerator by high pressure BFW pumps (51-P-1A&B) to the gasifiers and the HRSGs.

The gasifier steam generation system is shown in Process Flow Diagram EALC-21-1. BFW for the gasifiers' steam generation is heated to 510°F by exchanging heat against gasifier effluents in the waste heat recovery vessels (20-1-V-2:E-7 and E-5). The hot BFW makeup then flows to the steam drums, 20-1-V-1. Saturated high pressure steam is generated in the gasifiers' walls and flows back to 20-1-V-1. The high pressure (HP) steam is further superheated to 900°F in 20-1-V-2:E-3 and flows to the superheated HP steam header.

Boiler feedwater heating, IP and HP steam generation and superheating are accomplished in the HRSGs by heat recovery from the flue gases produced in the gas turbines, 50-1-GT-1. BFW is first heated in the economizer two (51-1-B-1:E-6). Makeup BFW for IP steam generation is withdrawn downstream of the economizer two and transported to the IP evaporator (51-1-B-1:E-5). A portion of the hot BFW also flows through the process air heaters, 11-1-V-1 and 11-1-V-2, and back to the deaerator. A small quantity of the BFW is let down to supply the desuperheating water for the 50 psig steam header. The balance of the BFW is then further heated to HP steam saturation temperature in the economizer one (51-1-B-1:E-4) and flows to the HP evaporators (51-1-B-1:E-3) where HP steam is generated. The HP steam is superheated to 900°F in 51-1-B-1:E-2 and flows to the superheated HP steam header.

COMBINED CYCLE SYSTEM

Process Flow Diagrams EALC-50/51-1 and EALC-50/51-2 schematically represent the combined cycle system for Case EALC and Sub-Case EALC-LP, respectively. These diagrams show the total power block flows.

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

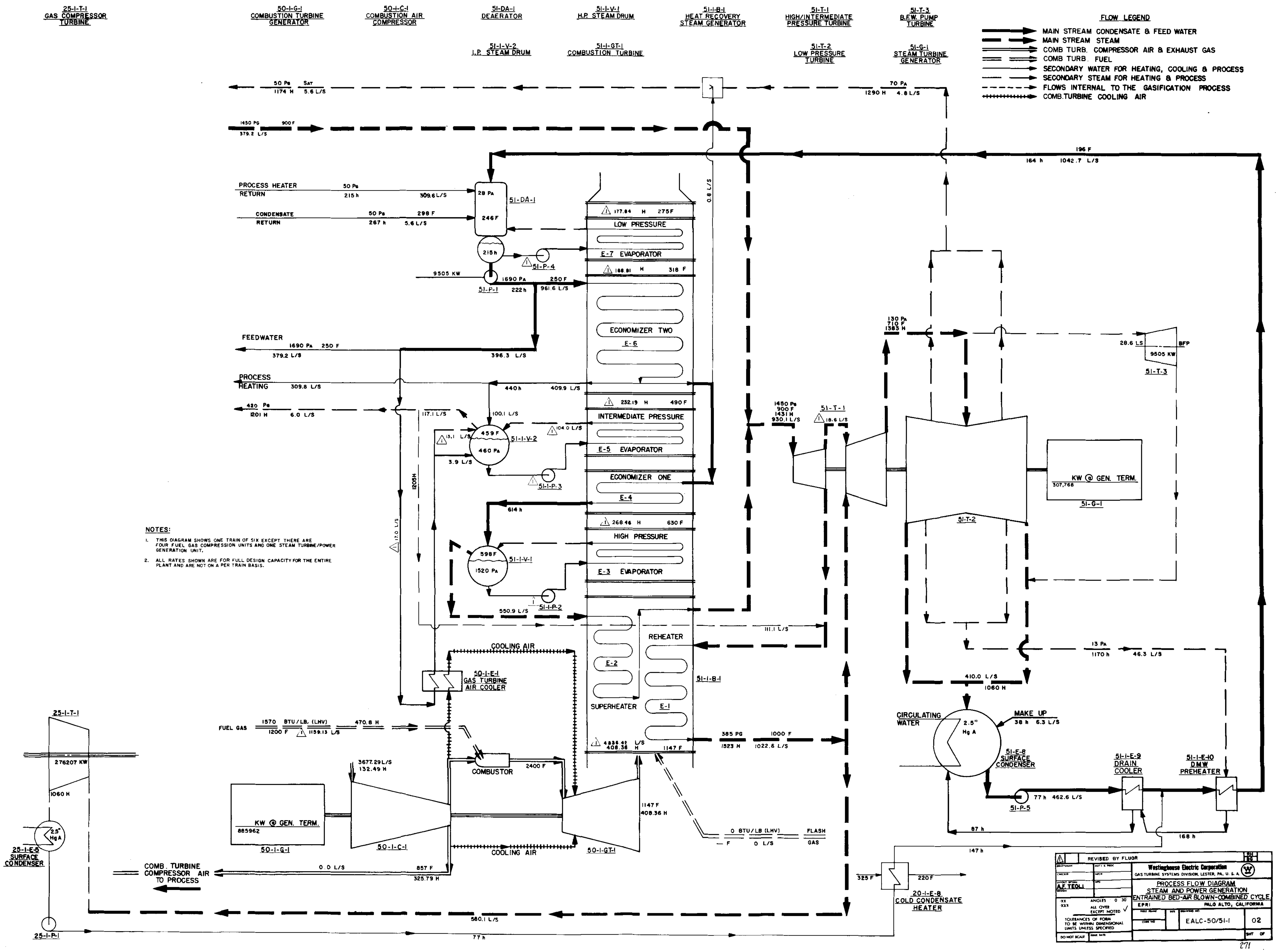
Approximately 40 percent of the total superheated high pressure (HP) steam generated in the plant is supplied by the process steam generators in Case EALC. About 34 percent of the total superheated HP steam produced in the plant is supplied by the process steam generators in Subcase EALC-LP. In both cases the superheated HP process steam together with superheated HP steam generated in the HRSGs is used for power recovery in the turbine, 51-T-1.

The following features are applicable to both Case EALC and Subcase EALC-LP:

- Steam for the boiler feed water pump drives is taken from steam cross over from IP to LP steam turbine cylinders. Steam for the fuel gas compressor drives is taken from the hot reheat line in both cases.

- Hot feed water at 450°F from the outlet of economizer two, 51-1-B-1: E-6, of the HRSG is supplied to the process plants to meet the process heating requirement in both cases.

The process cooling loads, where possible, are integrated into the condensate and makeup system. Approximately 146 MM Btu/hr of low level process heat is recovered by heating a cold condensate stream in the waste heat recovery vessel, 20-1-V-2:E-8 (Flow Diagram: EALC-21-1). However, Case EALC requires a small quantity of steam extraction from 51-T-2 at 13 psia for additional condensate heating while in Case EALC-LP steam extraction from 51-T-2 is required at 15.3 psia and 5.6 psia for this service.



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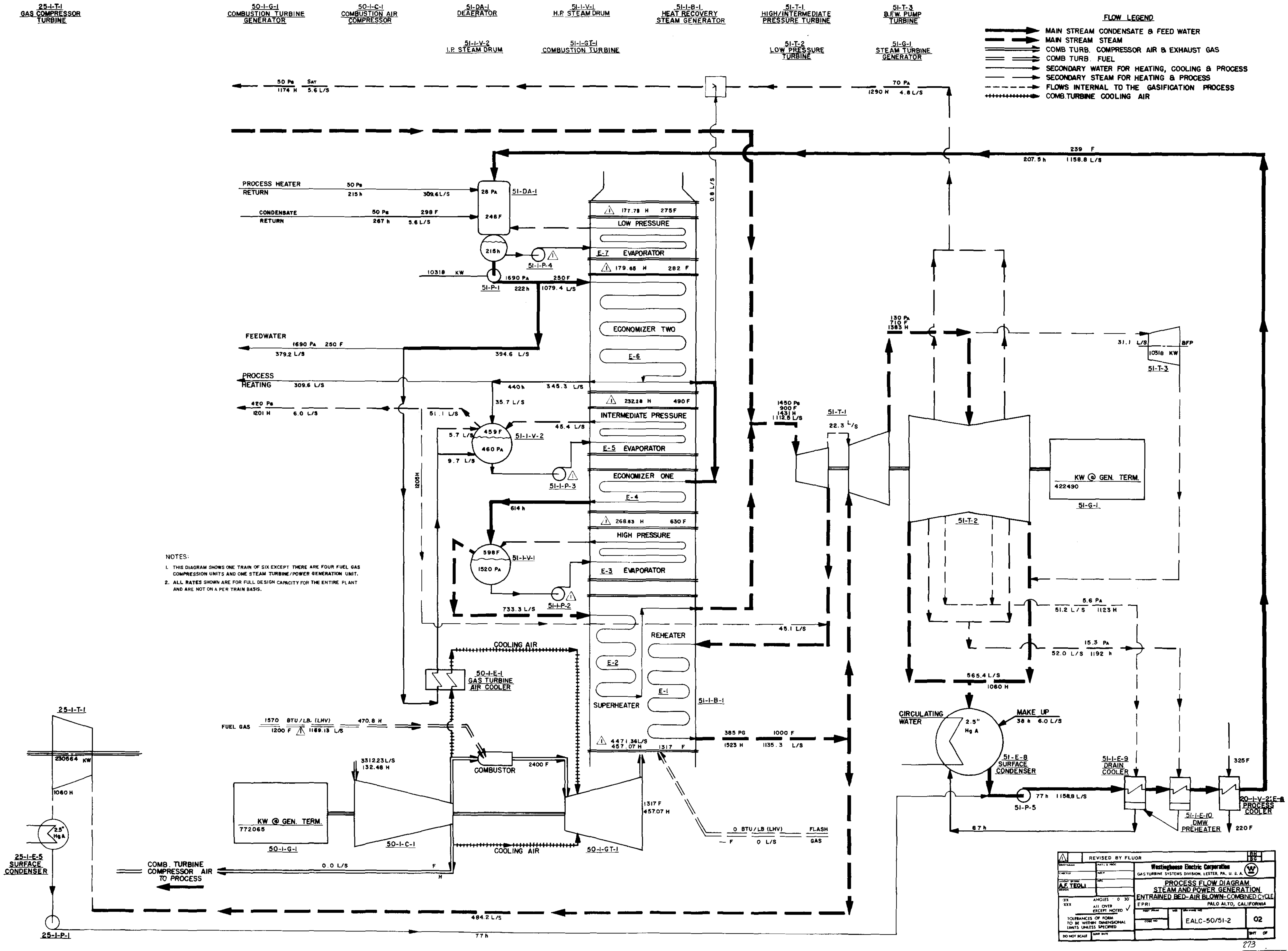

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Equipment Notes

The thermodynamic design of gas turbines selected for Subcase EALC-LP differs from Case EALC in operating pressure. The gas turbine for Subcase EALC-LP operates at the specified lower 10:1 pressure ratio while the gas turbine for Case EALC operates at a pressure ratio to result in combustor shell pressure of 225 psig. This latter pressure is consistent with that of the other combined cycle cases.

The general comments made in Appendix A on the equipment state-of-art are applicable to both Case EALC and Subcase EALC-LP.

The following additional comments are applicable to Subcase EALC-LP only.

A design prerequisite for the gas turbine (501 engine) is the temperature limit of the last rotating turbine blade. This blade is characterized by large taper ratio and high twist and is not readily adaptable to conventional cooling techniques. Therefore, the stress limit of this uncooled turbine blade defines a growth or conditional limit.

For a given blade size and shape, stress becomes a function of blade material and temperature, with temperature a function of turbine inlet temperature and cycle pressure ratio.

In Subcase EALC-LP, with the turbine inlet temperature specified at 2400°F at a pressure ratio of 10:1, the exhaust gas temperature is approximately 1320°F, which corresponds to a temperature approximately 100°F hotter or 1420°F, on the last rotating turbine blade.

For a blade stressed to the level of the 501 size unit, blade metal temperatures in the range of 1200 to 1280°F are required for a 100,000 hour design life. Even at these temperatures, development work is required to forge or cast large blades from the newer identified blade materials, such as U-710 and IN-792.

To operate at the temperature levels of the 10:1 pressure ratio cycles, a departure from the standard design philosophy must be made and alternatives sought. Possible approaches that would permit the higher last row blade temperature are:

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1. Develop a stronger high temperature material for use in the last rotating row of turbine blading.

2. Develop blade cooling.

3. Reduce blade loading, thus stress, by reduced turbine flow. This can be accomplished by:

   a. Using small, lower mass flow, multiple combustion turbine units.

   b. The current design reported here is based on providing the gas turbine air compressor, the gas turbine and the generator operating at a speed of 3600 rpm on the same shaft. The gas turbine provides the power to drive the air compressor and the excess power is transmitted to the generator.

      The blade loading can be reduced by providing the air compressor and gas turbine on the same shaft and using a double flow expander for transmitting power to a separate shaft. Fuel gas from the process is fired in the gas turbine to extract adequate power to drive the air compressor. The exhaust gases from the gas turbine, still at high pressure and temperature, then flow to a double flow expander (with two turbine elements) where power is extracted and supplied to the power shaft.

      As the total mass flow rate is split in half for each turbine element, the size of the last wheel blade of each turbine element of the flow expander is reduced.

   c. Provide the gas turbine and air compressor on the same shaft and use a slower speed (1800 rpm) single flow expander for transmitting power to a separate shaft. The gas turbine will supply adequate power to drive the air compressor. The gas turbine exhaust gases flow to the single flow expander where power is extracted and supplied to the power shaft.

4. Reduce blade design life to a level commensurate with other machine "hot parts" life.

Although each of the alternates are possible solutions to the blade temperature limit problem, they are not considered in any way as part of this study. A final design configuration may employ any one of the above or a combination, thereof.

PROCESS DISCUSSION

The gasifier used for case EALC is an atmospheric pressure, two stage, entrained device currently being developed by Combustion Engineering, Inc. Compression of the cleaned product gas to gas turbine combustor pressure is therefore required. The discharge pressure of the product gas compressor is determined by the gas turbine combustor pressure. In Case EALC the selected gas turbine combustor pressure is 239.4 psia, which is consistent with other cases considered in this report. The product gas compressor operates at 315 psia discharge pressure. A total pressure drop of approximately 75 psi is allowed for fuel gas heating in exchangers 20-1-V-2:E-6, 20-1-V-2:E-4 and 20-1-V-2:E-1 (Flow Diagram: EALC-21-1) and other friction losses.

For the other cases presented in this report the selection of a 239.4 psia combustor/turbine combination posed no difficulty, in that the gasifer pressures were sufficiently high to allow the product to freely flow to the combustor. This was not the case for EALC. Using a high pressure ratio turbine, so that an expensive product gas compressor is required, somewhat unfairly penalizes this case. Accordingly, Fluor investigated an alternate case using a lower compression ratio turbine (combustor pressure at 140.4 psia). This alternate case is designated EALC-LP. The product gas compressor discharge pressure in this case is 215 psia. The economics of this case are somewhat more favorable than for EALC, as is discussed in the economics section. This alternate case does, in fact, have a significantly reduced requirement for product gas compression. However, the net power produced from the coal feedstock is substantially equal in the two cases.

Table EL-2 summarizes the pertinent results.

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TABLE EL-2

SUMMARY OF SYSTEM PERFORMANCE

GASIFICATION AND GAS CLEANING SYSTEM	Case EALC	Subcase EALC-LP
Coal Feed Rate, lbs/hr (m.f.)	---798,333---	
Oxygen or Air(1)/Coal Ratio, lbs/lb m.f.	--- 4.37 ---	
Oxidant Temperature, °F	--- 437 ---	
Gasifier Exit Pressure, inches H_2O	--- -0.5 ---	
Crude Gas Temperature, °F	--- 1700 ---	
Crude Gas HHV (dry basis), Btu/SCF (2)	--- 113.0 ---	
Temperature Fuel Gas to Gas Turbine, °F	--- 1200 ---	
POWER SYSTEM		
Gas Turbine Outlet Temperature, °F	2400	2400
Pressure Ratio	17:1	10:1
Turbine Exhaust Temperature, °F	1147	1317
Steam Conditions, psig/°F/°F	1450/900/1000	1450/900/1000
Condenser Pressure, inches Hg abs.	2.5	2.5
Stack Temperature, °F	275	275
Gas Turbine Power, MW (3)	886	772
Steam Turbine Power, MW (3)	307	422
Power Consumed, MW	55	55
Net System Power, MW	1138	1139
OVERALL SYSTEM		
Process and Deaerator Makeup Water, gpm/1000 MW	157	157
Cooling Tower Makeup Water, gpm/1000 MW	7439	7850
Cooling Water Circulation Rate, gpm/MW	343	361
Cooling Tower Heat Rejection, % of coal HHV	37.6	39.6
Air Cooler Heat Rejection, % of coal HHV	4.9	4.7
Net Heat Rate, Btu/kWh	8959	8951
Overall System Efficiency (Coal → Power), % of coal HHV	38.10	38.13

Notes:

(1) Dry basis, 100% O_2 for oxygen blown.
(2) Excluding HHV of H_2S, COS, and NH_3.
(3) At generator terminals.

Gasifier Material Balance

Table EL-3 presents gasifier material balance for both Case EALC and Subcase EALC-LP. Several observations about these cases are notable:

- There is no steam feed to the gasifier as there is with other gasification processes considered in this study.
- There is practically no methane in the effluent and no higher hydrocarbons.
- There is very little ammonia in the effluent. All the process condensate produced in gas cooling is recycled to the gasifier and a process condensate unit is not required.

Approximately 309 short tons/day of elemental by-product sulfur is made in the plant.

Acid Gas Removal

The Stretford Process is used for this section because of the low pressure prevailing in this case. The Stretford Process can efficiently absorb H_2S at low concentrations from low pressure gas streams because it is a chemical absorption process rather than a physical one (like Selexol).

As noted in the Plant Description, the Stretford units are two to three times larger than the largest units currently in operation. This scaleup does not present any unusual problems. The process does have the disadvantage of requiring very large plot areas.

As specified for these studies, the process removes H_2S to give an SO_2 emission level of 1.0 lb SO_2 per million Btu (HHV) coal to the gasifiers from combustion of the product fuel gas. H_2S removal is on the order of 90 percent. In an actual application the H_2S removal may be nearly 100 percent because there is no practical means to limit H_2S absorption. Whereas in most commercial acid gas removal processes absorption is limited by equilibrium relations, there is no such limitation with the Stretford solution. The process typically removes H_2S down to several parts per million in the treated gas. In both cases H_2S in the treated gas is 998 ppm. Should the unit have to be designed for 100 percent H_2S removal, it would be about 10 percent larger.

TABLE EL-3
MATERIAL BALANCE

CASE EALC AND SUBCASE EALC-LP

FEEDS

	T (°F)	LB/HR	LB MOL/HR
Coal	77		
Moisture		35,000	1,942.8
Ash		80,000	
MAF Coal			
Carbon		554,985	46,205.9
Hydrogen		42,525	21,094.6
Oxygen		80,022	2,500.8
Nitrogen		9,985	356.4
Sulfur		30,816	961.1
TOTAL COAL		833,333	
Recycle Char	162		
Ash		54,771	
Carbon		169,400	14,103.6
TOTAL RECYCLE CHAR		224,171	
Oxidant (1)	437.4		
Oxygen		813,340	25,417.8
Nitrogen		2,678,679	95,612.4
Moisture		56,599	3,141.6
TOTAL OXIDANT		3,548,618	124,171.8
TOTAL FEEDS		4,606,122	

EFFLUENTS

	T (°F)	LB/HR	LB MOL/HR	MOL % (WET)
Gasifier Effluent	1700			
CH_4		10	0.8	0.00
H_2		38,044	18,872.0	11.22
CO		1,047,623	37,401.2	22.24
CO_2		376,463	8,554.0	5.08
H_2O		116,180	6,448.7	3.85
H_2S		30,801	904.0	0.54
COS		3,457	57.6	0.03
N_2		2,687,168	95,915.6	57.04
NH_3		28	1.6	0.00
TOTAL GASIFIER EFFLUENT		4,299,774	168,155.5	100.00
Entrained Char	1700			
Ash		54,771		
Carbon		169,400	14,103.6	
TOTAL ENTRAINED CHAR		224,171		
Slag	3000			
Ash		80,000		
Carbon		2,177	181.2	
TOTAL SLAG		82,177		
TOTAL EFFLUENTS		4,606,122		

Note:

(1) Oxidant stream temperature is an average of 600°F process air and 325°F transport air.

Process Energy Balances

Tables EL-4 and EL-4A present overall plant energy balances for Case EALC and Subcase EALC-LP at 100 percent capacity operation. The boundary for each balance encompasses the entire plant. Energy content of streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kWh. These energy balances close to less than one percent. The discrepancies result from approximations used for some process units and for calculating some heat loads. In both cases, product gas compressor steam rate data used by Westinghouse for the power block analysis have been taken as a basis. The product gas compressor power requirements were provided by Fluor for both cases.

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

As shown in the Table EL-5, Case EALC and Subcase EALC-LP, respectively, produce 3885 and 3888 MM Btu/hr based on the higher heating value (HHV) of coal. The system efficiency (net power at 3,413 Btu/kWh as a percent of coal HHV) is 38.10 in Case EALC and 38.13 in Subcase EALC-LP. The heat rates based on net power produced and coal HHV are 8959 Btu/kWh and 8,951 Btu/kWh in Case EALC and Subcase EALC-LP respectively.

In both the cases superheated intermediate pressure (IP) steam at 385 psig and 1000°F produced in the power block is used to drive the fuel gas compressor (25-1-C-1). Because the discharge pressure 25-1-C-1 in Subcase EALC-LP is significantly lower than in Case EALC (215 psia in Case EALC-LP and 315 psia in Case EALC), higher power generation is obtained from the steam turbine/generator in Subcase EALC-LP. However, due to the lower pressure of the fuel gas, less power is produced by the gas turbine generators. The total net power generated in Subcase EALC-LP is, therefore, only marginally higher than Case EALC.

The major heat losses from the system occur in the stack gases and at the surface condensers. As the total flue gas mass flow in Subcase EALC-LP is smaller than Case EALC because of lower combustion air requirement, the HRSG stack gas loss as a percent of coal HHV in Case EALC-LP is lower than the one in Case EALC (13.98 percent in Subcase EALC-LP, 15.11 percent in Case EALC). The heat losses at the

TABLE EL-4
ENERGY BALANCE - CASE EALC

BASIS: 60°F, Water as Liquid, 3413 Btu/kWh.

			MM Btu/hr			
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Blower Air		24	47			71
Gas Turbine Air		90	218			308
Makeup Water		2				2
Auxiliary Power Inputs					189	189
	10,196	121	265	0	189	10,771
HEAT OUT						
Ash/Slag	31	107				138
Gasifier Losses				153		153
Sulfur Product	102	1				103
Gas Turbine Power					3024	3024
Steam Turbine Power					1050	1050
Power Block Losses (1)				40	114	154
Steam Turbine Condenser			1452			1452
Compressor Turbine Condenser			2053			2053
Fuel Gas Compressor Coolers		487	332			819
Stretford Cooling			179			179
HRSG Stack Losses		930	611			1541
Waste Water Effluent		16				16
Steam Heat Losses		14	4			18
	133	1555	4631	193	4188	10,700

$$\frac{\text{Input - Output}}{\text{Input}} = 0.65\%$$

Note:

(1) Includes electrical and mechanical losses.

TABLE EL-4A

ENERGY BALANCE - SUBCASE EALC-LP

BASIS: 60°F, Water as Liquid, 3413 Btu/kWh.

			MM Btu/hr			
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Blower Air		24	47			71
Gas Turbine Air		81	196			277
Makeup Water		2				2
Auxiliary Power Inputs					189	189
	10,196	112	243		189	10,740
HEAT OUT						
Ash/Slag	31	107				138
Gasifier Losses				153		153
Sulfur Product	102	1				103
Gas Turbine Power					2635	2635
Steam Turbine Power					1442	1442
Power Block Losses (1)				45	111	156
Steam Turbine Condenser			2001			2001
Compressor Turbine Condenser			1714			1714
Compressor Coolers		459	330			789
Stretford Cooling			179			179
Flue Gases		860	565			1425
Waste Water Effluent		16				16
Steam Heat Losses		14	4			18
	133	1457	4793	198	4188	10,769

$$\frac{\text{Output - Input}}{\text{Output}} = 0.27\%$$

Note:

(1) Includes mechanical and electrical losses.

TABLE EL-5
ENERGY BALANCES AS PERCENT OF COAL HHV

	CASE EALC		SUBCASE EALC-LP	
	MMBTU/HR	PERCENT	MMBTU/HR	PERCENT
IN				
Coal HHV	10,196	100.00	10,196	100.00
OUT				
Net Power	3,885	38.10	3,888	38.13
Power Block Losses	154	1.51	156	1.53
Sulfur Product	103	1.01	103	1.01
Stretford Sensible & Latent	179	1.76	179	1.76
Fuel Gas Compression Cooling	819	8.03	789	7.73
Ash/Slag	138	1.35	138	1.35
HRSG Stack Gases	1,541	15.11	1,425	13.98
Surface Condensers	3,505	34.38	3,715	36.44
Gasifier Heat Losses	153	1.50	153	1.50
Other Sensible Losses	(91)	(0.89)	(82)	(0.80)
Other Latent Losses	(261)	(2.56)	(239)	(2.34)
	10,125	99.30	10,225	100.29

surface condensers based on percent coal HHV input are 34.38 and 36.44 percent respectively in Case EALC and Subcase EALC-LP. This occurs due to the larger power generation in the steam turbine generator in Subcase EALC-LP. Also, losses for fuel gas compression amount to 819 MM Btu/hr or 8.03 percent of coal HHV input in Case EALC and 789 MM Btu/hr (7.73 percent of coal HHV) in Subcase EALC-LP.

The acid gas removal unit accounts for losses of approximatley 1.76 percent of coal HHV input in both cases.

ECONOMICS

Important economic results are summarized below.

TABLE EL-6

SUMMARY OF ECONOMICS - CASE EALC

PRODUCTION AT DESIGN CAPACITY	
Net Power, MW (1)	1,138.1
Overall Plant Heat Rate, Btu/kWh	8,959
TOTAL CAPITAL (2)	
Total Capital @ $1/MM Btu Coal, $1000	1,059,156
Total Capital @$1/MM Btu, $/kW	931
Total Capital @ $2/MM Btu Coal, $1000	1,076,274
Total Capital @ $2/MM Btu, $/kW	946
AVERAGE COST OF SERVICES (2)	
Annual Cost @ $1/MM Btu Coal, $1000/yr	288,563
Per Unit @ $1/MM Btu Coal, mills/kWh	41.35
Annual Cost @ $2/MM Btu Coal, $1000/yr	353,756
Per Unit @ $2/MM Btu Coal, mills/kWh	50.69

NOTES:

(1) At 100% Operating Load Factor.
(2) Mid-1976 Dollars and 70% Operating Load Factor.

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The tables and discussion given below apply to Case EALC. Near the end of this study an alternate case, called Subcase EALC-LP, was briefly investigated. This case involved a lower compression ratio turbine, thus a less expensive fuel gas compressor. This alternate case was not fully evaluated, but appears to have somewhat more favorable economics than Case EALC. A discussion is included at the end of this section. Subcase EALC-LP is described in the Process Discussion Section.

Tables EL-7 and EL-8 give detailed breakdowns of plant investment, capital charges and working capital at 70% operating load factor for the two coal costs. Plant investment is the same at both coal costs. Capital charges are about 1 percent higher for the $2.00/MM Btu coal and working capital nearly 30 percent higher.

The accuracy of plant investment estimates is judged to be ±25%. Since other capital charges and working capital are keyed to elements of plant investment, this accuracy is reflected in other capital charges as well. Therefore, caution must be exercised in comparing Case EALC with cases representing other gasification technologies.

Major elements of plant investment are Fuel Gas Compression and the Combined Cycle System. Together they represent nearly 66 percent of the total plant investment.

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

Table EL-9 summarizes cost of services based on coal charged at $1.00/MM Btu and $2.00/MM Btu HHV. Costs are computed in accordance with criteria given by EPRI (Criteria section). They are presented as averages for the plant.

Operating labor requirement is a function of the number of units and trains. Requirements on a per shift basis are:

	CASE EALC
"A" Operators	5
"B" Operators	16
Foremen	2
Lab and Instrument Technicians	4

Catalyst and chemical costs are primarily for chemicals consumed in the demineralizer, cooling tower, and boiler feedwater treating. There are some minor costs associated with making up solution losses in the Stretford Sulfur Recovery Unit. Utility costs are for raw water makeup to the plant. The plant is otherwise self-sufficient in utilities.

Operating charges constitute about 42 percent of cost of services with coal at $1.00/MM Btu and about 52 percent at a coal cost of $2.00/MM Btu. Coal is the largest single operating charge, representing nearly 53 percent with coal at $1.00/MM Btu and almost 69 percent at the higher coal cost. The relationships as percentages are summarized below:

	CASE EALC	
Coal Cost, $/MM Btu, HHV	1.00	2.00
Coal as % of Operating Charges	50.7	67.3
Coal as % of Total Cost of Services	21.7	35.3
Operating Charges as % of Total Cost of Services	42.7	52.5
Capital Charges as % of Total Cost of Services	57.3	47.5

TABLE EL-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR
AND $1.00/MM BTU COAL

CASE EALC

	$1000 (1)	$/kW (2)	Percent
PLANT INVESTMENT			
Coal Handling	47,118	41.40	6.88
Oxidant Feed	2,111	1.85	0.31
Gasification and Ash Handling	89,483	78.63	13.06
Gas Cooling	14,025	12.32	2.05
Acid Gas Removal and Sulfur Recovery	24,360	21.40	3.55
Fuel Gas Compression	175,235	153.97	25.58
Steam, Condensate and BFW	845	0.74	0.12
Support Facilities	55,904	49.13	8.16
Combined Cycle	276,052	242.56	40.29
Subtotal	685,133	602.00	100.00
Contingency	143,259	125.88	
TOTAL PLANT INVESTMENT	828,392	727.88	
ILLINOIS SALES TAX	18,694	16.43	
CAPITAL CHARGES			
Preproduction Costs	52,072	45.75	
Paid-up Royalties	4,142	3.64	
Initial Catalyst and Chemical Charges	528	0.47	
Construction Loan Interest	103,466	90.91	
TOTAL CAPITAL CHARGES	160,208	140.77	
DEPRECIABLE CAPITAL	1,007,294	885.08	
WORKING CAPITAL	51,862	45.57	
TOTAL CAPITAL	1,059,156	930.65	

NOTES
(1) Mid-1976 dollars.
(2) Based on 100% Operating Load Factor.

TABLE EL-8

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR
AND $2.00/MM BTU COAL

CASE EALC

	$1000 (1)	$/kW (2)	Percent
PLANT INVESTMENT			
Coal Handling	47,118	41.40	6.88
Oxidant Feed	2,111	1.85	0.31
Gasification and Ash Handling	89,483	78.63	13.06
Gas Cooling	14,025	12.32	2.05
Acid Gas Removal & Sulfur Recovery	24,360	21.40	3.55
Fuel Gas Compression	175,235	153.97	25.58
Steam, Condensate and BFW	845	0.74	0.12
Support Facilities	55,904	49.13	8.16
Combined Cycle	276,052	242.56	40.29
Subtotal	685,133	602.00	100.00
Contingency	143,259	125.88	
TOTAL PLANT INVESTMENT	828,392	727.88	
ILLINOIS SALES TAX	18,694	16.43	
CAPITAL CHARGES			
Preproduction Costs	53,932	47.39	
Paid-up Royalties	4,142	3.64	
Initial Catalyst and Chemical Charges	528	0.47	
Construction Loan Interest	103,466	90.91	
TOTAL CAPITAL CHARGES	162,068	142.41	
DEPRECIABLE CAPITAL	1,009,154	886.72	
WORKING CAPITAL	67,120	58.98	
TOTAL CAPITAL	1,076,274	945.70	

NOTES

(1) Mid-1976 dollars.

(2) Based on 100% Operating Load Factor.

TABLE EL-9

COST OF SERVICES AT 70% OPERATING LOAD FACTOR

CASE EALC

COAL COST, HHV	$1.00/MM Btu	$2.00/MM Btu
NET PRODUCTION (1)		
Net Power, MW	1,138.1	1,138.1
By-product Ammonia ST/SD	0	0
By-product Sulfur ST/SD	309	309
OPERATING CHARGES, $1000/YR		
Coal	62,522	125,044
Operating Labor	2,596	2,596
Catalyst and Chemicals	334	334
Utilities	1,272	1,272
Maintenance Labor	9,798	9,798
Maintenance Materials	14,697	14,697
Administration and Support Labor	3,718	3,718
General and Administration Expenses	7,436	7,436
Ash Disposal	252	252
Property Tax and Insurance	20,710	20,710
By-product Ammonia	(0)	(0)
By-product Sulfur	(0)	(0)
TOTAL OPERATING CHARGES, $1000/yr	123,335	185,857
CAPITAL CHARGES, $1000/YR		
Total Capital Charges	165,228	167,899
COST OF SERVICES		
Total, $1000/yr	288,563	353,756
Per Unit Production, mills/kWh	41.35	50.69

NOTES

(1) At 100% Operating Load Factor.

Subcase EALC-LP was investigated briefly to determine the value of selecting a low pressure ratio gas turbine in the Combined Cycle System. This turbine selection reduced the fuel gas supply pressure requirement for the gas turbine, resulting in less expensive fuel gas compressors. No plant investment estimate was prepared for Subcase EALC-LP, but a comparison of the major equipment differences is shown below:

	Case EALC	Subcase EALC-LP
Fuel Gas Compressor		
Number Required	4	4
Total Compressor Cost	$44,200,000	$36,584,000
Combined Cycle Power Block	$136,646,000	$131,851,000
Total	180,846,000	168,435,000

For both cases, net power production is essentially the same. Differences in unit costs, other than the fuel gas compressors and Combined Cycle System as shown above, should be minor. Referring to Tables EL-7 and EL-8, Fuel Gas Compressor and the Combined Cycle System represent nearly 66 percent of the total plant investment.

A reduction in equipment cost in these two areas of 6.8 percent for Subcase EALC-LP should translate to a significant savings in overall plant installed cost of approximately 7.4 percent which translates into a savings of approximately $68/kW for the Total Capital requirement resulting in an approximate reduction in the cost of electricity of 2 mills/kWh. Based on these comparisons, subcase EALC-LP appears to be the more favorable of the two cases.

CASE EXTC AND EXTC-DF - TEXACO

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PLANT DESCRIPTION - CASE EXTC

GENERAL

Grass roots plants for electric power generation based on single stage, entrained bed, oxygen blown gasifiers of the Texaco type integrated with combined cycle generating equipment, are shown schematically on the block flow diagrams EXTC-1-1 and EXTC-1-2. These plants consume 10,000 ST/day of Illinois No. 6 coal. The first diagram (Flow Diagram: EXTC-1-1) represents a case where coal is fed to the gasifiers as a water slurry containing approximately 66.5 wt % coal. This case is identified as Case EXTC (Slurry Feed).

The second diagram (Flow Diagram: EXTC-1-2) represents a subcase based on feeding dry pulverized coal and process steam to the gasifiers. This case is referred to as Case EXTC (Dry Feed). At present, Texaco does not offer nor is actively developing a mechanism for feeding dry coal to the gasifier. The purpose of studying this subcase was to try to identify the economic incentives (if any) for the development of a dry coal feeding mechanism.

The main plant consists of oxidant feed, gasification, gas cooling, acid gas removal units and combined cycle power systems. The oxidant feed unit is in five parallel operating trains. The gasification unit consists of five parallel operating trains and one spare train. The gas cooling and acid gas removal units are in three operating parallel trains. There are seven parallel gas turbine, heat recovery steam generator sets and a single steam turbine.

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

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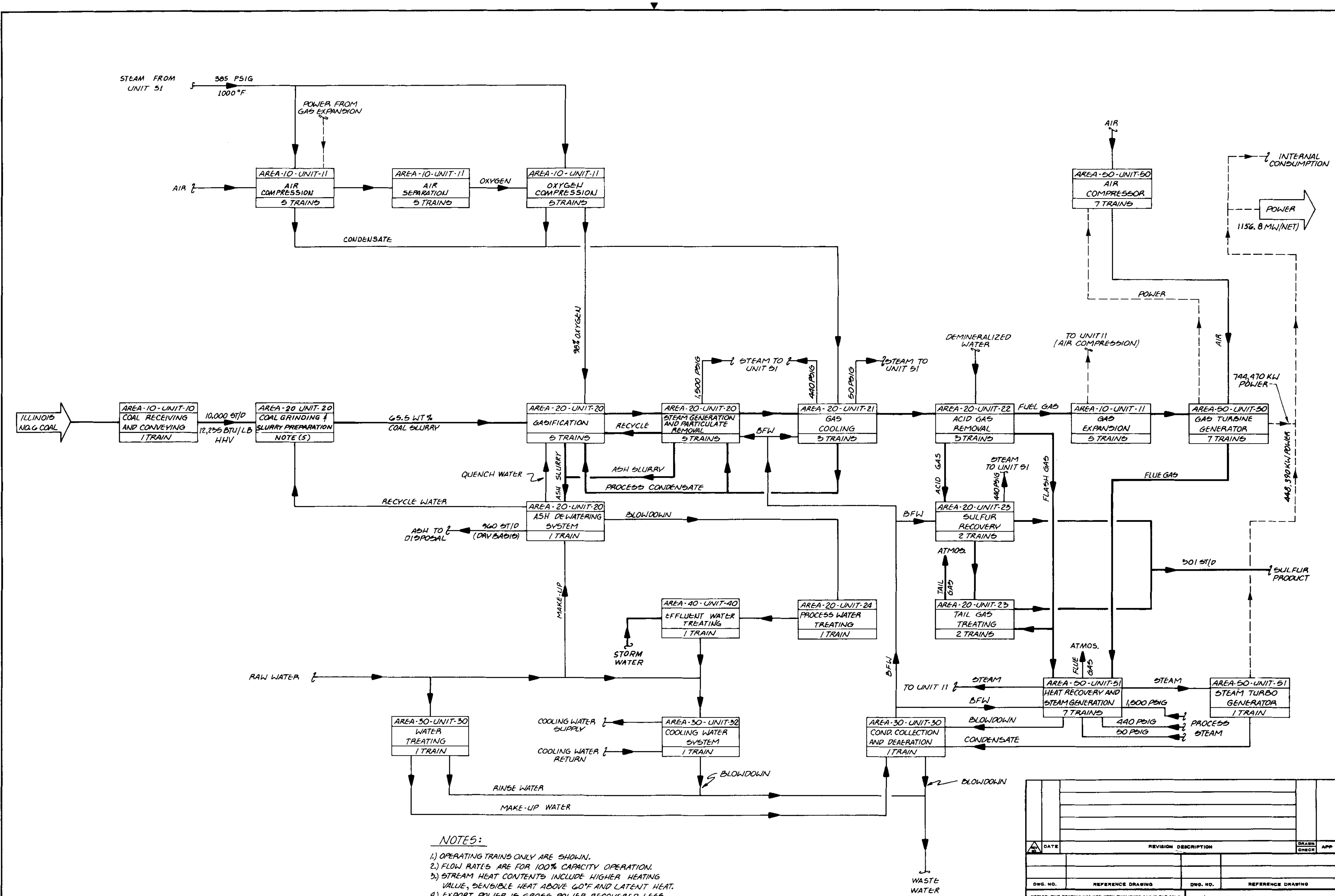

NOTES:

1.) OPERATING TRAINS ONLY ARE SHOWN.
2.) FLOW RATES ARE FOR 100% CAPACITY OPERATION.
3.) STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE, SENSIBLE HEAT ABOVE 60°F AND LATENT HEAT.
4.) EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
5.) TWO TRAINS OF COAL GRINDING & ONE TRAIN OF SLURRY PREPARATION ARE PROVIDED.

FLUOR

DRAWN BY B. GEORGE / R. KRESS
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
FEERRAR

OVERALL BLOCK FLOW DIAGRAM
TEXACO PROCESS COAL GASIFICATION
OXYGEN BLOWN - SLURRY FEED
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE NONE — DRAWING NUMBER EXTC-1-1 — REVISION 02

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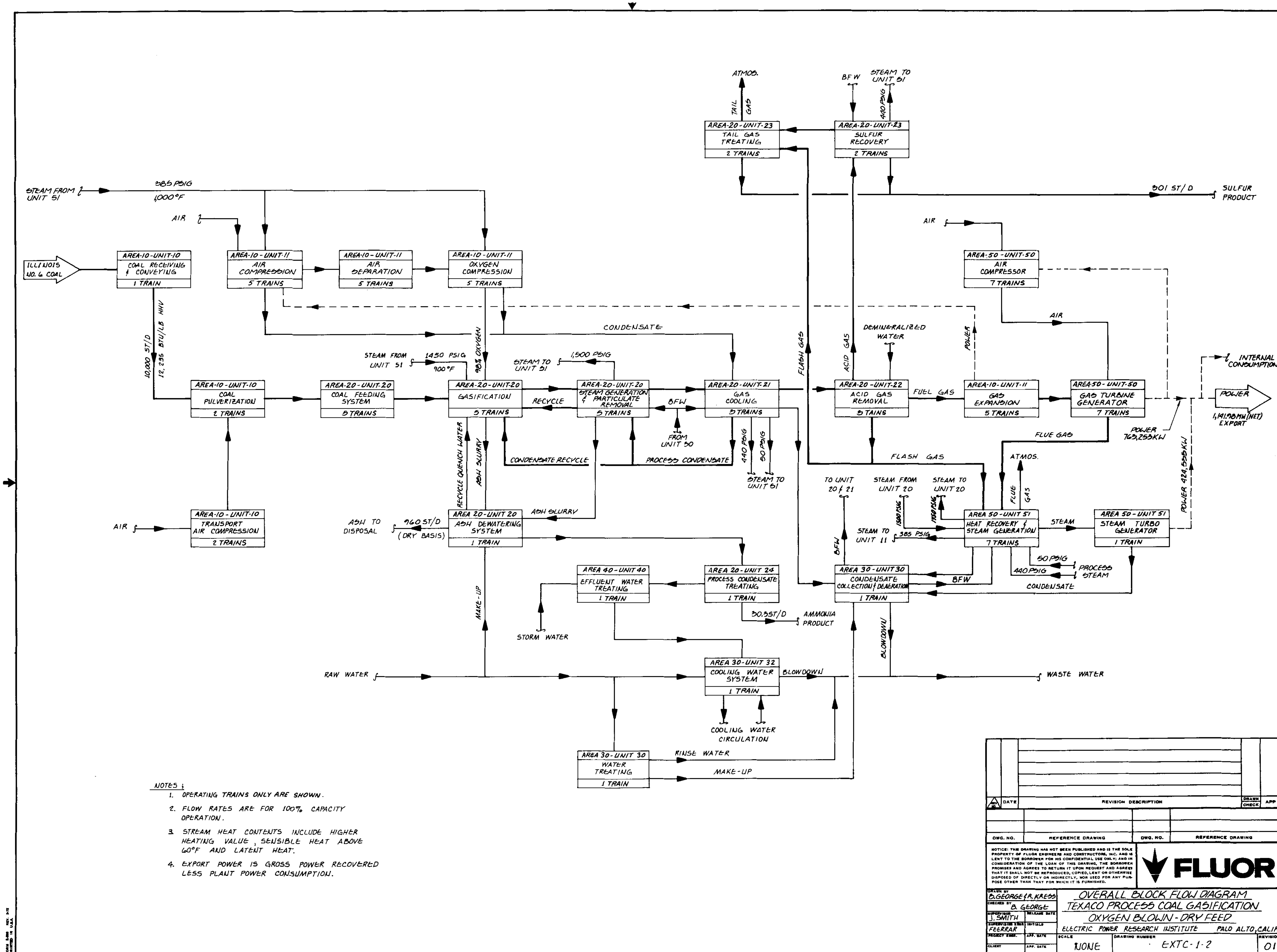

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TABLE ET-1

MAJOR EQUIPMENT SECTIONS: CASE EXTC

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

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COAL PREPARATION

Process Flow Diagram EXTC-10-1 depicts the process arrangement of equipment in this section for Case EXTC (Slurry Feed). There is one 100% train of coal unloading, stacking, reclamation and conveying equipment.

The coal preparation area for Subcase EXTC (Dry Feed) is similar to the one described for Case EAHC. Refer to process description of Coal Preparation Unit, Case EAHC and Process Flow Diagram EAHC-10-1. There is one 100% train of coal unloading, stacking, reclamation and conveying equipment. The coal pulverization system is in two 50% parallel trains.

Case EXTC (Slurry Feed)

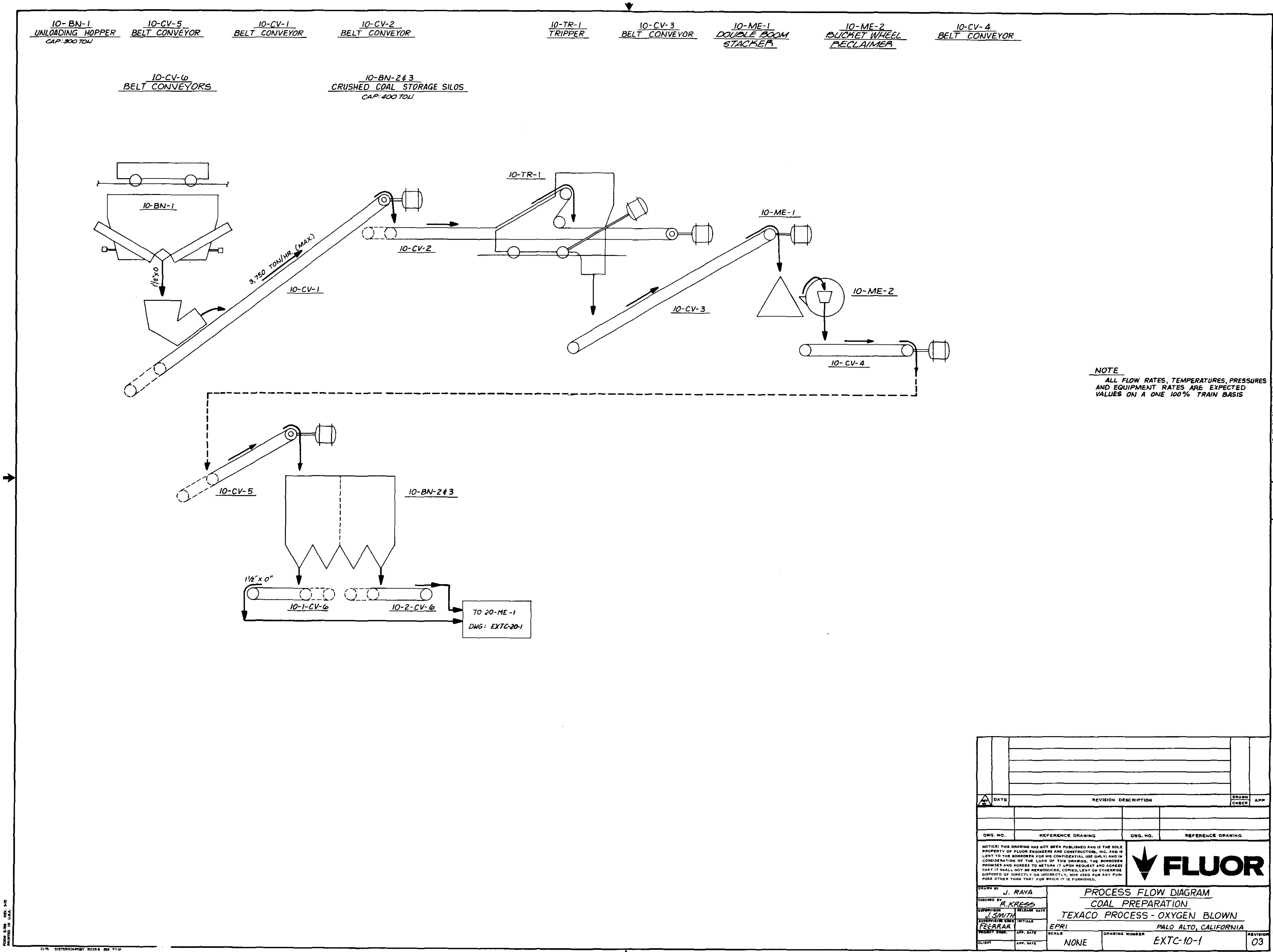
Washed, 1-1/2" x 0 coal is received at the plant site by unit train. The coal is unloaded from 100-ton bottom dump cars into an unloading hopper, withdrawn from the hopper by two vibrating feeders and transported by belt conveyors to a tripper. The tripper distributes coal to a traveling belt stacking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

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

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

Equipment Notes

All the equipment is commercially available.


10-BN-1
UNLOADING HOPPER
CAP: 300 TON
10-CV-5
BELT CONVEYOR
10-CV-1
BELT CONVEYOR
10-CV-2
BELT CONVEYOR
10-TR-1
TRIPPER
10-CV-3
BELT CONVEYOR
10-ME-1
DOUBLE BOOM STACKER
10-ME-2
BUCKET WHEEL RECLAIMER
10-CV-4
BELT CONVEYOR
10-CV-6
BELT CONVEYORS
10-BN-2 & 3
CRUSHED COAL STORAGE SILOS
CAP: 400 TON
1½x0
3,750 TON/HR (MAX)
1½" x 0"
10-1-CV-6
10-2-CV-6
TO 20-ME-1
DWG: EXTC-20-1
NOTE
ALL FLOW RATES, TEMPERATURES, PRESSURES AND EQUIPMENT RATES ARE EXPECTED VALUES ON A ONE 100% TRAIN BASIS
REDUCED PRINT SCALES
DATE
REVISION DESCRIPTION
DRAWN
CHECK
APP
DWG. NO.
REFERENCE DRAWING
FLUOR
DRAWN BY J. RAYA
CHECKED BY R. KRESS
J. SMITH
FERRAR
PROCESS FLOW DIAGRAM
COAL PREPARATION
TEXACO PROCESS - OXYGEN BLOWN
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EXTC-10-1
REVISION 03
305
EXTC-10-1

OXIDANT FEED

Process Flow Diagram EXTC-11-1 shows the oxidant feed system for Case EXTC (Slurry Feed). The process scheme for the oxidant feed system for Case EXTC (Dry Feed) is similar to Case EXTC (Slurry Feed). The capacity of this system for the dry feed case is approximately 94% of that for the Slurry Feed Case.

There are five parallel operating trains. Each train has one air compressor, one air separation plant and one oxygen compressor. No spare train is provided in this section.

Atmospheric air in Case EXTC is compressed to 96 psig in a two stage centrifugal machine, 11-1-C-1. Heat of compression is rejected to air in interstage airfan cooler (11-1-E-2).

The 34,100 hp required by each air feed compressor is supplied by combination of a steam turbine, 11-1-T-1 and a fuel gas expander, 11-1-EX-1. The steam turbine driver is a condensing type machine operating at inlet conditions of 385 psig, 1000°F, with exhaust pressure at 2-1/2" Hg abs. to meet the overall steam balance requirements. The steam turbine is designed with excess capacity to provide response capabilities during turndown or upset conditions. Each of the five operating fuel gas expanders extracts approximately 7280 bhp from the fuel gas, which has been preheated to 950°F in the heat recovery steam generators (HRSGs) located in Unit 51. The fuel gas is expanded from 495 psig, 950°F to approximately 280 psig, 781°F and flows back to the gas turbines.

The compressed air is processed in an air separation unit which produces 1676 tons per day (100% basis) of 98% oxygen. Liquid oxygen storage of approximately 5550 tons is provided, with attendant cryogenic pumps and vaporizer. Storage is equivalent to approximately three days of rated capacity operation of a single train. The three days of storage is anticipated to adequately cover any outage of the cryogenic unit.

The air separation plant produces oxygen at 2 psig and 90°F. The oxygen is compressed to 400 psig in six stages. As in the case of the air feed compressor, interstage heat of compression is rejected to Interstage Air Fan Coolers 11-1-E-5 through 9.

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11-1-E-2
AIR INTERCOOLER
38 MM BTU/HR
16,652 SQ. FT. BARE
TUBES: CARBON STL.

11-1-E-3
AIR AFTERCOOLER
39 MM BTU/HR
10,937 SQ FT. BARE
TUBES: CARBON STL.

11-1-C-1
AIR
COMPRESSOR
34,100 BHP

11-1-E-1
SURFACE
CONDENSER
140 MM BTU/HR
24,500 SQ. FT.
TUBES : CARBON STEEL
SHELL : CARBON STEEL

11-1-E-4
TRIM COOLER
13.0 MM BTU/HR
8,670 SQ. FT
TUBES: CARBON STL.
SHELL: CARBON STL.

11-1-V-1
K.O. DRUM
120" I.D.x 10'-6" T-T
CARBON STL.
DESIGN: 105 PSIG@150°F

11-1-ME-1
AIR SEPARATION
UNIT
CAP: 1676 TPD NORMAL
1850 TPD DESIGN

11-1-C-2
OXYGEN
COMPRESSOR
11,700 BHP

11-1-E-5
1ST STAGE
O_2 COOLER
4.7 MM BTU/HR
3,489 SQ. FT. BARE
TUBES: 316 SS

11-1-E-6
2ND STAGE
O_2 COOLER
4.8 MM BTU/HR
2,632 SQ. FT. BARE
TUBES: 316 SS

11-1-E-7
3RD STAGE
O_2 COOLER
4.8 MM BTU/HR
1,999 SQ. FT. BARE
TUBES: 316 SS

11-1-E-8
4TH STAGE
O_2 COOLER
4.7 MM BTU/HR
1,475 SQ. FT. BARE
TUBES: 316 SS

11-1-E-9
5TH STAGE
O_2 COOLER
4.4 MM BTU/HR
1,126 SQ. FT. BARE
TUBES: 316 SS

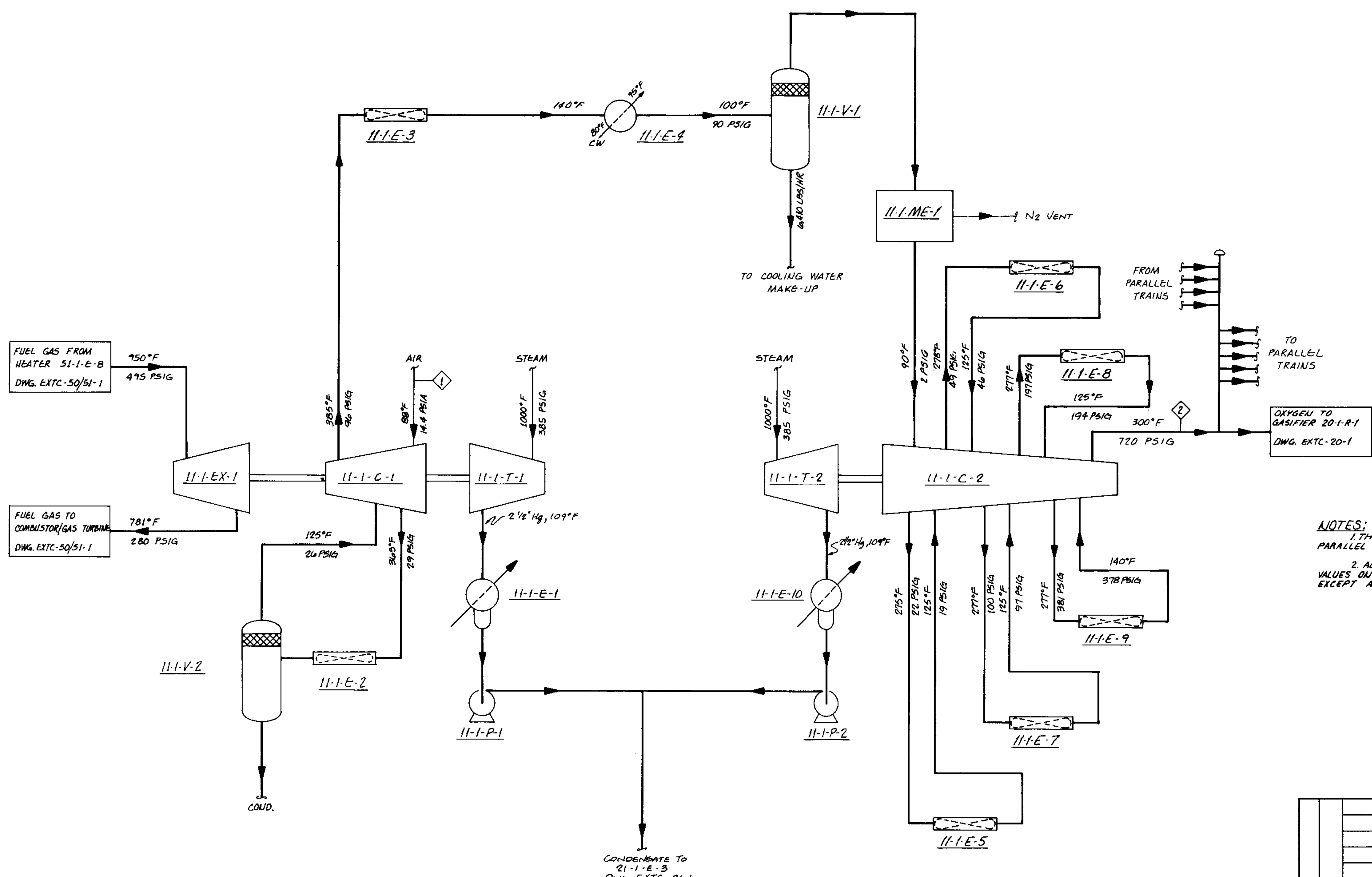
11-1-EX-1
FUEL GAS
EXPANDER
7280 BHP

11-1-V-2
K.O. DRUM
162" I.D.x 13'-0" T/T
DESIGN: 50 PSIG @ 175°F
CARBON STEEL

11-1-T-1
AIR COMP TURBINE
26,820 BHP

11-1-E-10
SURFACE CONDENSER
61 MMBTU/HR
10,680 SQ. FT.
TUBES: CARBON STEEL
SHELL : CARBON STEEL

11-1-T-2
OXYGEN COMP TURBINE
11,700 BHP



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

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

MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER: 1 AIR MPH	MOL%	2 OXIDANT MPH	MOL%		
O_2	4626.4	20.48	4272.5	98.00		
N_2	17,184.2	76.07	66.0	1.50		
AR	221.4	0.98	21.7	0.50		
H_2O	558.0	2.47	—	—		
TOTAL MPH	22,590.0	100.00	4367.2	100.00		
LB/HR	648,310.		139,650			
MOL. WT.	28.70		31.98			

FORM E-388 REV. 9-72 PRINTED IN U.S.A.
11-75 DIETERICH-POST 2111X-6 .003 FILM

REV NO	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
TEXACO PROCESS-OXYGEN BLOWN
EPRI PALO ALTO, CALIFORNIA

DRAWN BY: P. STONE
CHECKED BY: J. RAYA
SUPERVISOR: J. SMITH
RELEASE DATE:
SUPERVISING ENGR: FEERRAR
INITIALS:
PROJECT ENGR.:
APP. DATE:
CLIENT:
APP. DATE:
SCALE: NONE
DRAWING NUMBER: EXTC-11-1
REVISION: 02

EXTC-11-1
AA8334

REDUCED PRINT SCALES

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The 11,700 hp compression requirement is supplied by a condensing type steam turbine. The inlet steam condition is 385 psig, 1000°F with back pressure at 2-1/2" Hg abs.

Equipment Notes

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

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GASIFICATION AND ASH HANDLING

Case EXTC (Slurry Feed)

Process Flow Diagram EXTC-20-1 shows the gasification step for Case EXTC (Slurry Feed). There are six parallel trains. One train is a spare. The ash handling system consists of one 100% capacity train. The boxes on EXTC-20-1 represent proprietary sections of the Texaco Coal Gasification Process. Each of these sections contains many units of equipment.

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

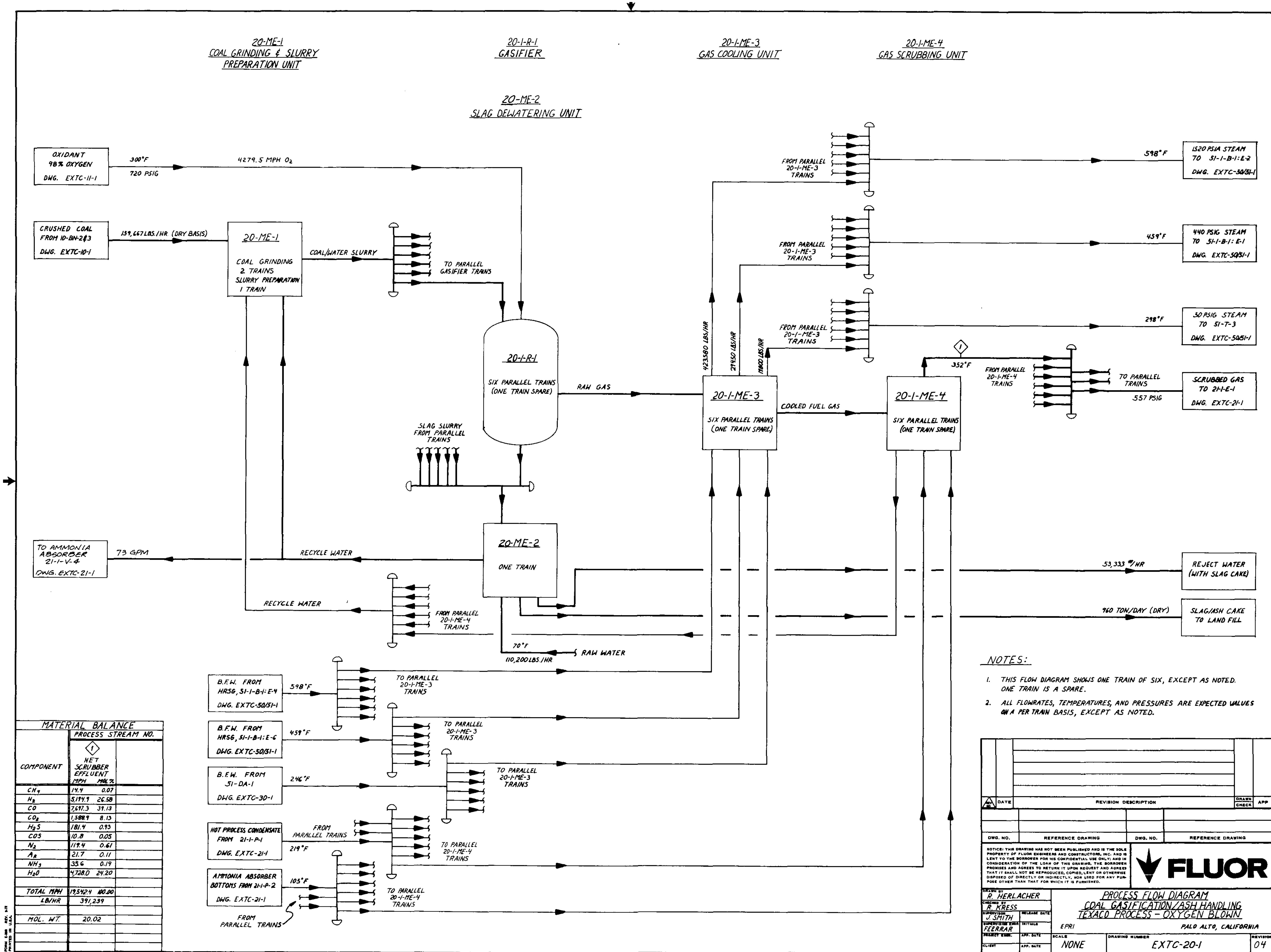
Coal slurry and oxygen combine at the gasifier burners. Each burner is oriented downward from the top head of the gasifier. The burners have circulating, tempered water cooling coils.

The gasification section, 20-1-R-1, operates at an average pressure of 600 psig and temperatures in the range of 2300°F to 2600°F. Part of the coal burns with oxygen to produce a hot flue gas. This reaction provides heat for the endothermic steam/carbon and CO/CO_2 reactions. The coal's hydrogen and carbon react to form CO, CO_2, H_2 and a small amount of CH_4. Most of the sulfur is converted to H_2S and COS. Nitrogen in the coal transforms to free nitrogen and a small quantity of ammonia. The ash melts to form slag. The gasification temperature must be sufficiently above the ash flow point to ensure free flowing molten slag. At the high temperatures prevailing in the gasifier, some of the ammonia in the recycled water is destroyed.

Most of the ash in the form of slag falls into a water quench at the bottom of the gasifier. The resultant ash slurry leaves the gasifier and enters the slag dewatering unit.

A slag/ash cake from dewatering unit 20-ME-2 is disposed to landfill. Overflow from the slag dewatering unit is recycled to the coal grinding system.

Raw hot gas from the gasifier is cooled in a gas cooling unit, 20-1-ME-3, to a temperature well below the ash softening point. This gas cooling system is of



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NO. 1: NET SCRUBBER EFFLUENT MPH	MOL %	
CH_4	14.4	0.07	
H_2	5,194.9	26.58	
CO	7,647.3	39.13	
CO_2	1,588.9	8.13	
H_2S	181.4	0.93	
COS	10.8	0.05	
N_2	119.4	0.61	
AR	21.7	0.11	
NH_3	35.6	0.19	
H_2O	4,728.0	24.20	
TOTAL MPH	19,542.4	100.00	
LB/HR	391,239		
MOL. WT.	20.02		

proprietary design and allows for the removal of ash entrained in the crude gas for the protection of downstream heat exchange equipment. Hot boiler feedwater at IP steam saturation temperature (459°F) is supplied from the heat recovery steam generation (HRSG) units located in Unit 51. Boiler feedwater is also supplied from deaerator 51-DA-1 located in Unit 51 for LP steam production. High pressure (HP) steam at 1520 psia, saturated intermediate pressure (IP) steam at 420 psig and saturated low pressure (LP) steam at 50 psig, are produced in Unit 20-1-ME-3.

The raw gas leaves 20-1-ME-3 and flows to the gas scrubbing unit, 20-1-ME-4. Ammonia absorber bottoms from the gas cooling area (Flow Diagram: EXTC-21-1) and hot process condensate are used for gas scrubbing. Water from 20-1-ME-4 is recycled to 20-ME-1.

The clean gases from 20-1-ME-4 flow to the gas cooling section.

Case EXTC (Dry Feed)

This conceptual case was developed by Fluor and EPRI. It uses a pulverized coal feeding system for the gasifiers similar to the one used for Case EAHC, with the exception that high pressure process steam at 1450 psig, 900°F is also used as transport gas for feeding the pulverized coal from 20-1-V-2 to the gasifiers. Refer to Process Flow Diagram EAHC-20-1 and process description for Case EAHC for the arrangement of the equipment 20-1-ME-1, 20-1-BN-1, 20-1-V-1A, B and 20-1-V-2. The gasification and ash handling system is similar to the one described for Case EXTC (Slurry Feed). One major difference between the slurry feed and dry coal feed cases is that in the dry coal feed case steam is injected through the burners with the oxygen and coal in order to moderate the temperature in the gasifier.

There are six parallel trains (five operating, one spare) for the coal feeding and the gasification step. The slag dewatering unit consists of one 100% capacity train.

Reclaimed water flows to the process condensate unit for ammonia recovery.

Equipment Notes

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

The Texaco coal gasification research facility at Montebello, California, is presently testing coals and chars in a 350 psia 15 ton/day gasifier. A 150 ton/day Texaco coal gasifier is scheduled to start up in Germany early in 1978.

The slag dewatering unit is commercially proven.

The gas scrubbing unit equipment is commercially available.

GAS COOLING

Case EXTC (Slurry Feed)

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

Clean gasifier effluent from the particulate scrubbing section is cooled to approximately 105°F in a series of exchangers, 21-1-E-1, 21-1-E-2 and 21-1-E-3. Heat is recovered in exchanger 21-1-E-1 by the generation of saturated 50 psig steam. The effluent, after separation of condensate in the knockout drum 21-1-V-2, is then cooled by exchanging heat against fuel gas from the acid gas removal section. The condensate produced in cooling is separated in 21-1-V-2. Further gas cooling is obtained in exchanger 21-1-E-3 by heating a condensate stream and the resultant condensate is separated in knockout drum 21-1-V-3. Condensate from knockout drums, 21-1-V-1 and 21-1-V-2 also flows to 21-1-V-3. Hot condensate from 21-1-V-3 is then pumped to the gasification unit (Flow Diagram: EXTC-20-1).

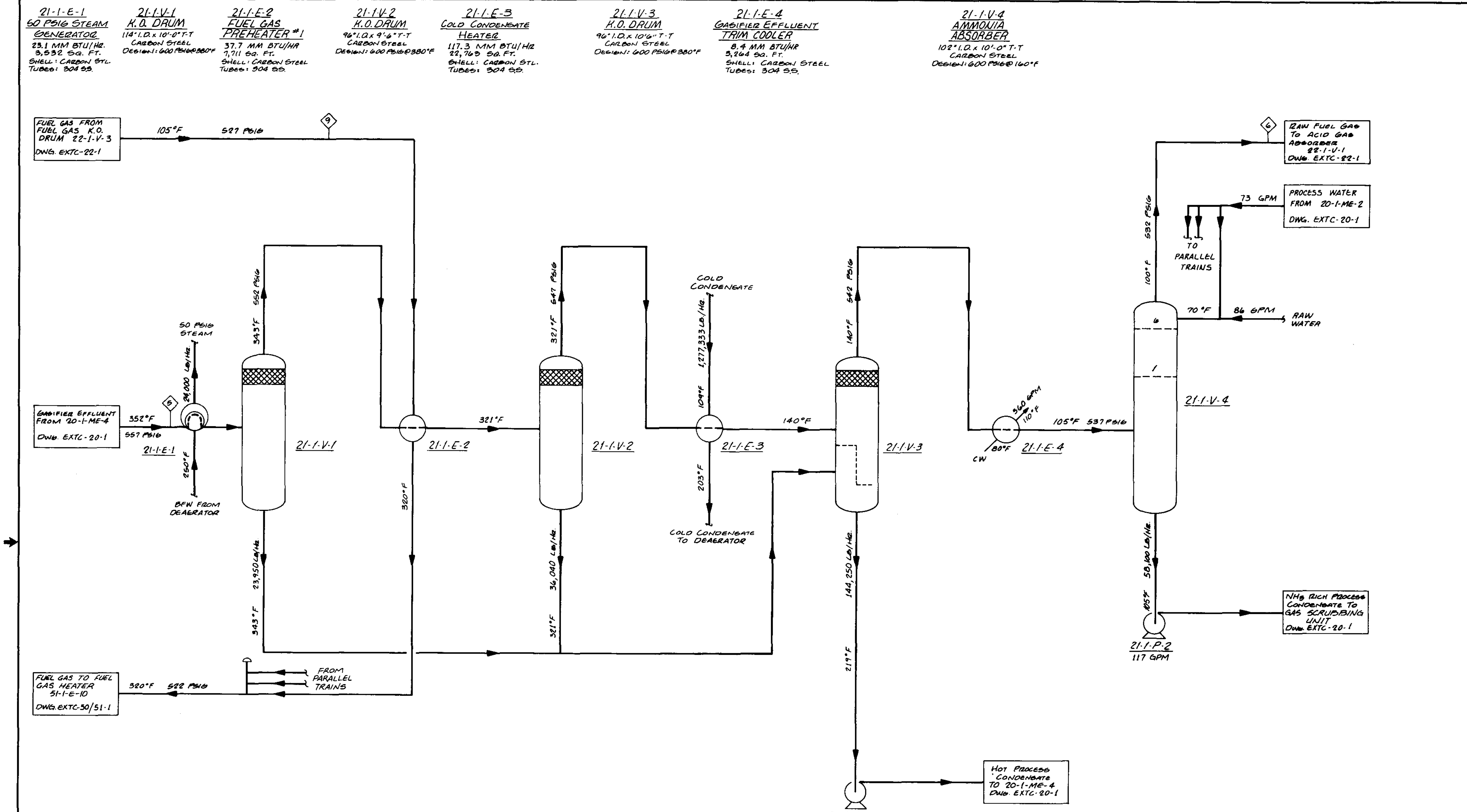
The overhead gases from knockout drum 21-1-V-3 are further cooled with water to approximately 105°F in a trim cooler, 21-1-E-4, and flow to an ammonia absorber, 21-1-V-4. Ammonia is removed by contacting the gas countercurrently with the water on the trays. The ammonia-free overhead gases from the absorber then flow to the acid gas removal unit for H_2S removal. The ammonia-rich process condensate from the bottom of the absorber is pumped to the particulate scrubber, 20-1-V-4.

Case EXTC (Dry Feed)

The basic process scheme for this case is similar to the one described for Case EXTC (Slurry Feed) with small differences in gas throughput. There are three parallel trains.

Equipment Notes

All equipment is commercially available.



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER: 5 PARTICULATE SCRUBBER OVHD MPH	5 MOL%	6 AMMONIA SCRUBBER OVHD MPH	6 MOL%	9 FUEL GAS MPH	9 MOL%
CH_4	24.0	0.07	24.0	0.10	23.6	0.10
H_2	8658.2	26.35	8658.2	35.19	8632.4	36.38
CO	12,745.5	38.78	12,745.5	51.81	12,662.5	53.36
CO_2	2,648.2	8.06	2648.2	10.77	2196.0	9.26
H_2S	303.0	0.92	302.3	1.23	24.0	0.10
COS	18.0	0.06	18.0	0.07	11.6	0.05
N_2	199.0	0.60	119.0	0.48	118.1	0.50
AR	36.2	0.11	36.2	0.14	36.1	0.15
NH_3	59.3	0.19	—	—	—	—
H_2O VAPOR	8,172.0	24.86	51.65	0.21	22.5	0.10
TOTAL VAPOR MPH	32,863.3	100.00	24,599.4	100.00	23,726.8	100.00
H_2O LIQ	—		—		—	
TOTAL LB/HR	658,100		508,039		475,570	
MOL WT VAPOR	20.02		20.65		20.04	

NOTES:

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

DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY: P. STONE
CHECKED BY: B. GEORGE
SUPERVISOR: J. SMITH
RELEASE DATE
SUPERVISING ENGR: FEERRAR
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
GAS COOLING
TEXACO PROCESS-OXYGEN BLOWN
EPRI — PALO ALTO, CALIFORNIA

SCALE: NONE
DRAWING NUMBER: EXTC-21-1
REVISION: 02

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ACID GAS REMOVAL

Case EXTC (Slurry Feed)

Process Flow Diagram EXTC-22-1 depicts one of the three parallel acid gas removal trains. No spare train is provided.

The acid gas removal system employs Allied Chemical Corporation's Selexol® process for selective removal of hydrogen sulfide (H_2S). Hydrogen sulfide in the crude gas is absorbed in Selexol® solvent in order to reduce sulfur in the treated gas to 1.0 pound sulfur dioxide (SO_2) equivalent per million Btu (HHV) coal charged to the plant.

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

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

The rich solvent solution from the flash drum exchanges heat with hot regenerated solution in 22-1-E-2 and flows to the top of the regenerator, 22-1-V-4. In the regenerator the absorbed H_2S and CO_2 are stripped from the solution. Reboil heat is supplied by 115 psig steam in a vertical thermosyphon reboiler, 22-1-E-3. Hot regenerated solvent is pumped back to Absorber 22-1-V-1 through exchangers 22-1-E-2 and 22-1-E-1. Heat is first exchanged with rich solution in 22-1-E-2 in order to reduce reboiler duty. The lean solution is cooled down to operating temperature with cooling water in exchanger 22-1-E-1.

Acid gas from the regenerator overhead is cooled to 120°F in air fan cooler 22-1-E-4. The condensate produced in cooling is separated in the knockout drum, 22-1-V-5, and then pumped back to the regenerator through 22-1-P-2. A small

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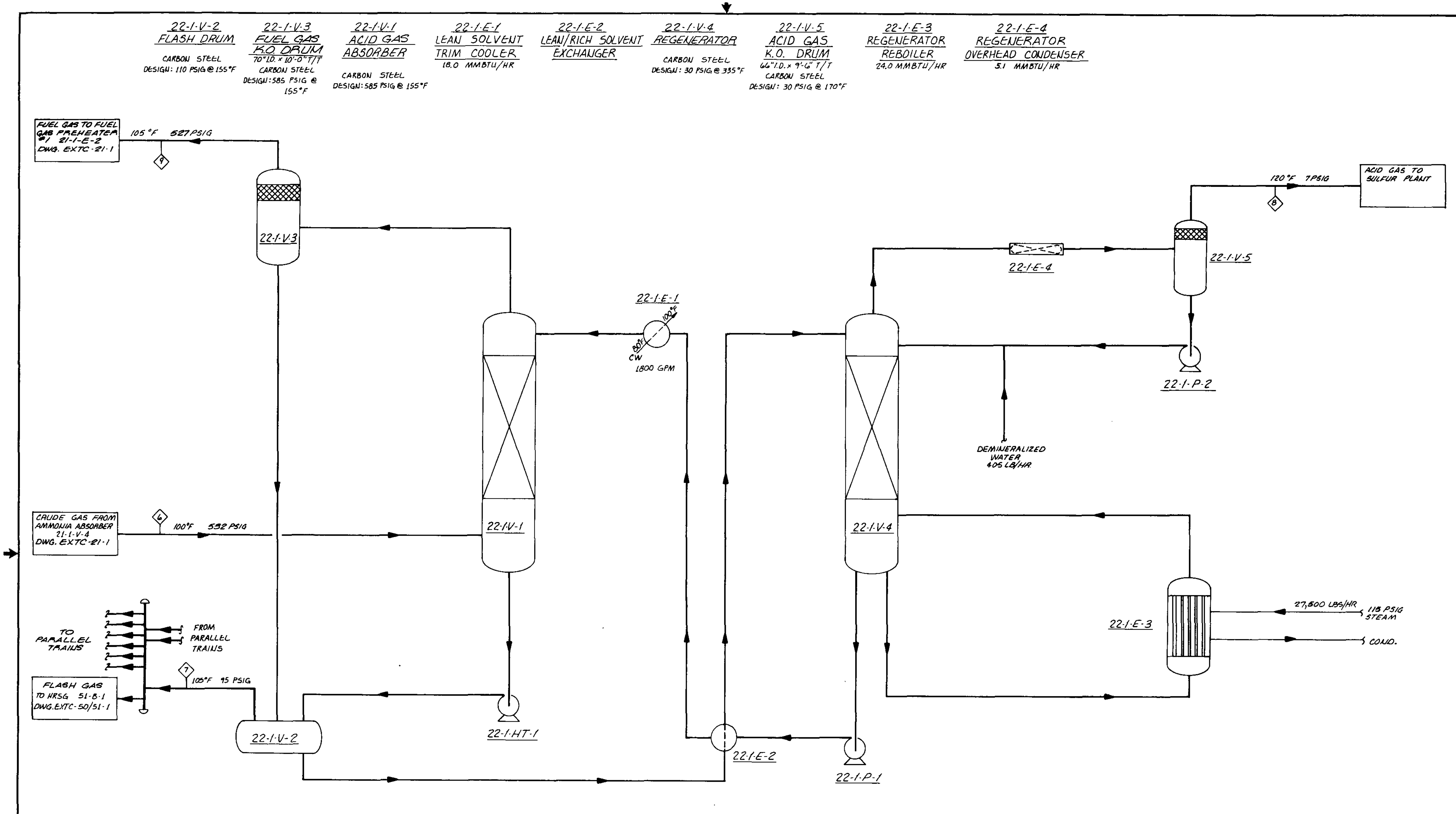

NOTES:

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

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

MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER							
	6 CRUDE GAS		7 FLASH GAS		8 ACID GAS		9 FUEL GAS	
	MPH	MOL%	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	24.0	0.10	0.3	0.16	0.1	0.01	23.6	0.10
H_2	8658.2	35.79	24.1	12.61	1.7	0.25	8632.4	36.38
CO	12,745.1	51.81	71.9	37.62	10.7	1.59	12,662.5	53.36
CO_2	2648.2	10.77	81.2	42.49	337.2	49.98	2196.0	9.26
H_2S	302.3	1.23	10.9	5.70	267.4	39.63	24.0	0.10
COS	18.0	0.07	0.6	0.31	5.8	0.86	11.6	0.05
N_2	119.0	0.48	0.8	0.42	0.1	0.01	118.1	0.50
AR	36.2	0.14	0.1	0.05	—	—	36.1	0.15
NH_3	—	—	—	—	—	—	—	—
H_2O	51.65	0.21	1.2	0.63	51.7	7.66	22.5	0.10
TOTAL MPH	24,599.4	100.00	191.1	100.00	674.7	100.00	23,726.8	100.00
TOTAL LB/HR	508,039		6096		26539		475,570	
MOL WT	20.65		31.90		39.33		20.04	

FORM E-306 REV. 9-72 PRINTED IN U.S.A.

11-75 DIETERICH-POST 3111X-8 .003 FILM

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

EPRI
PALO ALTO, CALIFORNIA

DRAWN BY P. STONE
CHECKED BY B. GEORGE
J. SMITH
FEERRAR

SCALE: NONE
DRAWING NUMBER: EXTC-22-1
REVISION: 03

FLUOR

EXTC-22-1

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stream of demineralized water is added to the condensate at the discharge of 22-1-P-2 to maintain the water balance in the absorption system. The cooled acid gas from 22-1-V-5 contains about 39.6% H_2S on a volume basis and flows to the sulfur recovery unit for further processing.

Case EXTC (Dry Feed)

The basic process scheme for this case is similar to the one described for Case EXTC (Slurry Feed) with small differences in gas throughput.

There are three parallel operating trains. No spare train is provided.

Equipment Notes

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

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SULFUR RECOVERY AND TAIL GAS TREATING

The processes used in these units for both Case EXTC (Slurry Feed) and Case EXTC (Dry Feed) are the same as for Case MACW. Refer to Case MACW and Process Flow Diagrams MACW-23-1, MACW-23-2 and MACW-23-3 for the detailed process descriptions of these units.

There are two 50% parallel operating sulfur recovery trains each followed by a tail gas treating unit. Sulfur recovery per train is 150.7 short tons per day for both cases. There is a third (spare) train because of the important environmental requirements these units fulfill.

PROCESS CONDENSATE TREATING

Case EXTC (Slurry Feed)

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

A small stream of ammonia (as ammonium salts) contaminated effluent leaving the process units is obtained in the gasification area (Flow Diagram: EXTC-20-1). This effluent is small and treated in the effluent water treating unit (Unit 40).

A unit for the recovery of byproduct ammonia is therefore not provided in Case EXTC (Slurry Feed).

Case EXTC (Dry Feed)

The process condensate treating unit employs U.S. Steel's proprietary PHOSAM-W process.

The PHOSAM-W is a widely accepted process to reclaim anhydrous ammonia from the sour process condensate. About 30.3 short tons per day of ammonia is recovered in this case.

The flash gas stream, containing H_2S and COS, from the process condensate unit is sent to tail gas unit (Unit 23) for further processing.

Equipment Notes

All equipment is commercially available.

STEAM, BOILER FEEDWATER AND CONDENSATE

Case EXTC (Slurry Feed)

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

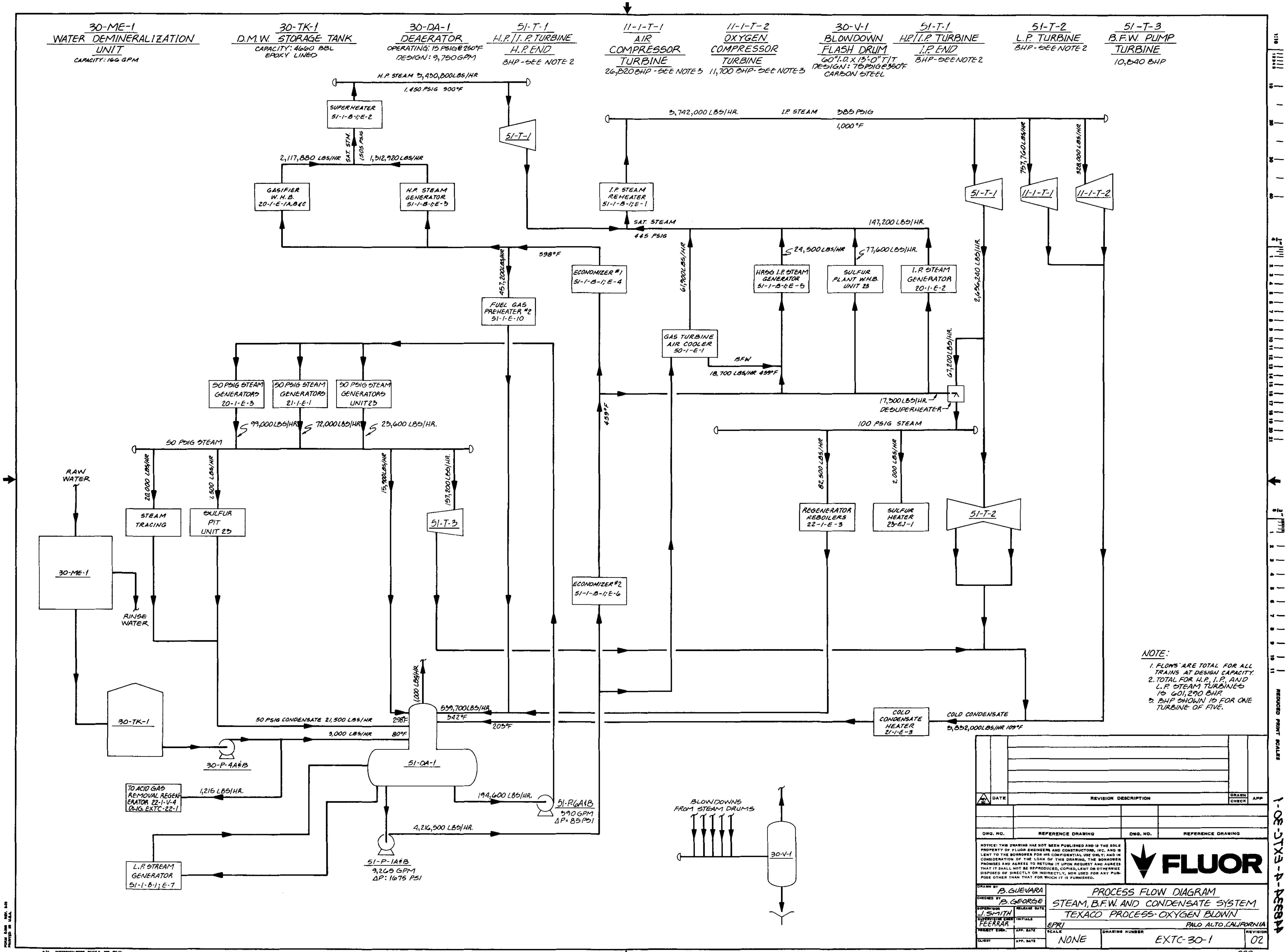
The process plant steam generation is integrated with the combined cycle system. The steam system operates at four levels:

High Pressure	-	1450 psig, 900°F
Intermediate Pressure	-	400 psig
Low Pressure	-	100 psig
Low Pressure	-	50 psig

High pressure steam (H.P.) generation is carried out in heat recovery steam generators (HRSG) 51-1-B-1 of gas turbines 50-1-GT-1. There are seven gas turbines and each gas turbine has its own HRSG. The HRSG is described in detail in Appendix A. Additional H.P. steam generation is obtained in the gas cooling units, 20-1-ME-3. The saturated H.P. steam from the process units combines with the saturated steam produced in the H.P. evaporators (51-B-1:-E-3) and is superheated to 900°F in the HRSG superheaters (51-1-B-1:E-2). All the superheated H.P. steam is used to drive the single back pressure type turbine, 51-T-1. The H.P. end of Turbine 51-T-1 takes steam at 1450 psig, 900°F and exhausts at 400 psig.

Saturated intermediate pressure (IP) steam generation at 420 psig is also obtained in the IP steam generators located in the sulfur plant, gasification unit and the HRSGs (51-1-B-1:E-5). The saturated IP steam together with the exhaust steam from 51-T-1 is superheated to 1000°F in the HRSGs' reheaters (51-1-B-1:E-1). The superheated IP steam at 385 psig, 1000°F is then used in the IP end of 51-T-1 and the condensing turbines, 11-1-T-1 and 11-1-T-2. The IP end of 51-T-1 exhausts steam to the 100 psig steam header.

A portion of the 100 psig steam is desuperheated and supplied to the process exchangers (22-1-E-3) in the acid gas removal unit. A small quantity of the desuperheated IP steam is also used in the sulfur plant (Unit 23). The balance of the 100 psig exhaust steam is used in the LP turbine, 51-T-2. LP Turbine 51-T-2 is a condensing turbine exhausting at 2-1/2" Hg abs.


30-ME-1
WATER DEMINERALIZATION UNIT
CAPACITY: 166 GPM
30-TK-1
D.M.W. STORAGE TANK
CAPACITY: 4660 BBL
EPOXY LINED
30-DA-1
DEAERATOR
OPERATING: 15 PSIG@250°F
DESIGN: 9,750 GPM
51-T-1
H.P./I.P. TURBINE
H.P. END
BHP-SEE NOTE 2
11-1-T-1
AIR COMPRESSOR TURBINE
26,820 BHP - SEE NOTE 3
11-1-T-2
OXYGEN COMPRESSOR TURBINE
11,700 BHP-SEE NOTE 3
30-V-1
BLOWDOWN FLASH DRUM
60"I.D. x 13'-0" T/T
DESIGN: 75 PSIG@350°F
CARBON STEEL
51-T-1
HP/I.P. TURBINE
I.P. END
BHP-SEE NOTE 2
51-T-2
L.P. TURBINE
BHP-SEE NOTE 2
51-T-3
B.F.W. PUMP TURBINE
10,840 BHP
H.P. STEAM 3,430,800 LBS/HR
1,450 PSIG 900°F
SUPERHEATER
51-1-B-1;E-2
SAT. STM.
1,505 PSIG
2,117,880 LBS/HR
1,312,920 LBS/HR
GASIFIER W.H.B.
20-1-E-1A,B&C
H.P. STEAM GENERATOR
51-1-B-1;E-3
3,742,000 LBS/HR.
I.P. STEAM 385 PSIG
1,000°F
757,760 LBS/HR
328,000 LBS/HR
I.P. STEAM REHEATER
51-1-B-1;E-1
SAT. STEAM
445 PSIG
147,200 LBS/HR.
598°F
457,200 LBS/HR
ECONOMIZER #1
51-1-B-1;E-4
61,900 LBS/HR
24,500 LBS/HR
77,600 LBS/HR.
HRSG I.P. STEAM GENERATOR
51-1-B-1;E-5
SULFUR PLANT W.H.B.
UNIT 23
I.P. STEAM GENERATOR
20-1-E-2
2,636,240 LBS/HR.
FUEL GAS PREHEATER #2
51-1-E-10
GAS TURBINE AIR COOLER
50-1-E-1
BFW
18,700 LBS/HR 459°F
67,200 LBS/HR.
50 PSIG STEAM GENERATORS
20-1-E-3
50 PSIG STEAM GENERATORS
21-1-E-1
50 PSIG STEAM GENERATORS
UNIT 23
99,000 LBS/HR
72,000 LBS/HR
23,600 LBS/HR.
459°F
17,300 LBS/HR.
DESUPERHEATER
100 PSIG STEAM
50 PSIG STEAM
RAW WATER
20,000 LBS/HR
1,500 LBS/HR
15,900 LBS/HR
157,700 LBS/HR.
82,300 LBS/HR
2,000 LBS/HR
STEAM TRACING
SULFUR PIT
UNIT 23
51-T-3
REGENERATOR REBOILERS
22-1-E-3
SULFUR HEATER
23-EJ-1
51-T-2
30-ME-1
RINSE WATER
ECONOMIZER #2
51-1-B-1;E-6
NOTE:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. TOTAL FOR H.P., I.P., AND L.P. STEAM TURBINES IS 601,290 BHP.
3. BHP SHOWN IS FOR ONE TURBINE OF FIVE.
1,000 LBS/HR
30-TK-1
50 PSIG CONDENSATE 21,500 LBS/HR
298°F
539,700 LBS/HR
342°F
203°F
3,000 LBS/HR
80°F
COLD CONDENSATE HEATER
21-1-E-3
COLD CONDENSATE
3,832,000 LBS/HR 109°F
30-P-4A&B
51-DA-1
TO ACID GAS REMOVAL REGENERATOR 22-1-V-4
DWG. EXTC-22-1
1,216 LBS/HR.
194,600 LBS/HR.
51-P-6A&B
390 GPM
ΔP: 85 PSI
BLOWDOWNS FROM STEAM DRUMS
4,216,500 LBS/HR.
30-V-1
L.P. STEAM GENERATOR
51-1-B-1;E-7
51-P-1A&B
9,265 GPM
ΔP: 1675 PSI
DATE
REVISION DESCRIPTION
DRAWN
CHECK
APP
DWG. NO.
REFERENCE DRAWING
FLUOR
DRAWN BY B. GUEVARA
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
FEERRAR
PROCESS FLOW DIAGRAM
STEAM, B.F.W. AND CONDENSATE SYSTEM
TEXACO PROCESS-OXYGEN BLOWN
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EXTC-30-1
REVISION 02
329
EXTC-30-1

The 50 psig steam header is supplied by steam generation in process exchangers, 21-1-E-1, generators in the gas cooling unit, 20-1-ME-3, and the steam generators in the sulfur plant. The 50 psig steam is mainly used in condensing turbine 51-T-3, driving the boiler feedwater pump. A small amount of 50 psig steam is also used for steam tracing and in the sulfur pit.

Raw water is treated in a semiautomatic, resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for 1500 psig boiler. Demineralized water is stored in Tank 30-TK-1. Demineralized water from the storage tank is transported to the deaerator through Pumps 30-P-4A&B. A small quantity of the makeup water is withdrawn from the discharge of 30-P-4A&B and transported to Unit 22. The balance of the demineralized water flows to the deaerator, 51-DA-1. The condensate from the 100 psig and 50 psig steam users also flows to the deaerator.

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

The deaerator is a tray type unit operating at 15 psig. The deaerator provides for 10 minute storage. The deaerating steam is generated in the HRSGs' LP steam evaporators 51-1-B-1:E-7.

Boiler feedwater (BFW) from the deaerator is pumped through high pressure boiler feedwater pumps (51-P-1A&B) to the HRSGs, process HP steam generators and process IP steam generators located in Unit 20, sulfur plant waste heat boilers and HRSG's IP evaporators (51-1-B-1:E-5).

BFW to the HRSGs is first heated to the IP steam saturation temperature (459°F) in economizers 51-1-B-1:E-6. Part of the BFW is withdrawn downstream of 51-1-B-1:E-6 and supplied to the process IP steam generators, HRSG's IP evaporators 51-1-B-1:E-5 and the desuperheating station for the 100 psig steam header. The balance of the BFW is heated to the HP steam saturation temperature (598°F) in the HRSG's economizers 51-1-B-1:E-4. A portion of the hot high pressure BFW from 51-1-B-1:E-4 is used to preheat fuel gas in 51-1-E-10. The balance of the high pressure BFW flows to the HP steam generators located in gasification unit and HRSGs where saturated high pressure steam is generated.

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The 50 psig steam generators are supplied boiler feedwater by a separate set of pumps, 51-P-6A&B.

Case EXTC (Dry Feed)

The steam, BFW and condensate system for this case is similar to Case EXTC (Slurry Feed) with the exception that a portion of high pressure superheated steam is supplied to the gasification plant (Unit 20) to meet the process steam requirement of the gasifiers.

The demineralization unit in Case EXTC (Dry Feed) is also considerably larger in size than the one required for Case EXTC (Slurry Feed) as the high pressure process steam is not recovered as condensate.

COMBINED CYCLE SYSTEM

Case EXTC (Slurry Feed)

Process Flow Diagram EXTC-50/50-1 depicts the combined cycle system for Case EXTC (Slurry Feed). This diagram shows the total power block flows.

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

The combined cycle system for Case EXTC (Slurry Feed) has the following distinct features:

- Equipment for fuel gas preheating is provided in Case EXTC. Fuel gas produced in the process plant at 522 psig and 320°F is first heated to 580°F by heat exchange against hot feed water at 598°F extracted from the outlet of economizer one 51-1-B-1:E-4 in fuel gas heater #3, 51-1-E-10. The cooled feed water flows back to the deaerator. The fuel gas is further heated to 950°F in a coil 51-1-B-1:E-8 provided in the reheater section of HRSG. The hot fuel gas from the HRSG is subsequently expanded from 495 psig to 280 psig in expanders 11-1-EX-1 to supply a portion of the air compressors' power. The balance of the air compressors' power is provided by condensing steam turbines which take steam from the hot reheat line.

- Steam for the oxygen compressor drives is also taken from the hot reheat line.

- The boiler feed water pump drives use 50 psig saturated steam.

The process cooling loads, where possible, are integrated into the condensate and makeup. Approximately 338 MM Btu/hr of low level process heat is recovered by heating a cold condensate stream.

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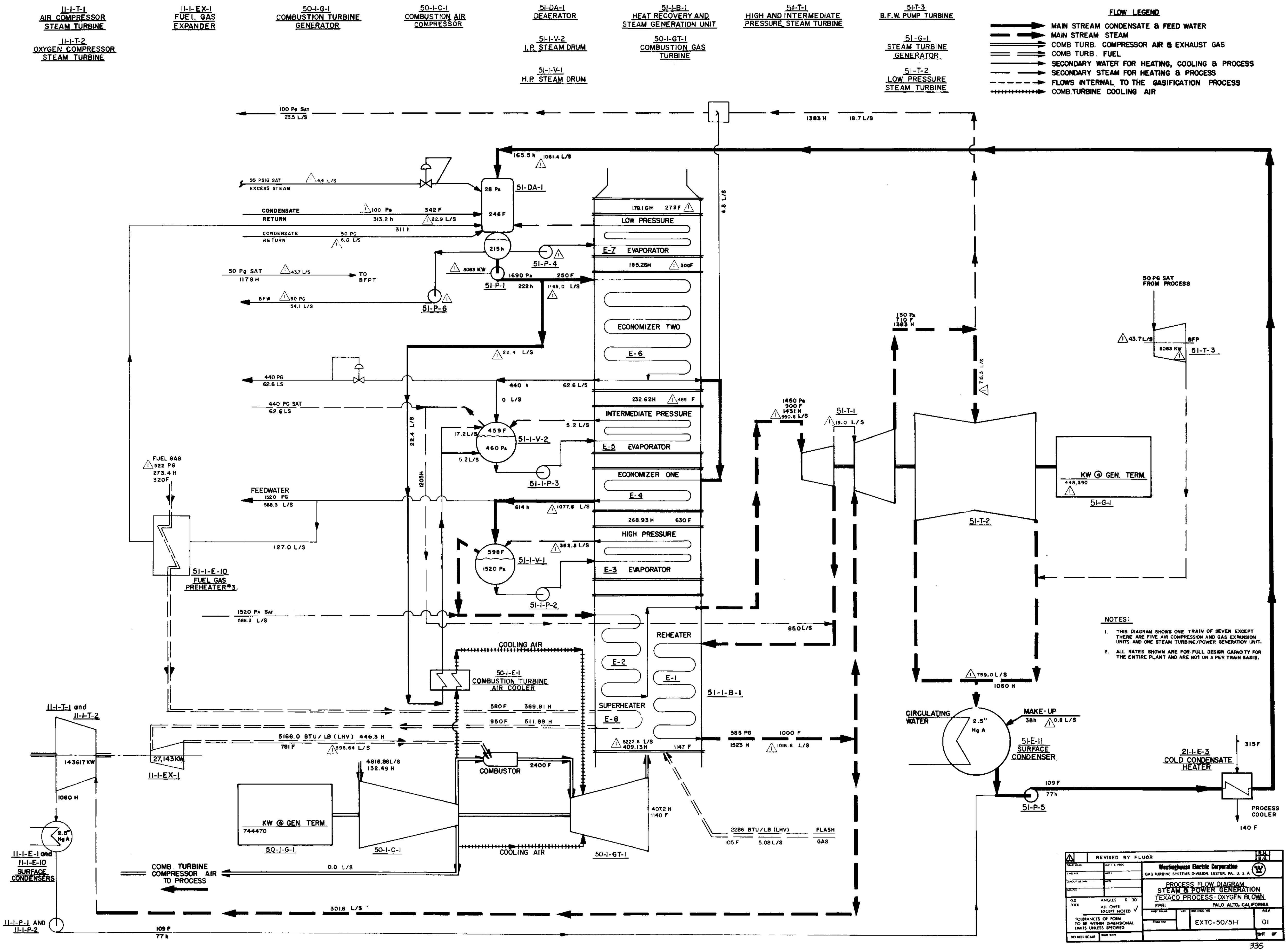

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Case EXTC (Dry Feed)

The equipment arrangement for the combined cycle system for this subcase is similar to Case EXTC (Slurry Feed) with small difference in total throughput.

There are seven parallel trains of gas turbines 50-1-GT-1, generators 50-1-G-1 and HRSG 51-1-B-1 and one 100 percent steam turbine (51-T-1 & T-2) and generators (50-G-1).

Equipment Notes

The proposed scheme of heating fuel gas in HRSG (51-1-B-1:E-8) is a standard practice in ammonia plants and refinery units such as hydrocrackers, hydrotreaters and crude units. Gaseous or liquid hydrocarbon streams are commonly heated in coils placed in fired heaters which have heat recovery sections similar to HRSGs.

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

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

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

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PROCESS DISCUSSION

The table below summarizes pertinent heat and material balance results.

TABLE ET-2

SUMMARY OF SYSTEM PERFORMANCE - CASE EXTC

	Case EXTC (Slurry Feed)	Case EXTC (Dry Feed)
GASIFICATION AND GAS CLEANING SYSTEM		
Coal Feed Rate, lbs/hr (m.f.)	798,333	798,333
Oxygen (1)/Coal Ratio, lbs/lb m.f.	0.858	0.806
Oxidant Temperature, °F	300	300
Steam/Coal Ratio, lbs/lb m.f.	0	0.610
Slurry Water/Coal Ratio, lbs/lb m.f.	0.503	0
Gasification Section Average Pressure, psig	600	600
Crude Gas Temperature, °F	2300-2600	2300-2600
Crude Gas HHV (dry basis), Btu/SCF (2)	281.1	280.7
Temperature of Fuel Gas to Gas Turbine, °F	781	781
POWER SYSTEM		
Gas Turbine Inlet Temperature, °F	2,400	2,400
Pressure Ratio	17:1	17:1
Turbine Exhaust Temperature, °F	1,140	1,140
Steam Conditions, psig/°F/°F	1,450/900/1,000	1,450/900/1,000
Condenser Pressure, Inches Hg abs	2.5	2.5
Stack Temperature, °F	272	272
Gas Turbine Power (3), MW	745	763
Steam Turbine Power (3), MW	448	425
Power Consumed, MW	36	46
Net System Power, MW	1,157	1,142
OVERALL SYSTEM		
Process and Deaerator Makeup Water, gpm/1000 MW	362	1,072
Cooling Tower Makeup Water, gpm/1000 MW	7,588	7,255
Cooling Water Circulation Rate, gpm/MW	347	352
Cooling Tower Heat Rejection, % of Coal HHV	38.7	35.6
Air Cooler Heat Rejection, % of Coal HHV	5.2	4.6
Net Heat Rate, Btu/kWh	8,813	8,928
Overall System Efficiency (Coal → Power), % of Coal HHV	38.7	38.2

(1) Dry Basis, 100% O_2
(2) Excluding the HHV of H_2S, COS and NH_3
(3) At Generator Terminals

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Gasifier Material Balances

Gasifier material balances for full capacity operation are given in Tables ET-3SF and ET-3DF for the oxygen blown Texaco gasifier cases.

Most of the data presented in the above two tables were received from Texaco Development Corporation. For the particular coal used for this study, Texaco indicated that slurry concentrations in the range of 60% solids to possibly 70% solids could be achieved. For the purposes of this work, EPRI selected a slurry concentration of 66.5% solids. It is important to keep in mind, however, the fact that slurrying characteristics of coals vary greatly and that it is not valid to extrapolate performance estimates presented in this report to other coals that will process different slurrying characteristics.

The dry feed case was a conceptual case developed by EPRI and Fluor. It was analyzed in order to estimate what incentive there might be to develop a method of feeding coal dry to the gasifier, instead of as a water slurry. Some thermal incentive was expected because vaporizing the water in the coal feed slurry requires heat, thereby increasing the oxygen requirement of the gasifier. Analysis of the dry feed case shows, however, that overall thermal efficiency is not improved compared to the slurry feed case. The requirement for gasifier feed steam is met at the expense of steam turbine power recovery, and effluent water treating requirements are increased. We thus conclude that there is no incentive to develop a dry feed system if the coal being considered can be slurried with less than 35% water.

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

TABLE ET-3SF

MATERIAL BALANCE-CASE EXTC (Slurry Feed)

FEEDS

	T (°F)	lb/hr	lb mol/hr
Coal	140		
Moisture		35,000	1,942.8
Ash		80,000	
MAF Coal			
Carbon		554,985	46,205.9
Hydrogen		42,525	21,094.6
Oxygen		80,022	2,500.8
Nitrogen		9,985	356.4
Sulfur		30,816	961.1
TOTAL COAL		833,333	
Oxidant	300		
Oxygen		684,687	21,397.3
Argon		4,326	108.3
Nitrogen		9,241	329.9
TOTAL OXIDANT		698,254	21,835.5
Water	140	366,553	20,364.1
TOTAL FEEDS		1,898,140	

EFFLUENTS

	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	2,300-2,600			
CH_4		1,158	72.2	0.08
H_2		52,364	25,974.2	28.84
CO		1,071,001	38,236.4	42.45
CO_2		345,232	7,844.4	8.71
H_2S		30,907	906.9	1.01
COS		3,256	54.2	0.06
N_2		16,725	597.1	0.66
Ar		4,326	108.3	0.12
H_2O		290,137	16,106.4	17.88
NH_3		3,034	178.1	0.19
TOTAL GASIFIER EFFLUENT		1,818,140	90,078.2	100.00
Ash	2,300-2,600			
Carbon		Nil		
Ash		80,000		
TOTAL ASH		80,000		
TOTAL EFFLUENTS		1,898,140		

TABLE ET-3DF

MATERIAL BALANCE-CASE EXTC (Dry Feed)

FEEDS

	T (°F)	lb/hr	lb mol/hr
Coal	140		
Moisture		35,000	1,942.8
Ash		80,000	
MAF Coal			
Carbon		554,985	46,205.9
Hydrogen		42,525	21,094.9
Oxygen		80,022	2,500.8
Nitrogen		9,985	356.4
Sulfur		30,816	961.1
TOTAL COAL		833,333	
Oxidant	300		
Oxygen		643,744	20,117
Argon		6,520	163
Nitrogen		4,006	143
TOTAL OXIDANT		654,270	20,423
Steam	900	340,765	18,914
Water(1)	140	110,900	6,156
TOTAL FEEDS		1,939,268	

EFFLUENTS

	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	2,300-2,600			
CH_4		1,153	71.9	0.08
H_2		58,787	29,160.1	30.85
CO		1,053,419	37,608.7	39.79
CO_2		372,870	8,472.4	8.97
H_2S		30,897	906.6	0.96
COS		3,268	54.4	0.06
N_2		14,288	410.1	0.43
Ar		6,520	163.4	0.17
H_2O		315,025	17,488.0	18.50
NH_3		3,034	178.1	0.19
TOTAL GASIFIER EFFLUENT		1,859,268	94,513.0	100.00
Ash	2,300-2,600			
Carbon		Nil		
Ash		80,000		
TOTAL ASH		80,000		
TOTAL EFFLUENTS		1,939,268		

(1) Water Content of Char Slurry

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In the dry feed case, the condensates cannot be recycled to the gasifier; therefore, they are treated in a water treatment plant and the ammonia is recovered.

The figures given in the tables and flow sheets for ammonia should be regarded as tentative estimates only.

In both cases, gasifier temperatures are believed to be high enough to destroy all hydrocarbon except methane.

Acid Gas Removal

One of the important design considerations in coal gasification is acid gas removal. Acid gas removal processes tend to absorb both hydrogen sulfide (H_2S) and carbon dioxide (CO_2). While in many applications removal of both is desired, for gas turbine power generation there are substantial disadvantages to removing CO_2. Absorption of CO_2 increases solvent circulation rates, equipment sizes and wasteful heat loads and takes away "working fluid" from the gas turbine generator. Further, the design and size of the downstream sulfur recovery units are affected in directions that increase cost. The Selexol® process removes H_2S in preference to CO_2 and, therefore, accomplishes an important objective. The process is used in these cases because it accomplishes this objective and it compares favorably with other similar processes economically.

The Selexol® process results in an H_2S concentration over 20 percent in the acid gas feed to the sulfur recovery unit in both the slurry feed and dry feed cases. At H_2S concentrations in this range, a sulfur plant design commonly referred to as "split flow" may be employed that avoids use of fuel gas in the sulfur furnace. Fuel gas must be burned in the furnace to sustain a flame if H_2S concentration is under 15 percent. In the split flow design the flame can be sustained by burning acid gas only.

Process Energy Balances

Tables ET-4SF and ET-4DF present overall process energy balances for slurry and dry feed cases at 100 percent capacity operation. The boundary for each balance encompasses the entire plant. Energy content of streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F and latent heat of water at 60°F. Electric power is converted to equivalent

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theoretical heat energy at 3413 Btu/kWh. These energy balances close to less than one percent. The discrepancies result from approximations used for some process units and for calculating some heat loads.

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

The tables show that the slurry feed case results in slightly more of the coal energy charged to the plant converted to power than does the dry feed case. Coal charged at 10,000 ton/day is equivalent to 10,196 MM Btu/hr HHV. The slurry feed case produces 3948 MM Btu/hr power equivalent or 38.7 percent of the coal HHV. The dry feed case produces 3898 MM Btu/hr or 38.2 percent of the coal HHV.

If all power consumed by the process units is included, the net system efficiency is 38.7 percent for the slurry feed case compared to 38.2 percent for the dry feed case. This reflects the slightly larger process power consumption by the dry feed case. The heat rate based on net power produced is 8813 Btu/kWh for the slurry feed case and 8928 Btu/kWh for the dry feed case. These net rates are close enough to be considered equivalent within the accuracy of the calculations done for these cases.

Comparisons drawn from the tables illustrate some of the differences between the slurry feed and dry feed cases. Oxygen consumption is less for the dry feed case, because it is not necessary to supply combustion heat to vaporize water. The gasifier steam consumption requirement for the dry feed case is satisfied at the expense of power generation from the steam cycle. Power generated in the gas turbine, however, is marginally higher for the dry feed case. This is offset by lower steam turbine power recovery and higher process power requirements for coal handling and transport gas compressors.

TABLE ET-4SF

ENERGY BALANCE CASE EXTC (Slurry Feed)

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

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

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

(1) Includes mechanical and electrical losses.

TABLE ET-4DF

ENERGY BALANCE - CASE EXTC (Dry Feed)

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

	MM Btu/hr					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Air Compressor Suction Air		21	50			71
Gas Turbine Combustion Air		121	291			412
Demineralized and Raw Water		9				9
Auxiliary Power Inputs					156	156
TOTAL	10,196	156	341	0	156	10,849
HEAT OUT						
Ash Slurry		81				81
Gasifier Heat Losses				26		26
Gas Cooling		20	6			26
Sulfur Product	105	1				106
Ammonia Product	25					25
Process Condensate Treating		47	23			70
Oxidant Compressor Interstage Cooling		503	31			534
Oxidant Compressor Surface Condensers			1,005			1,005
Gas Turbines					2,605	2,605
Sulfur Plant Effluent Gas		2	19			21
Steam Turbines					1,449	1,449
Power Block Losses (1)				43	171	214
Steam Turbine Condenser		2,570			2,570	
HRSG Stack Gas		1,067	821			1,888
Steam Heat Losses			22			22
Selexol Overhead Condenser			25			25
Selexol Solvent Cooler		56				56
Air Separation Plant Waste Gas		17				17
Waste Water Effluent		19				19
Pulverizer Vent Gas		24	21			45
TOTAL	130	1,837	4,543	69	4,225	10,804

$$\frac{\text{Input - Output}}{\text{Input}} = 0.41\%$$

(1) Includes mechanical and electrical losses.

TABLE ET-5

ENERGY BALANCE AS PERCENT COAL HHV

	Case EXTC (Slurry Feed)		Case EXTC (Dry Feed)	
	MM Btu/hr	Percent	MM Btu/hr	Percent
IN				
Coal HHV	10196	100.0	10196	100.0
OUT				
Net Power	3948	38.72	3898	38.23
Sulfur Product, HHV	105	1.03	105	1.03
Ammonia Product, HHV	0	0	25	0.25
Selexol Sensible and Latent	78	.77	81	0.79
Oxidant Interstage Cooling	568	5.57	534	5.24
Ash Slurry Sensible	81	0.79	81	0.79
HRSG Stack Gases	1817	17.82	1888	18.52
Rejected at Condensers	3754	36.82	3575	35.06
Other Sensible Losses	(94)	(.92)	(26)	(0.25)
Other Latent Losses	(288)	(2.82)	(250)	(2.45)
Gasifier Heat Losses	26	0.26	26	0.26
Power Block Losses	215	2.11	214	2.10
	10210	100.15	10151	99.57

Energy leaving the plant as stack gas is marginally higher for the dry feed case: 1888 MM Btu/hr versus 1817 MM Btu/hr for the slurry feed case. This represents 17.8 percent of the coal feed HHV for the slurry feed case and 18.5 percent for the dry feed case, and reflects the slightly higher fuel flow rate (particularly hydrogen flow rate) existing in the dry feed case.

ECONOMICS

Important economic results are summarized below.

TABLE ET-6

SUMMARY OF ECONOMICS - CASE EXTC

	Case EXTC (Slurry Feed)	Case EXTC (Dry Feed)
PRODUCTION AT DESIGN CAPACITY		
Net Power, MW (1)	1,156.8	1,142.0
Overall Plant Heat Rate, Btu/kWh	8,813	8,928
TOTAL CAPITAL (2)		
Total Capital @ $1/MM Btu Coal, $1,000	944,563	974,683
Total Capital @ $1/MM Btu, $/kW	816	854
Total Capital @ $2/MM Btu Coal, $1,000	961,681	991,855
Total Capital @ $2/MM Btu, $/kW	831	869
AVERAGE COSTS OF SERVICES (2)		
Annual Cost @ $1/MM Btu Coal, $1000/yr	262,088	267,861
Per Unit @ $1/MM Btu Coal, mills/kWh	37.21	38.25
Annual Cost @ $2/MM Btu Coal, $1000/yr	327,280	333,062
Per Unit @ $2/MM Btu Coal, mills/kWh	46.47	47.56

NOTES

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

The total capital requirements are about 3 percent higher and average annual cost of services nearly 2 percent higher in the dry feed case.

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

For all units, the dry feed plant investment is higher or equal to the slurry feed case except for the oxidant feed. This reflects the slightly higher oxygen requirements of the slurry feed case. The coal handling unit, waste water treating and steam, condensate and boiler feed water units are significantly higher in the dry feed case. The increase in coal handling reflects the added cost of a dry feed system for feeding coal to the gasifier. In the slurry feed system, process condensate can be returned to the gasifier, reducing waste water treating requirements and the size of the steam, condensate and boiler feed water system. This is reflected in the lower investment requirements for these units in the slurry feed case.

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

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

Operating labor requirements are a function of the number of units and trains. Requirements were the same for both cases and are shown below on a shift basis.

	Case EXTC (Slurry and Dry Feed)
"A" Operators	5
"B" Operators	17
Foreman	2
Lab and Instrument Technicians	4

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

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

Operating charges constitute about 43 percent of cost of services with coal at \$1.00/MM Btu and nearly 54 percent at a coal cost of \$2.00/MM Btu. For both cases, coal is the largest single operating charge. The relationship as percentage is summarized below:

	Case EXTC (Slurry Feed)		Case EXTC (Dry Feed)	
Cost of Coal, \$/MM Btu, HHV	1.00	2.00	1.00	2.00
Coal as % of Operating Charges	54.5	70.5	54.0	70.1
Coal as % of Total Cost of Services	23.9	38.2	23.3	37.5
Operating Charges as % of Total Cost of Services	43.8	54.1	43.2	53.5
Capital Charges as % of Total Cost of Services	56.2	45.9	56.8	46.5

TABLE ET-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $1.00/MM BTU COAL

	Case EXTC (Slurry Feed)			Case EXTC (Dry Feed)		
	$1,000 (1)	$/kW (2)	Percent	$1,000 (1)	$/kW (2)	Percent
PLANT INVESTMENT						
Coal Handling	22,061	19.07	3.56	29,320	25.67	4.58
Oxidant Feed	117,389	101.48	18.95	113,177	99.11	17.69
Gasification and Ash Handling	24,261	20.97	3.92	27,445	24.03	4.29
Gas Cooling	66,986	57.91	10.81	67,523	59.13	10.56
Acid Gas Removal and Sulfur Recovery	28,585	24.71	4.61	28,585	25.03	4.47
Process Condensate Treating	-	-	-	7,138	6.25	1.12
Steam, Condensate and BFW	827	0.71	0.14	2,335	2.05	0.37
Support Facilities	55,205	47.72	8.91	58,095	50.87	9.08
Combined Cycle	304,156	262.94	49.10	306,014	267.97	47.84
Subtotal	619,470	535.51	100.00	639,632	560.11	100.00
Contingency	118,160	102.14		122,001	106.83	
Total Plant Investment	737,630	637.65		761,633	666.94	
ILLINOIS SALES TAX	16,656	14.40		17,177	15.04	
CAPITAL CHARGES						
Preproduction Costs	46,342	40.06		47,746	41.81	
Paid-up Royalties	3688	3.19		3,808	3.34	
Initial Catalyst and Chemical Charge	515	0.44		515	0.45	
Construction Loan Interest	92,130	79.64		95,128	83.30	
Total Capital Charges	142,675	123.33		147,197	128.90	
DEPRECIABLE CAPITAL	896,961	775.39		926,007	810.88	
WORKING CAPITAL	47,602	41.15		48,676	42.62	
TOTAL CAPITAL	944,563	816.53		974,683	853.50	

NOTE

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

TABLE ET-8

CAPITAL INVESMENT AT 70% OPERATING LOAD FACTOR AND $2.00/MM BTU COAL

	Case EXTC (Slurry Feed)			Case EXTC (Dry Feed)		
	$1,000 (1)	$/kW (2)	Percent	$1,000 (1)	$/kW (2)	Percent
PLANT INVESTMENT						
Coal Handling	22,061	19.07	3.56	29,320	25.67	4.58
Oxidant Feed	117,389	101.48	18.95	113,177	99.11	17.69
Gasification and Ash Handling	24,261	20.97	3.92	27,445	24.03	4.29
Gas Cooling	66,986	57.91	10.81	67,523	59.13	10.56
Acid Gas Removal and Sulfur Recovery	28,585	24.71	4.61	28,585	25.03	4.47
Process Condensate Treating	-			7,138	6.25	1.12
Steam, Condensate & BFW	827	0.71	0.14	2,335	2.05	0.37
Support Facilities	55,205	47.72	8.91	58,095	50.87	9.08
Combined Cycle	304,156	262.94	49.10	306,014	267.97	47.84
Subtotal	619,470	535.51	100.00	639,632	560.11	100.00
Contingency	118,160	102.14		122,001	106.83	
Total Plant Investment	737,630	637.65		761,633	666.94	
ILLINOIS SALES TAX	16,656	14.40		17,177	15.04	
CAPITAL CHARGES						
Preproduction Cost	48,202	41.67		49,660	43.49	
Paid-up Royalties	3,688	3.19		3,808	3.34	
Initial Catalyst and Chemical Charge	515	0.45		515	0.45	
Construction Loan Interest	92,130	79.64		95,128	83.30	
Total Capital Charges	144,535	124.95		149,111	130.58	
DEPRECIABLE CAPITAL	898,821	777.00		927,921	812.56	
WORKING CAPITAL	62,860	54.34		63,934	55.99	
TOTAL CAPITAL	961,681	831.34		991,855	868.55	

NOTE

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

TABLE ET-9

COST OF SERVICES AT 70% OPERATING LOAD FACTOR

	Case EXTC (Slurry Feed)		Case EXTC (Dry Feed)	
COAL COST, HHV	$1/MM Btu	$2/MM Btu	$1/MM Btu	$2/MM Btu
NET PRODUCTION (1)				
Net Power, MW	1,156.8	1,156.8	1,142.0	1,142.0
By-product Ammonia ST/SD	0	0	30.3	30.3
By-product Sulfur ST/SD	301	301	301	301
OPERATING CHARGES, $1000/YEAR				
Coal	62,522	125,044	62,522	125,044
Operating Labor	2,692	2,692	2,692	2,692
Catalyst and Chemicals	262	262	284	284
Utilities	1,354	1,354	1,440	1,440
Maintenance, Labor	7,882	7,882	8,229	8,229
Maintenance, Materials	11,822	11,822	12,343	12,343
Administrative and Support Labor	3,172	3,172	3,276	3,276
General and Administrative Expenses	6,344	6,344	6,553	6,553
Ash Disposal	245	245	245	245
Property Tax/Insurance	18,441	18,441	19,041	19,041
By-product, Ammonia	(0)	(0)	(774)	(774)
By-product, Sulfur	(0)	(0)	(0)	(0)
Total Operating Charges, $1000/Year	114,736	177,258	115,811	178,333
CAPITAL CHARGES, $1,000/YEAR				
Total Capital Charges	147,352	150,022	152,050	154,729
COST OF SERVICES				
Total, $1,000/Year	262,088	327,280	267,861	333,062
Per Unit Production, mills/kWh	37.21	46.47	38.25	47.56

NOTES

(1) At 100% Operating Load Factor.

APPENDIX A

COMBINED CYCLE SYSTEM DETAILS

GENERAL

For each of the coal gasification processes a separate but similar combined cycle was selected. The work on this portion of the plant was performed by Westinghouse Electric Corporation, Lester, Pennsylvania based on interface conditions between the fuel processing and power section supplied by Fluor.

In some instances this work was done before correct final heat and material balances were available for the plant. Therefore, it was necessary for Fluor to update certain steam and condensate flows on the Westinghouse flow sheets. This has been done and noted as a Fluor revision.

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

A summary of the calculated power output for the power block equipment and heat loads rejected to the station cooling tower is presented in Table CS-1. The power output is calculated at the generator terminals without margins for design or manufacturing tolerances. The calculated power outputs include approximately 2.5 percent deduction for mechanical and electrical losses which include lube and seal oil pumps.

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TABLE CS-1

POWER BLOCK PERFORMANCE SUMMARY

	EALC	EALC-LP	MACW	MXSC	EXHC	EAHC	EXTC
GENERATION							
Gas Turbine, kW	885,962	772,065	589,905	856,737	802,731	750,596	744,470
Steam Turbine, kW	307,768	422,490	430,500[2]	385,142	384,446	500,786	448,390[2]
Total, Power Block, kW	1,193,730	1,194,555	1,020,405[2]	1,241,879	1,187,177	1,251,382	1,192,860[2]
HEAT REJECTION							
Process Cooling, M^2 Btu/hr	332.1[2]	330.0[2]	403.8[2]	397.2[2]	373.5[2]	258.9[2]	379.8
Process Cooling Absorbed in C.C., M^2 Btu/hr	146.2	146.6	114.5	0	78.7	4.8	338.1[2]
Process Cooling Rejection to Tower, M^2 Btu/hr	185.9[2]	183.4[2]	289.3[2]	397.2[2]	294.8[2]	254.1[2]	41.7[2]
Power Block Heat Rejection,[1] M^2 Btu/hr	3,592.6	3,798.9	2,882.8[2]	2,985.7[2]	3,034.9	3,436.1	3,806.2
Total Heat Rejection to Tower, M^2 Btu/hr	3,778.5[2]	3,982.3[2]	3,172.10[2]	3,382.9[2]	3,329.7[2]	3,690.2[2]	3,847.9[2]

Notes:

1. Includes mechanical and electrical losses of the power block.

2. Fluor revision.

TECHNICAL INPUT DATA

A common set of technical data was established and followed for all cases to provide equivalent treatment of the several coal gasification processes.

Gas Turbine

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

Steam Conditions

Steam conditions used for the combined cycle system are:

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

Heat Recovery Steam Generator (HRSG) Conditions

Steam production of the HRSG was calculated using a flue gas stack temperature of approximately 275°F. On the basis of demineralized boiler feedwater, boiler blowdown was assumed zero. The low pressure flash gas, where available from the gasification processes, has been burned as supplementary fuel in the HRSG.

Process Interface

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

Auxiliary Equipment

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

POWER BLOCK

Gas Turbine-Generator Unit

Gas Turbine (50-1-GT-1)

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

Overall engine performance was calculated using the Westinghouse computer program. The performance included both inlet air and exhaust duct losses to account for pressure drops through air silencers, ducting, afterburner and HRSG heat recovery sections.

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

Generator (50-1-G-1)

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

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

Steam Cycle Selection

HRSG

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

TABLE CS-2

GAS TURBINE PERFORMANCE SUMMARY

DESCRIPTION	EALC	EALC-LP	MXSC	EXHC	EAHC	MACW	EXTC
Compressor Inlet Air Duct Loss, H_2O	—	—	—	4.0	—	—	—
Compressor Disch. Pressure, psia	239.4	140.4	—	239.4	—	—	239.1
Compressor Disch. Temp., °F	857.0	669	—	857	—	—	857
Turbine Inlet Pressure, psia	229.1	134.4	—	229.1	—	—	228.8
Rotor Cooling Air, % of Inlet Flow	—	—	—	2.9	—	—	—
Rotor Coolant Temperature, °F	—	—	—	350	—	—	—
Turbine Exhaust System Loss, H_2O	—	—	—	20.5	—	—	—
Compressor Air Flow, lb/sec	3,677.29	3,312.23	5,798.79	5,227.98	5,379.71	3,906.96	4,818.86
Air to Process, lb/sec	0.0	0.0	0.0	0.0	646.05	449.19	0.0
Fuel Flow, lb/sec	1,159.13[2]	1,159.13[2]	350.49	378.88	848.57	669.28	399.14
Turbine Exhaust Temp., °F	1,147	1,317	1,128	1,133	1,127	1,137	1,140
Rotor Cooling Air Cooler Duty, Duty, MM Btu/hr	49.52	27.7	78.09	70.41	72.45	52.62	64.9
Power Output, kW[1]	885,962	772,065	856,737	802,731	750,596	589,905	744,470
Flash Gas Fuel Flow, lb/sec	0.0	0.0	1.63	3.66	2.25	27.22	5.08
Total Exh. Gas Flow, lb/sec	4,836.42[2]	4,471.36[2]	6,150.91	5,610.52	5,584.48	4,154.27	5,223.10
Exh. Gas Temp., Into HRSG, °F	1,147	1,317	1,133	1,139	1,131	1,180	1,147

Notes:

1. At generator terminals
2. Fluor revision.

The HRSG generates saturated steam at three pressure levels; high pressure (HP) at 1520 psia, intermediate pressure (IP) at 460 psia and low pressure (LP) at 28 psia. The low pressure steam generated in the HRSG is used in the deaerator. An approach temperature of 30°F (temperature of gas leaving - saturation temperature of steam) was used in calculating the steam generated in each of the evaporator sections. This approach temperature is in line with current HRSG design practice.

The typical arrangement of the heat recovery sections of the HRSG for all cases is shown in Figure CS-1. In the direction of exhaust gas flow the HRSG heat recovery sections for all the cases are as follows:

. Reheater	51-1-B-1:E-1
. Superheater	51-1-B-1:E-2
. HP Evaporator	51-1-B-1:E-3
. Economizer One	51-1-B-1:E-4
. IP Evaporator	51-1-B-1:E-5
. Economizer Two	51-1-B-1:E-6
. LP Evaporator	51-1-B-1:E-7

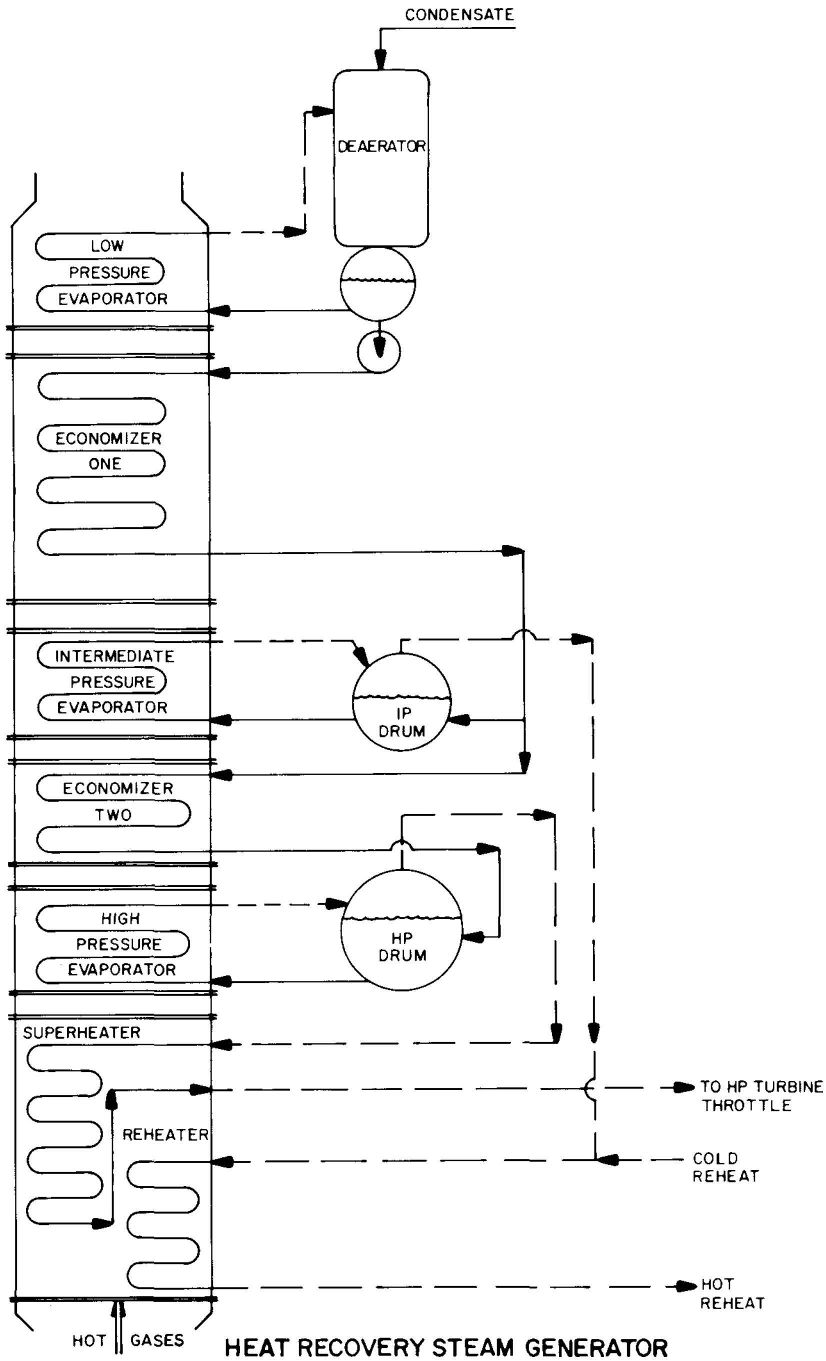
Case EXTC is provided with an additional heat recovery coil 51-1-B-1:E-8 in the high temperature zone of the HRSG for fuel gas heating.

The high pressure saturated steam generated in the HP evaporator is heated to 900°F in the superheater. If HP saturated steam is available from the process, it is combined with the HP steam from 51-1-B-1:E-3 ahead of the superheater.

The saturated steam produced in the IP evaporator section is combined with cold reheat steam from the high pressure steam turbine, 51-T-1, and superheated to 1000°F by passing through the reheater.

The LP evaporators located in the topmost section of the HRSG's supply deaerating steam to a tray type deaerator, 51-DA-1. The deaerator operates at approximately 28 psia. One common deaerator for the multiple HRSGs and the process steam generators is provided for each case.

The boiler feedwater (BFW) from the deaerator is first preheated to the IP steam saturation temperature (459°F) in Economizer Two. A portion of the BFW flows to the IP evaporator to meet the makeup water demand of the IP steam generation



HEAT RECOVERY STEAM GENERATOR
Figure CS-1

system. The balance of the BFW is further preheated to the HP steam saturation temperature (598°F) and flows to the HP evaporator. Interface between the process and HRSG's boiler feedwater system is provided if necessary.

Each HRSG is provided with its own LP, IP and HP steam drums and corresponding BFW circulation pumps.

The HRSG exhaust gas (stack) temperature of approximately 275°F, established in conjunction with the low pressure (28 psia) evaporator section, allows the gas side surface of the LP evaporator to operate safely above the sulfur dew point of the exhaust gas. The performance of the HRSG for each case is summarized in Table CS-3.

Steam Turbine - Generator Unit

Steam Turbine (51-T-1 and 2)

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

The HP end of 51-T-1 receives superheated HP steam at 1450 psig, 900°F and exhausts to the IP steam header operating at approximately 400 psig. The IP steam available after meeting the process IP steam demand is reheated to 1000°F in the HRSG reheaters (51-1-B-1:E-1) and flows to the IP end of 51-T-1. The inlet and exhaust conditions for the IP end of 51-T-1 are 385 psig, 1000°F and 115 psig respectively.

The LP end (51-T-2) is a condensing type unit receiving steam at 115 psig and exhausting at 2-1/2" Hg abs. The turbine is provided with the capacity of steam extraction at 50 psig (in Case EALC, also at 13 psia) to satisfy the overall steam balance of the plant. The surface condenser, 51-E-8, associated with 51-T-2 is designed for cooling water (CW) flow in two tube side passes with 85°F CW inlet temperature and 15°F CW temperature rise.

TABLE CS-3

HRSG PERFORMANCE SUMMARY

	EALC	EALC-LP	MACW	MXSC	EXHC	EAHC	EXTC
Exhaust Gas Flow, lb/sec	4,836.42[1]	4,471.36[1]	4,154.27	6,150.91	5,610.52	5,584.48	5,223.10
HP EVAP., SH & RH SECTIONS							
Exhaust Gas Temperature In, °F	1,147	1,317	1,180	1,133	1,139	1,131	1,147
SH Temperature Out, °F				900			
SH Pressure Out, psig				1,450			
SH Enthalpy Out, Btu/lb				1,431			
Sat. Steam Evap. lb/sec	550.9	733.3	628.1	786.3	644.1	553.5	362.3[1]
Sat. Steam from Process, lb/sec	0.0	0.0	0.0	0.0	253.0	290.3	588.3
SH Outlet Flow, lb/sec	550.9	733.3	628.1	786.3	897.1	843.8	950.6
SH Duty, MM Btu/hr	519.5	691.7	592.4	741.6	846.1	795.8	896.60
HP Drum Temperature, °F				598			
HP Drum Pressure, psia				1,520			
HP Drum Wtr. Enthalpy In, Btu/lb				614			
HP Evap. Duty, MM Btu/hr	1,100.5	1,465.1	1,254.9	1,571.0	1,286.9	1,105.9	724.0
RH Enthalpy In, Btu/lb	1,301.3	1,209.0	1,298.3	1,293.6	1,311.8	1,294.0	1,303.9
RH Temperature Out, °F				1,000			
RH Pressure Out, psig				385			
RH Enthalpy Out, Btu/lb				1,523			
RH Flow, lb/sec	1,022.6	1,135.3	521.5[1]	855.0	888.5	1,003.1	1,016.6[1]
RH Flow to Process, lb/sec	580.1	484.2	0.0	192.9	241.7	95.2	301.6
RH Duty, MM Btu/hr	816.2	875.9	421.8[1]	706.1	675.5	826.9	801.9[1]
Fuel Gas Heater Enthalpy In, Btu/lb	-	-	-	-	-	-	369.81
Fuel Gas Heater Enthalpy Out,Btu/lb	-	-	-	-	-	-	511.89
Fuel Gas Heater Flow, lb/sec	-	-	-	-	-	-	399.14
Fuel Gas Heater Duty, MM Btu/hr	-	-	-	-	-	-	204.2
Exhaust Gas Temperature Out, °F				630			

Note:
1. Fluor revision.

TABLE CS-3 (Continued)

ECONOMIZER NO. 1 SECTION	EALC	EALC-LP	MACW	MXSC	EXHC	EAHC	EXTC
Water Enthalpy In, Btu/lb				440			
Water Flow, lb/sec	550.9	733.3	628.1	786.3	644.1	553.5	1,077.6[1]
Water Flow to Process, lb/sec	0.0	0.0	0.0	0.0	0.0	0.0	588.3[1]
Water Flow to Fuel Gas Heater, lb/sec	0.0	0.0	0.0	126.1	0.0	0.0	127.0
Duty, MM Btu/hr	345.3	459.3	393.4	571.5	403.5	346.7	675[1]

IP EVAPORATOR SECTION	EALC	EALC-LP	MACW	MXSC	EXHC	EAHC	EXTC
IP Drum Temperature, °F				459			
IP Drum Pressure, psia				460			
IP Steam Enthalpy Out, Btu/lb				1,205			
IP Steam Evap., lb/sec	104.0	45.4	57.8	84.7	121.7	139.2	5.2
IP Evap. Duty, MM Btu/hr	286.5	125.0	159.2	233.3	335.2	383.3	14.4
IP Steam from Air Cooler, lb/sec	13.1	5.7	13.9	20.7	18.7	19.2	17.2
IP Steam to (from) Process, lb/sec	6.0	6.0	0.0	(48.1)	130.5	(17.9)	(62.6)
IP Steam to Cold RH, lb/sec	111.1	45.1	71.7	153.5	9.9	176.3	85.0
Water Enthalpy In, Btu/lb				440			
Exhaust Gas Temperature Out, °F				490			

TABLE CS-3 (Continued)

ECONOMIZER NO. 2 SECTION	EALC	EALC-LP	MACW	MXSC	EXHC	EAHC	EXTC
Water Enthalpy In, Btu/lb				222			
Water Flow, lb/sec	961.6	1,079.4	687.4	1,003.2	893.6	796.6	1,145.0[1]
Outlet Flow to Process, lb/sec	310.4	309.6	5.7	12.3	133.4	109.7	62.6
Duty, MM Btu/hr	754.7	847.1	539.5	787.3	701.3	625.1	898.6[1]
Exhaust Gas Temperature Out, °F	318	282	350	350	354	367	300[1]

LP EVAPORATOR & DA SECTION	EALC	EALC-LP	MACW	MXSC	EXHC	EAHC	EXTC
LP Drum Temperature °F				246			
LP Drum Pressure, psia				28			
Cond. Flow In, lb/sec	1,042.7	1,158.8	1,004.1[1]	888.8	973.1	980.7	1,061.4[1]
Cond. Enthalpy In, Btu/lb	164.0	207.5	107.0	72.3	94.0	77.0	165.5
Process Flows In, lb/sec	315.2	315.2	86.4	196.5	215.2	194.0	160.3[1]
FW Flow to Process, lb/sec	379.2	379.2	164.4	54.7	270.4	353.1	54.1
Duty, MM Btu/hr	190.9	30.1	288.6	419.7	405.0	469.6	133.5
Exhaust Gas Temperature Out, °F				275			272

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

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

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

TABLE CS-4

STEAM TURBINE

PERFORMANCE SUMMARY

HP ELEMENT	EALC	EALC-LP	MACW	MXSC	EXHC	EAHC	EXTC
Throttle Conditions:	1,450 psig/900°F TT						
Steam Enthalpy In, Btu/lb	1,431						
Throttle Flow from HRSG, lb/sec	550.9	733.3	628.1	786.3	897.1	843.8	950.6[2]
Throttle Flow from Process, lb/sec	379.2	379.2	220.3[2]	0.0	0.0	0.0	0.0
Total Throttle Flow, lb/sec	930.1	1,112.5	848.4[2]	786.3	897.1	843.8	950.6[2]
Exh. Flow to Process, lb/sec	0.0	0.0	381.6[2]	69.1	0.0	0.0	0.0
Exh. Enthalpy, Btu/lb	1,313						
IP ELEMENT							
Inlet Conditions:	385 psig/1,000°F TT						
Inlet Enthalpy, Btu/lb	1,523						
Inlet Flow, lb/sec	442.5	651.1	741.8[2]	662.1	646.8	907.9	715.0[2]
Exh. Enthalpy, Btu/lb	1,383						
Exh. Flow to Process, lb/sec	0.0	0.0	48.1	41.4	60.4	53.6	18.7
Exh. Flow to BFP, lb/sec	28.6	31.1	20.6[2]	22.9	25.0	24.8	0.0
LP ELEMENT							
Extr. Flow to Process, lb/sec	4.8	4.8	0.0	10.2	12.4	19.6	0.0
Extr. Flow to FWH, lb/sec	46.3	103.2	0.0	0.0	0.0	0.0	0.0
Turbine Exh. Flow, lb/sec	381.4	534.3	690.1[2]	603.1	567.3	826.6	715.3[2]
Exh. Enthalpy, Btu/lb	1,060						
Total Flow to Condenser, lb/sec	410.0	565.4	710.7[2]	626.0	592.3	851.4	759.0[2]
Cond. Circ. Water Flow, gpm	162,660	224,416	281,342[2]	246,805	481,015	335,968	282,461[2]
Power Output, kW[1]	307,768	422,490	430,500[2]	385,142	384,446	500,786	448,390[2]

Notes:
1. At Generator Terminals
2. Fluor revision

EQUIPMENT STATE OF THE ART

Gas Turbine

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

- gas turbine compressor pressure ratio
- high temperature turbine operation

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

Additional constraints apply to Case EALC-LP. Refer to process description of the Combined Cycle System for Case EALC for a discussion of these constraints.

Gas Turbine Compressor Pressure Ratio

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

Performance studies on combined cycles operating with gas turbine inlet temperatures in the 2400°F range have shown the optimum pressure ratio to be near 14 to 1. See Figure CS-2. Because of this no one has yet undertaken development of large single spool gas turbines having fixed (maximum of two variable stage) compressor geometry for these higher compression ratios.

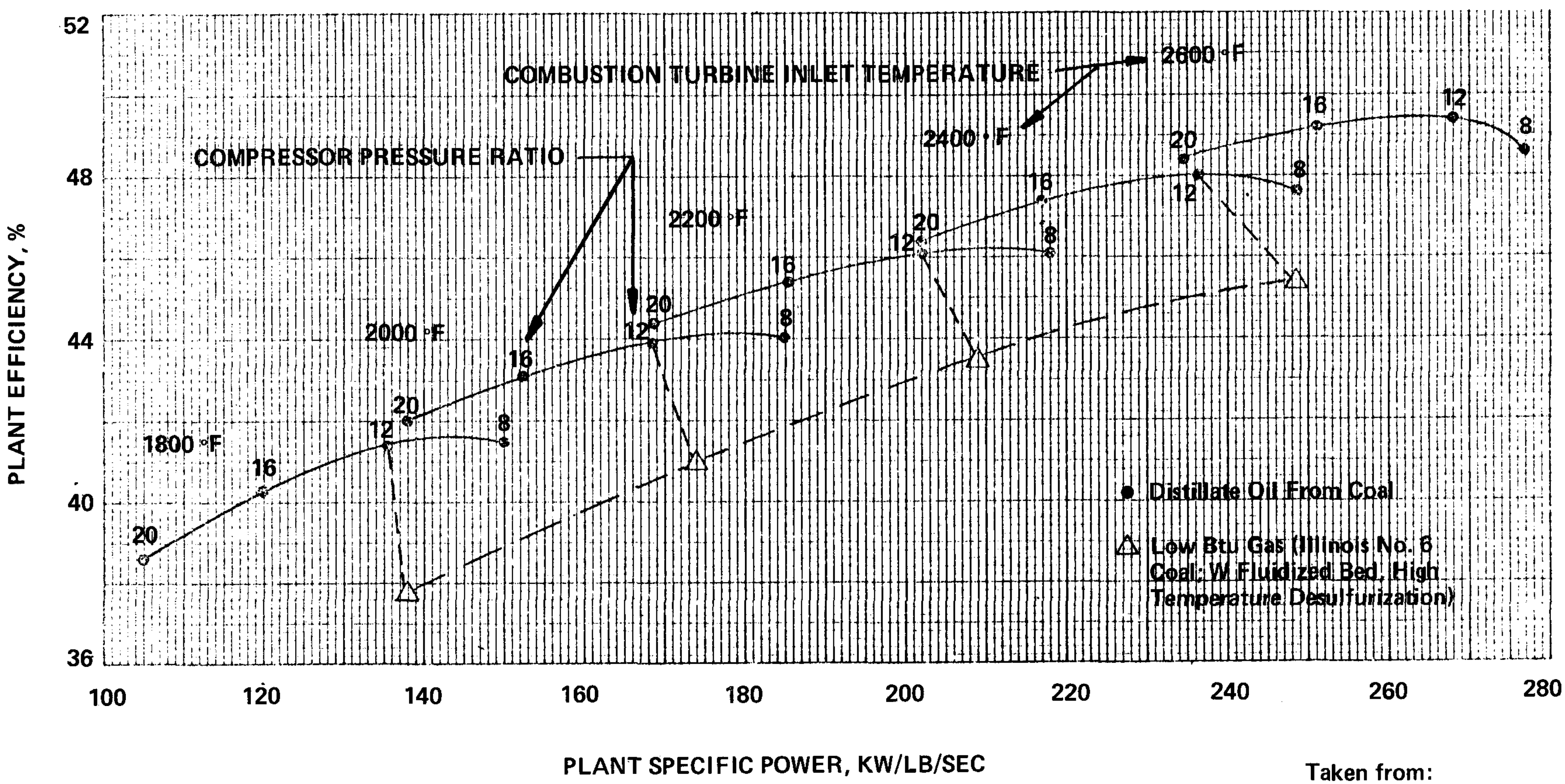
A more desirable pressure ratio for a gas turbine operating at 2400°F turbine inlet temperature is 14:1 at ISO conditions rather than the approximate 19:1 selected for this study, based on Figure CS-2.

ISO AMBIENT

ADVANCED GAS TURBINE

2400 PSIG/1000° F/1000° F STEAM CYCLE

WET COOLING TOWER 2" HG ABS



COMBINED CYCLE PLANT PERFORMANCE

FIGURE CS-2

Taken from:
W Phase I, Vol. V.
ECAS Report
NAS3-19407

Although Figure CS-2 shows a pressure ratio of approximately 12:1 to be nearer optimization for performance, the 14:1 ratio is recommended for mechanical reasons, such as temperature limit on the uncooled last row of combustion turbine blading.

Assuming that the gas turbine operating characteristics when burning a coal-derived gas are similar to that when burning a coal-derived oil, an approximation of performance improvement can be made. From Figure CS-2, the performance gain would be 1.5-2 percent in efficiency with a corresponding saleable power increase of 8-10 percent. These performance figures would then indicate an overall coal gasification combined cycle plant, coal to busbar. Further parametric study on each system would be required to determine final affect upon performance due to operating pressure ratio.

High Temperature Turbine Operation

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

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

HRSG

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

Steam Turbine

Although the selected throttle steam conditions of 1450 psig/900°F/1000°F reheat present no problem to the state of art, current industry practice with machines in the size range of this study would have throttle pressures of 1800 and 2400 psig.

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APPENDIX B

COAL FIRED BOILER WITH STACK GAS SCRUBBER

TABLE B-1

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $1.00/MM BTU COAL - COAL FIRED BOILER

	$1000 (1)	$/KW (2)
TOTAL PLANT INVESTMENT (3)	629,000	629.00
CAPITAL CHARGES		
Preproduction Costs	41,707	41.71
Paid-up Royalty	0	0
Initial Catalyst and Chemical Charges	14	0.01
Construction Loan Interest	121,460	121.46
Total Capital Charges	163,181	163.18
DEPRECIABLE CAPITAL	792,181	792.18
WORKING CAPITAL	46,229	46.23
TOTAL CAPITAL	838,410	838.41

NOTES
- (1) Mid-1976 Dollars.
- (2) Based on 100% Operating Load Factor.
- (3) Includes Illinois State Sales Tax.

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TABLE B-2

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $2.00/MM BTU COAL - COAL FIRED BOILER

	$1000 (1)	$/KW (2)
TOTAL PLANT INVESTMENT (3)	629,000	629.00
CAPITAL CHARGES		
Preproduction Costs	43,519	43.52
Paid-up Royalty	0	0
Initial Catalyst and Chemical Charge	14	0.01
Construction Loan Interest	121,460	121.46
DEPRECIABLE CAPITAL	793,993	793.99
WORKING CAPITAL	61,087	61.09
TOTAL CAPITAL	855,080	855.08

NOTES

(1) Mid-1976 Dollars.
(2) Based on 100% Operating Load Factor.
(3) Includes Illinois State Sales Tax.

TABLE B-3

COST OF SERVICE AT 70% OPERATING LOAD FACTOR - COAL FIRED BOILER

COAL COST, HHV	$1/MM BTU	$2/MM BTU
NET PRODUCTION		
Net Power, MW	1000	1000
OPERATING CHARGES, $1000/YEAR		
Coal	60,883	121,766
Operating Labor	867	867
Catalyst and Chemicals	4,726	4,726
Utilities	2,081	2,081
Sludge Disposal	12,264	12,264
Ash Disposal	32	32
Maintenance, Labor	6,600	6,600
Maintenance, Materials	10,000	10,000
Administrative and Support Labor	2,240	2,240
General and Administrative Expenses	4,480	4,480
Property Tax and Insurance	15,725	15,725
Total Operating Charges	119,898	180,781
CAPITAL CHARGES, $1000/YEAR		
Total Capital Charges	130,792	133,392
COST OF SERVICES		
Total, $1000/Year	250,690	314,173
Per Unit Production, mills/kwh	40.88	51.23

TABLE B-4

SYSTEM OPERATING REQUIREMENTS - COAL FIRED BOILER

Net Power Produced, MW	1000
Coal Feed Rate, lb/hr	983,040
Sludge Produced, tons/day	4,947
Ash to Disposal, tons/day	87
Chemical Requirements	
Limestone, tons/day	1205
Lime, tons/day	25.6
Chlorine, tons/day	0.9
Operating Labor, men/shift	9
Raw Water Makeup, gpm/1000MW	14,145
Cooling Tower Circulation, gpm/MW	360
Net Heat Rate, BTU/kWh	9928
Overall System Efficiency (Coal → Power) % of Coal HHV	34.4

APPENDIX C

AREA AND UNIT NUMBERING

Each plant consists of a number of facilities or systems called units. The units are grouped into areas having similar general purposes. The areas and units are numbered according to a consistent convention for identification. The table below shows the area and unit[1] numbering system.

AREA/UNIT NUMBERING SYSTEM

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

Note

1. Case EALC is provided with an additional onsite unit 25 (Area 20) designated as Product Gas Compression.

APPENDIX D

CASE DESIGNATIONS

A letter code has been prepared to shorten and simplify the references to the various cases given in this report. The first four letters characterize the case, as shown below.

Case MACW - Moving bed, Air blown, Combined cycle plant operating on Western coal. This plant uses the Lurgi dry ash gasifier.

Case MXSC - Moving bed, oXygen blown, Slagging gasifier, with a Combined cycle plant. This plant uses the British Gas Corporation Slagging gasifier.

Case EAHC (EXHC) - Entrained bed, Air (oXygen) blown, High pressure gasifier with a Combined cycle power plant. These cases are based on the Foster Wheeler Gasifier.

Case EALC - Entrained bed, Air blown, Low pressure, Combined cycle power plant. A subcase designated LP is based on a reduced gas turbine combustor pressure. These cases are based on the Combustion Engineering gasification process.

Case EXTC - Entrained bed, oXygen blown, Texaco gasifier, with a Combined cycle power plant. A subcase designated DF is based on Dry coal Feed, to provide a comparison with the normal approach of feeding the coal in a slurry. At certain locations in this report, the water Slurry Feed case is designated EXTC-SF.