Economic Evaluation of GCC Power Plants Based on the STEAG Combined-Cycle Design and Comparison With a U.S. Combined-Cycle-Based System

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ABSTRACT

This report presents the results of preliminary process design and economic screening studies of four integrated gasification-combined-cycle (GCC) power plant systems. Three of these GCC systems were based on the pressurized boiler combined-cycle design developed and demonstrated by STEAG in West Germany. As a basis for comparison, the fourth GCC system utilized a typical U.S. combined-cycle configuration as designed by United Technologies Corporation. All of the combined-cycle designs presented in this report used commercially available components, including the gas turbines. The major objective of this study was the preparation of a set of comparative evaluations which would indicate whether the STEAG combined-cycle design offers significant advantages that would make it a preferred choice for use in GCC systems now being developed for the U.S. electric power industry.

The gasification processes used in the three STEAG-based GCC systems included the air-blown Lurgi dry ash process, the oxygen-blown Texaco entrained-bed process, and the air-blown Texaco entrained-bed process. The U.S. combined-cycle-based GCC system also utilized the oxygen-blown Texaco process. All designs and accompanying evaluations were based on complete "grass roots" facilities sized to conform to a baseload power plant of approximately 1000 MW capacity.

The U.S. combined-cycle-based GCC system yielded both the highest overall thermal efficiency (34.8 percent) and the lowest first-year busbar power cost (38.9 mills /kWh) of the four systems studied. It was concluded that the three specific STEAG-based GCC systems supplied for this study provide no incentive for their development and use by the U.S. electric power industry. Of these three STEAG-based systems, the design which employed the oxygen-blown Texaco gasification process showed the most potential for improvement. It was also concluded that both the results of this study and the suggested improvements in the STEAG-based designs provide no evidence to suggest that the STEAG combined cycle is superior to the U.S. combined cycle in a GCC system application.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This final report, Economic Evaluation of GCC Power Plants Based on the STEAG Combined-Cycle Design and Comparison with a U.S. Combined-Cycle-Based System, is one of a series of gasification-combined-cycle (GCC) system evaluations performed by Fluor Engineers and Constructors, Inc. under Research Project (RP) 239-2. Results of a previous set of economic evaluations of GCC systems were presented in EPRI Final Report AF-642. The specific coal gasification technologies evaluated in the AF-642 study included the Lurgi dry ash gasifier, the British Gas Corporation Slagger, the two-stage entrained Foster Wheeler device, the Combustion Engineering atmospheric pressure entrained gasifier, and the oxygen-blown Texaco entrained gasifier. These GCC evaluations were further supplemented by a study of an air-blown Texaco-based system as presented in EPRI Final Report AF-753. A recent evaluation of the Toscodyne GCC system was published as EPRI Final Report AF-930. All of the GCC systems evaluated in previous studies for this project used U.S. combined-cycle designs based on highly advanced gas turbines with a 2400°F combustor exit temperature. Commercialization of these advanced gas turbines is projected for the mid- to late-1980s.

The present study extends the previous series of evaluations to GCC systems in which the combined-cycle designs are based on commercially available components, including the gas turbines. Particular emphasis is placed on systems that employ the pressurized boiler combined-cycle design developed and demonstrated by STEAG, a major producer of electric power in West Germany. The STEAG combined cycle design has been proven in a 170-MW GCC demonstration unit at Lünen, West Germany. This demonstration unit employs the air-blown Lurgi coal gasification process and has operated successfully for over 8600 hours. At the time of this writing, the 170-MW STEAG unit at Lünen is the only existing system that has demonstrated the successful coupled operation of a coal gasification process and a combined-cycle power plant.

This report further extends the previous studies by providing for a comparison between two different combined-cycle configurations in a GCC system application. The sequence of power generation steps in the STEAG combined cycle is essentially the reverse of that used in a U.S. combined cycle. Furthermore, the great majority of the gross power generated in the STEAG combined cycle is derived from the steam cycle. In contrast, the U.S. combined cycle derives the majority of its gross power output from the gas turbine cycle.

PROJECT OBJECTIVES

A major objective of this study was to determine if the STEAG combined-cycle design offers advantages that would make it the preferred choice for use in GCC systems now under development for the U.S. electric power industry. The approach taken to satisfy this objective was to prepare a set of four GCC system evaluations that would allow certain comparisons. In view of the fact that STEAG has published papers claiming up to 40 percent thermal efficiency for its Lünen unit, an evaluation was prepared for a GCC system based on the air-blown Lurgi gasification process and the STEAG combined cycle. Since the oxygen-blown Texaco process is considered to be the preferred second generation gasification technology for GCC system applications, two separate evaluations with different combined cycles were based on this oxygen-blown process. The STEAG combined-cycle design was used in one of these evaluations while a U.S. combined-cycle design provided by United Technologies Corporation (UTC) was employed in the other. Finally, the air-blown Texaco gasification process was chosen for integration with the STEAG combined cycle in the fourth evaluation. This choice was based on the premise that the steam-cycle-intensive STEAG system would benefit from the additional steam-raising capability which arises from cooling the air-blown Texaco gasifier effluent. The STEAG group in Essen, West Germany, provided all three of the STEAG combined-cycle designs used in this study.

The second major objective of this study was to develop evaluations in which the combined-cycle designs were based on commercially available components. Both STEAG and UTC were instructed to adhere to this criteria. In an attempt to encourage some innovation in the designs provided for this study, the suppliers of the combined cycles were given a certain degree of freedom in specifying basic parameters such as steam-cycle conditions, gas turbine combustor exit temperature, and the flue gas stack temperature. As a result, STEAG and UTC have adopted different values for these basic design parameters.

An additional objective in the present study was to maintain a degree of process consistency with previous GCC evaluations based on the same gasification technologies. Accordingly, the air-blown Lurgi gasifier performance data used in the present study were the same as those employed in the Lurgi-based evaluation contained in EPRI AF-642. These performance data are based on a Western U.S. coal. In the case of the oxygen-blown Texaco process, both the gasifier performance data and the design configuration adopted for heat integration are the same as those utilized in the EPRI AF-642 evaluation. The air-blown Texaco gasifier performance data and heat integration design are the same as those used in the EPRI AF-753 study. All Texaco systems are based on an Illinois No. 6 coal feed.

A final objective in the present work was to provide estimates of Total Plant Investment that were consistent with those developed for previous evaluations performed by Fluor. Each of the combined-cycle designers supplied cost estimates for the basic equipment contained in its system. Equipment costs for the gasification processes and related systems were estimated by Fluor. Development of both the total installed costs for all plant sections and the final Total Plant Investment for each case was also performed by Fluor. This methodology for cost estimating is exactly the same as that used in previous evaluations completed for this project. Use of mid-1976 dollars as the cost basis was also retained for purposes of consistency.

Owing to recent changes in EPRI Economic Criteria, the Total Capital Requirements and busbar power costs developed for the present study are not directly comparable with those presented in past studies. An updating effort is now under way that will present economics for all evaluations performed under this project on a mid-1978 dollar basis according to a single set of economic criteria.

PROJECT RESULTS

A major conclusion reached in this report is that the three specific STEAG-based GCC systems supplied by STEAG for this study provide no incentive for their development and use by the U.S. electric power industry. The relatively low thermal efficiency (30.6 percent) obtained for the Lurgi-based GCC system is particularly disappointing in view of STEAG's claims for its Lunen unit. A review of STEAG's publications suggests that the claims for high efficiency result from analyses that do not include energy debits for systems such as acid gas removal, process condensate treating, and the tar boiler operation. These systems will definitely be required for any baseload GCC power plant built in the U.S. The evaluation presented in this study accounts for all of these factors. In addition, it should be noted that European power companies such as STEAG usually report percent thermal efficiencies on a coal LHV basis rather than on a coal HHV basis as was done in this study. Use of coal LHV as the basis will increase calculated overall thermal efficiencies by as much as two to three percentage points above the efficiencies based on HHV.

STEAG's plans for piloting development of improvements in the Lurgi coal gasification process should be viewed with interest. Since the basic Lurgi process has been demonstrated commercially, additional improvements can be built on a foundation of "known" technology. Modifications that substantially improve the thermal efficiency of the Lurgi process may have a strong impact on the future use of this technology in the U.S. Depending upon STEAG's progress in its planned development work, it may be desirable to consider a future evaluation of a Lurgi-based GCC system that incorporates these improvements.

This report contains an extensive discussion of the cost-related problems associated with the heat integration designs used in the two Texaco-based cases that employ the STEAG combined-cycle design. These heat integration designs were adopted to satisfy the process consistency objective of this study. An examination of the resulting cost impact of steam generation conditions in the heat transfer equipment used for gasifier effluent cooling leads to an additional conclusion in this study. Specifically, Texaco-based GCC system designs tend to show more favorable economics if relatively mild steam generation conditions (saturated steam at less than 2000 psig, no superheating) are used in the gasifier effluent heat recovery equipment.

It should be recognized that the above-stated conclusion regarding steam generation conditions may not be applicable to a system designed to produce only medium-Btu fuel gas using the Texaco coal gasification process. An economic analysis for such a system may indicate that it is preferable to bear the increased cost for equipment to superheat steam by gasifier effluent cooling as opposed to burning valuable product gas in a fired superheater. Evaluation of a Texaco-based medium-Btu gas plant is a future study planned for this project.

The results obtained for the U.S. combined-cycle-based case (Case EXTC78) in this study are extremely significant. This case is the first GCC evaluation produced for this project that uses both the oxygen-blown Texaco coal gasification process and a U.S. combined-cycle design based on commercially available gas turbines. Within the limits of accuracy that apply to this type of evaluation, both the overall thermal efficiency and first-year power costs for Case EXTC78 are competitive with those for a coal-fired boiler plant equipped with flue gas desulfurization. Conservative economic evaluations of benchmark coal-fired power plants have been published in both EPRI AF-642 and EPRI AF-1182.

As discussed in the present report, Case EXTC78 is not an optimized system. Mention is made of the potential economic impact of reducing the cost of the combined-cycle section. In addition, studies now being conducted under Research Project 986 have shown that oxygen-blown Texaco-based GCC systems which utilize commercially available gas turbines can be configured to yield thermal efficiencies of 36 to 38 percent. These studies have been made by Westinghouse, General Electric, and UTC. An evaluation is now being prepared for this project to confirm the performance and develop the economics for a high-efficiency Texaco-based GCC system that employs commercially available gas turbines.

The final and most important conclusion in this study is reached by considering the basic results obtained for both the U.S. combined-cycle-based case EXTC78 and the three STEAG-based cases. This study contains no evidence to indicate that the STEAG combined-cycle design is superior to the U.S. combined cycle in a GCC system application. Consideration of <u>all</u> potential improvements in performance and economics presented for <u>both</u> the U.S. combined-cycle case and the STEAG cases does not alter this conclusion. The STEAG-based Case EXTS that employs the oxygen-blown Texaco process shows the greatest potential for improvement. However, the most favorable situation that can be envisioned at this time is that Case EXTS might be improved to the point where it becomes competitive with an improved version of Case EXTC78. Even if this most favorable situation is achieved for Case EXTS, no strong incentive will exist to pursue development of the STEAG combined cycle for the U.S. electric power industry. The U.S. combined-cycle configuration is already well-known and understood by the U.S. electric power industry. Switching to a totally different system configuration demands some evidence of <u>superior</u> performance and/or economics. This study contains no such evidence of superiority for the STEAG combined-cycle design.

Based on the results and conclusions of this study, it is recommended that:

- STEAG's progress with development work on Lurgi coal gasification process modifications be monitored.

- Unless system thermal efficiency clearly demands an exception, heat integration designs for oxygen-blown Texaco-based GCC plants should employ generation of saturated steam at less than 2000 psig in the gasifier effluent heat recovery equipment.

- The U.S. combined-cycle design should be preferred over the STEAG combined-cycle design in GCC systems now being developed for the U.S. electric power industry.

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SUMMARY

Combined-cycle configurations based on a pressurized boiler have been successfully developed and proven by STEAG, a major West German producer of electric power. The STEAG combined-cycle design differs from the systems used in North America in that fuel is combusted in a pressurized boiler which supplies steam to drive steam turbine generators. Exhaust gas from the pressurized boiler then passes to a gas turbine-generator system. Heat recovered from the turbine exhaust gas is used to heat boiler feedwater flowing to the pressurized boiler. The pressurized boiler cycle has the advantage of providing exhaust gas at temperatures under 1800°F from the boiler to the turbine inlet, allowing conventional turbine blading materials to be used. The STEAG combined cycle design has been proven in a 170 MW coal gasification combined cycle (GCC) plant operated by STEAG at Lünen, West Germany. This STEAG GCC plant utilizes the air-blown Lurgi gasification process and supplies electric power to a commercial distribution grid.

Combined cycle configurations widely used in North America utilize an atmospheric Heat Recovery Steam Generator (HRSG). These systems differ from the STEAG pressurized boiler design in that power is first produced by firing the fuel in a combustion turbine topping cycle with subsequent recovery of heat from the turbine exhaust in the HRSG. The steam generated and superheated in the HRSG is used to drive a steam turbine-based bottoming cycle. These cycles have been commercially demonstrated in many applications using oil or natural gas as the fuel.

These two alternate types of combined cycles are illustrated schematically in Figure S-1.

A major objective of this study was the development of a set of comparative evaluations which would indicate whether the STEAG design offers significant economic and/or performance advantages that would make it a preferred choice for use in GCC systems currently being developed for the U.S. electric power industry. In order to provide consistency with previous GCC evaluations, the design configurations adopted for heat integration of the gasification and combined-cycle

sections were analogous to those used in the Lurgi and Texaco-based systems described in EPRI Reports AF-642 and AF-753. Use of "commercially available" power generation equipment was a primary requirement for all combined-cycle designs developed for this study. Suppliers of these combined-cycle designs were given some freedom to specify certain operating parameters such as gas turbine combustor exit temperature, steam cycle conditions, and flue gas stack temperature.

Three gasification processes were studied in conjunction with the STEAG pressurized boiler combined cycle:

- Lurgi dry ash gasifier (Case MASW)
- Texaco oxygen-blown entrained gasification (Case EXTS)
- Texaco air-blown entrained gasification (Case EATS)

The combined-cycle designs for the above three cases were developed by STEAG. The designated letter codes for each case are explained in Appendix D.

The fourth GCC system in this study was based on integration of the oxygen-blown Texaco gasification process with a typical U.S. combined-cycle design supplied by United Technologies Corporation (UTC):

- Texaco oxygen-blown entrained gasification (Case EXTC78)

Current production model gas turbines with a combustor exit temperature of 1980°F were used in the combined-cycle system.

A major finding in this study was that the U.S. combined-cycle-based Case EXTC78 exhibited both the highest overall thermal efficiency and the lowest first-year and 30-year levelized power costs. Designs for GCC systems based on integration of the STEAG combined cycle with both the oxygen-blown and air-blown Texaco gasification processes showed somewhat lower thermal efficiencies and significantly higher power costs. A principal factor contributing to the higher power costs in each of these two cases was the estimated high cost of the required gas cooling system. These high costs for the gas cooling sections were a primary result of the heat integration designs used in each of these two GCC systems.

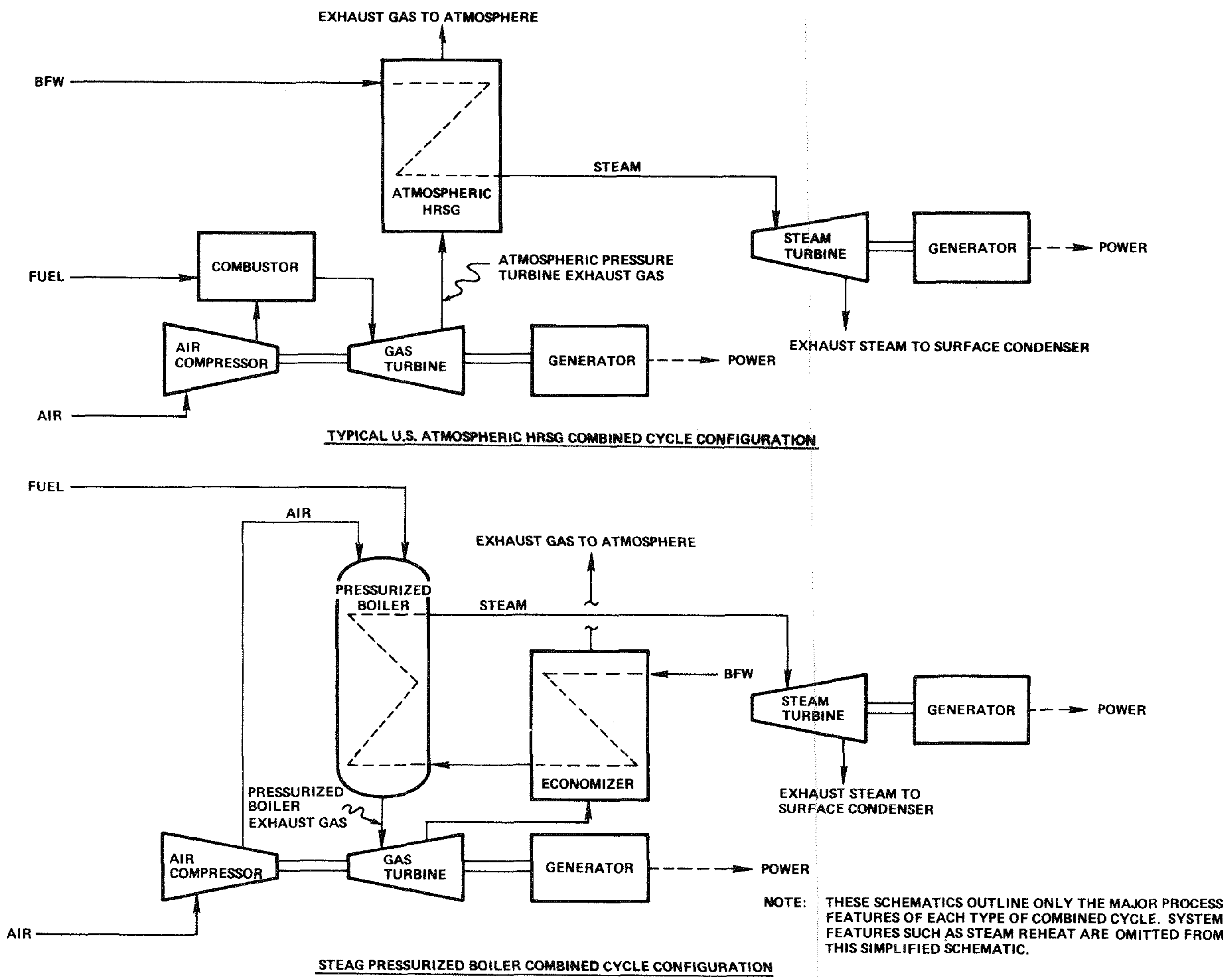


Figure S-1. Schematic Representation of Alternate Combined Cycle Configurations

Economic results are summarized in Table S-1. Operating results are contained in Table S-2. A brief case-by-case summary is as follows:

CASE MASW: AIR-BLOWN LURGI, STEAG COMBINED CYCLE

The overall thermal efficiency for this case was 30.6 percent, based on Western Coal HHV, which is the lowest of the four cases in the study. In terms of absolute dollars, the estimated Total Plant Investment was also the lowest of the four cases. The low Total Plant Investment for Case MASW was offset by the low thermal efficiency to yield a relatively high first-year busbar power cost of 45.4 mills/kWh based on 70 percent capacity factor, coal at $1/MM Btu and 1976 dollars.

STEAG has demonstrated a similar system to Case MASW in their 170 MW unit at Lünen. The combined cycle design in Case MASW utilizes a higher steam generation pressure in the pressurized boiler than that employed in the demonstration unit at Lünen. No major development problems are anticipated for the Case MASW design other than scale-up to larger equipment sizes and demonstration of the higher steam generation pressure in the pressurized boiler.

An improved verson of Case MASW has been proposed by STEAG. This improved design is based on proposed modifications to improve the efficiency of the Lurgi gasification process. STEAG is presently pursuing plans to demonstrate these modifications in a pilot unit.

Based on STEAG's in-house calculations and cost estimating, STEAG claims that the improved version of Case MASW will show a thermal efficiency of 38 percent and an accompanying first-year busbar power cost of 34.2 mills/kWh. Fluor has not checked either the thermal efficiency or power costs quoted by STEAG for this revised version of Case MASW.

CASE EXTS: O_2-BLOWN TEXACO, STEAG COMBINED CYCLE

Heat integration of the oxygen-blown Texaco coal gasification process with the STEAG combined cycle yielded a thermal efficiency of 32.6 percent, based on coal HHV. The accompanying first-year busbar power cost was 43.9 mills/kWh based on 70 percent capacity factor, coal at $1/MM Btu, and 1976 dollars. In case EXTS, saturated steam was generated at 3100 psig in the equipment used for heat recovery from the gasifier effluent. Estimated costs for this equipment were high and a major development program will be required to produce the necessary commercial units. The high cost of this gasifier effluent heat recovery equipment contributed directly to the relatively high Total Plant Investment and accompanying levelized fixed charges.

Improvements in the economics of Case EXTS will require changes in the heat integration design which reduce the cost of the gasifier effluent heat recovery equipment. Fluor has developed a rough design for a revised Case EXTS in which intermediate pressure (595 psig) steam is generated in the gasifier effluent heat recovery equipment. Approximate energy balance calculations and rough cost estimates showed a decrease in thermal efficiency to 31.6 percent and a decrease in first-year busbar power cost to 39.4 mills/kWh.

STEAG has proposed a complete revision of Case EXTS which involves a different steam cycle, a different gas turbine and generation of 1500 psig steam through cooling of the gasifier effluent. All process and cost estimating work for this revision of Case EXTS has been done by STEAG in-house. STEAG has quoted both an overall thermal efficiency of 34.97 percent and a first-year busbar power cost of 37.6 mills/kWh for this new version of Case EXTS. Fluor has not checked these claims for STEAG's revision of Case EXTS.

CASE EATS: AIR-BLOWN TEXACO, STEAG COMBINED CYCLE

Heat integration of the air-blown Texaco coal gasification process with the STEAG combined cycle yielded a thermal efficiency of 33.9 percent, based on coal HHV. The first-year busbar power cost was 50.8 mills/kWh based on 70 percent capacity factor, coal at $1/MM Btu, and 1976 dollars. In this case, superheated steam was generated in the gasifier effluent heat recovery equipment. Estimated costs for this equipment were very high and the development program required to produce commercial units will be extensive. As a result of the high cost of this heat recovery equipment, Case EATS exhibits the highest Total Plant Investment of all four cases in the study.

Improvements in the economics of Case EATS will also require changes in heat integration design which reduce the cost of the gasifier effluent heat recovery equipment. Since the development of the air-blown Texaco gasification process is not as advanced as for the oxygen-blown process, revisions to Case EATS have not been pursued by either Fluor or STEAG at this time.

CASE EXTC78: O_2-BLOWN TEXACO, U.S. COMBINED CYCLE

Integration of the oxygen-blown Texaco coal gasification process with a UTC-supplied U.S. combined cycle yielded an overall thermal efficiency of 34.8 percent, based on coal HHV. The first-year busbar power cost was 38.9 mills/kWh based on 70 percent capacity factor, coal at $1/MM Btu, and 1976 dollars. This heat-integrated GCC system exhibited both the highest thermal efficiency and lowest busbar power costs of all four cases in the study. Since the heat integration design was based on generation of 1500 psig steam through cooling of the gasifier effluent, the estimated cost of the required heat transfer equipment was significantly less than in Cases EXTS and EATS. In addition, the use of a steam generation pressure of only 1500 psig will reduce the development effort required to produce commercial equipment.

Some improvement in the economics of Case EXTC78 could be achieved by reducing the cost of the combined cycle system. Rough estimates suggested that switching to large single shaft gas turbines of approximately 100 MW output could reduce the installed cost of the combined-cycle section by nearly 10 percent. A 10 percent decrease in combined cycle installed cost will reduce the first-year busbar power cost for Case EXTC78 by approximately 1 mill/kWh.

Flowsheet development work now in progress under EPRI sponsorship may further improve the thermal efficiency and lower the cost of services for systems which are very similar to Case EXTC78. The implication of this flowsheet development work is that with further optimization the thermal efficiencies of GCC plants utilizing currently available combustion turbines could possibly be higher than the 34.8 percent presented in this report for Case EXTC78.

The following general conclusions can be drawn from the results of this study:

- Texaco-based GCC system designs tend to show more favorable economics if relatively mild steam generation conditions can be used in the gasifier effluent heat recovery equipment. High steam generation pressures and/or requirements for superheating high-pressure steam to approximately 1000°F are examples of design conditions which can markedly increase the Total Plant Invesment.

- The three specific STEAG-based GCC systems supplied for this study provide no incentive for their development and subsequent use by the U.S. electric power industry. Of these three systems, the oxygen-blown Texaco-based Case EXTS shows the most potential for improvement.

- The final results of this study plus the tentative results of further examinations by both STEAG and Fluor provide no evidence to indicate that the STEAG combined cycle is superior to the U.S. combined cycle in a GCC application.

Each evaluation in this study was performed by using data from process developers to prepare a process design and cost estimate for an integrated grass roots plant. Economic calculations for the cost of services were then made. This information is summarized in Table S-1. The economic results are not directly comparable to those presented in EPRI report AF-642 due to recent revisions in the economic criteria. However, a forthcoming EPRI report will present the cases contained in AF-642 as well as the cases contained in this report in terms of a consistent set of economic criteria.

The plant sizes were selected to match the current utility practice of building base load 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 power in the range of 800 to 1100 megawatts. All cases use Illinois No. 6 coal, except Case MASW, which uses a New Mexico coal. Operating results for this study are summarized in Table S-2.

The technical criteria used in preparing the plant designs are given in Section 2 of this report. Briefly these criteria are:

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

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

- Mid-1976 dollars with no escalation for construction funds
- Thirty-six month construction period
- Eight percent construction loan interest, compounded over the plant construction schedule of three years
- Coal cost of $1/MM Btu (HHV)

- Seventy percent design operating load factor
- Thirty-year plant life
- Fifty:fifty debt:equity ratio
- Eight percent annual bond interest
- Five percent Illinois state sales tax
- Six percent annual inflation rate for calculating 30-year levelized costs

Total capital requirements for each system were determined by adding capital related charges such as preproduction costs, prepaid royalties, initial chemical and catalyst costs, allowance for funds during construction, inventory capital, and land to the estimated plant investments.

Plant investments include a contingency which is divided into two parts. The first of these is a 15 percent project contingency which is intended to cover estimating uncertainty, and additional equipment and material 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. The process contingency is arrived at by applying a separate contingency to individual process units based on their state of development and then accumulating the results.

The Lurgi case (MASW) represents gasification technology that is available today as it has been widely used in commercial scale applications both in Africa and Europe for year.

The Texaco gasification process has been commercially used with heavy petroleum fractions for many years in synthesis gas applications. A demonstration scale plant using coal and oxygen for synthesis gas production has been operating successfully in West Germany since January 1978. Several additional coal-based demonstration plants have been announced. These include an oxygen-blown unit for ammonia synthesis gas for the TVA and an oxygen-blown combined-cycle unit to produce electric power for Southern California Edison at Cool Water, California. While the Texaco process has been commercially demonstrated on petroleum, it has not yet been demonstrated on a large scale with coal. Based on present favorable pilot data and considering the simplicity of the gasifier and its feed system, it is estimated that extension of oxygen-blown gasification to large scale plants should be relatively simple. However, the performance of the oxygen-blown Texaco coal gasification process used in the cases presented in this report is based on gasification data extrapolated to the 1980-1985 time period. Further development work appears necessary to prove this performance. No air-blown Texaco-based demonstration plants have been formally announced at this writing.

The STEAG pressurized boiler combined cycle is currently the only such cycle that has been operated extensively in coupled operation to a coal gasification system to produce electric power. STEAG has operated 170 MW combined cycle plant at Lünen, West Genermany, based on the pressurized boiler combined cycle and air-blown Lurgi moving-bed coal gasifiers. This plant has accumulated over 8600 hours of operation. The STEAG combined-cycle plant has reportedly operated reliably and demonstrated good operational behavior in integrated operation with the coal gasification plant.

The basic U.S. combined-cycle design has been used commercially for many years to produce electric power based on either oil or natural gas-firing of the gas turbines. In recognition of the need to estimate the operational behavior of a total integrated gasfication-combined-cycle system, EPRI is sponsoring a study (RP-913) to develop a dynamic simulation model of a Texaco/combined-cycle plant.

Table S-1

SUMMARY OF ECONOMIC RESULTS

GASIFICATION PROCESS	LURGI	TEXACO	TEXACO	TEXACO
OXIDANT	AIR	O_2	AIR	O_2
COMBINED CYCLE TYPE	STEAG	STEAG	STEAG	U.S.
CASE DESIGNATION	MASW	EXTS	EATS	EXTC78
PRODUCTION AT DESIGN CAPACITY*				
Net Power, MW	865	973	1,013	1,039
Overall Plant Heat Rate, Btu/kWh	11,149	10,479	10,065	9,813
TOTAL CAPITAL**				
Total Capital Requirement, $1000	851,105	949,098	1,194,422	898,425
Total Capital Requirement, $/kW	984	975	1,179	865
AVERAGE COSTS OF SERVICES**				
First-Year Cost, $1000/year	240,933	261,722	315,508	247,629
First-Year Cost, mills/kWh	45.42	43.87	50.79	38.87
Thirty-Year Levelized Cost, $1000/year	321,385	345,121	407,437	325,626
Thirty-Year Levelized Cost, mills/kWh	60.59	57.84	65.59	51.28

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

Table S-2

SUMMARY OF OPERATING RESULTS

GASIFICATION PROCESS OXIDANT COMBINED-CYCLE TYPE CASE DESIGNATION	LURGI AIR STEAG MASW	TEXACO O_2 STEAG EXTS	TEXACO AIR STEAG EATS	TEXACO O_2 U.S. EXTC78
GASIFICATION AND GAS CLEANING SYSTEM				
Coal Feed Rate, lb/hr m.f.	1,014,814	798,333	798,333	798,333
Oxidant/Coal Ratio, lbs/lb m.f.*	1.562	0.858	1.081	0.858
Oxidant Temperature, °F	319	300	1,000	300
Steam/Coal Ratio, lb/lb m.f.##	0.784	0	0.0086	0
Slurry Water/Coal Ratio lb/lb m.f.§	0.0	0.503	0.522	0.503
Gasifier Exit Pressure, psig	336	600	600	600
Crude Gas Temperature, °F	861	2300-2600	2300-2600	2300-2600
Crude Gas HHV (Dry Basis), Btu/SCF**	189.1	281.1	102.5	281.1
Temperature of Fuel Gas to Power System	179	320	142	1,000
POWER SYSTEM				
Gas Turbine Inlet Temperature, °F	1,742	1,742	1,742	1,980
Turbine Compressor Pressure Ratio	9.7:1	11.5:1	11.6:1	13.35:1
Gas Turbine Exhaust Temperature, °F	950	924	893	946
Steam Cycle Conditions, psig/°F/°F (at turbine inlet)	2669/977/977	------	2813/995/977	1450/900/900
Condenser Pressure, inches Hg abs.	2.5	2.5	2.5	2.5
Power Block Stack Temperature, °F	248	253	253	275
Gas Turbine Power, MW#	215	211	205	706
Steam Turbine Power, MW#	717	814	918	331
Fuel Gas Expander Power, MW	0	21	0	43
Power Consumed, MW	67	73	110	41
Net System Power, MW	865	973	1,013	1,039
OVERALL SYSTEM				
Process and Deaerator Makeup Water, gpm/1000 MW, Net	2,750	670	324	804
Cooling Tower Makeup Water, gpm/1000 MW, Net	9,491	11,312	11,588	8,100
Cooling Water Circulation Rate, gpm/MW	389	450	469	374
Cooling Tower Heat Rejection, % of Coal HHV	39.4	50.0	53.3	38.1
Air Cooler Heat Rejection, % of Coal HHV	6.2	3.1	0.0	5.2
Net Heat Rate, Btu/kWh	11,149	10,479	10,065	9,813
Overall System Efficiency (Coal→Power), % of Coal HHV	30.6	32.6	33.9	34.8

*Oxidant air on dry basis for the Lurgi case, or oxygen on 100 percent O_2 basis for the Texaco cases
**Excluding the HHV of H_2S, COS and NH_3
#At generator terminals
##Includes moisture in oxidant air
§Small changes in this ratio do not significantly alter the results presented here
§§Pressurized Boiler Exit Temperature, Cases MASW, EXTS, EATS, Combuster Exit Temperature, Case EXTC78

Section 1

INTRODUCTION AND SCOPE

The study reported here represents a continuation and extension of earlier economic studies done for EPRI by Fluor. The purpose of the new work was to evaluate the economics of producing electricity from coal using both a STEAG pressurized boiler combined cycle and alternately, a typical U.S. atmospheric pressure HRSG combined-cycle design. The pressurized boiler combined cycle is integrated with both oxygen-blown and air-blown Texaco coal gasification processes (Cases EXTS and EATS, respectively) and with an air-blown Lurgi process (Case MASW). The U.S. combined-cycle design (Case EXTC78) is based on modular gas turbines currently in production and available for delivery in 1978. This U.S. combined-cycle design is integrated with an oxygen-blown Texaco gasification process.

The overall results of this comparative evaluation show the economics which result from integration of the STEAG pressurized boiler cycle with the three different gasification processes. These results also provide a comparison of the STEAG pressurized boiler cycle with a typical U.S. combined cycle when each is heat-integrated in a similar manner with an oxygen-blown Texaco gasification process. All of the cases with Texaco gasification employ Illinois No. 6 coal, whereas the air-blown Lurgi case uses a western (New Mexico) coal. Finally, the results show the economics for coal gasification combined-cycle power plants which use current, rather than advanced technology components in the power block designs.

Designs for each of the gasification units, and for the Selexol acid gas removal units were based on information provided by appropriate licensors. The pressurized boiler cycle power systems were designed by STEAG, a major West German electric power company. The U.S. combined-cycle system was designed by United Technologies Corporation (UTC). Plant costs were estimated by Fluor. Economic evaluation criteria were supplied by EPRI.

Cases reported here are reported with a four letter acronym: EXTS, EATS, MASW, EXTC78. These acronyms are defined in Appendix D. Block flow diagrams to indicate the overall plant flow scheme are given for each case . Flow sheets are also provided for individual process units within each plant, where necessary, to depict what is included that is specific to each case.

Section 2

TECHNICAL CRITERIA

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

The Texaco Development Corporation and the American Lurgi Corporation provided EPRI the information used for the designs of the gasification systems. These designs are the same as published in EPRI Report AF-642 (Economic Studies of Coal Gasification Combined-Cycle Systems for Electric Power Generation). The power systems for the STEAG cases were designed by STEAG, whereas the power system for Case EXTC78 was designed by United Technologies Corporation.

The coal analysis is given in Table 2-1. Coal was assumed delivered to the site washed and sized.

The site for the plant is the Chicago area; Table 2-2 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 2-3. This data was extracted from EPRI Report AF-244 (Economics of Current and Advanced Gasification Processes for Fuel Gas Production).

In all cases, net plant products were restricted to electricity, sulfur, and ammonia. No hydrocarbon by-products were allowed. Total plant sulfur emissions are restricted to a maximum of 1.2 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 startup and emergency situations. Because each plant is assumed to be 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
- Fire water
- Flares
- Effluent water treating
- Electrical substation and distribution
- Buildings
- Maintenance
- Laboratory
- Rail
- Road

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

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

Redundant equipment or systems are provided where failure would jeopardize a substantial fraction of plant capacity. Major high cost equipment is not spared where experience indicates minimal probability of failure or where multiple trains are provided which limit the impact of a failure should it occur. In addition, redundancy is not provided where storage permits bypass of equipment for a sufficient period of time to accomplish reasonable maintenance and repair. The sparing provided is noted in the plant description section for each case, and on the flow diagrams. The degree of redundancy is compatible with a 90 percent 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 2-1

COAL ANALYSIS

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		
High Heating Value (HHV) (Btu/lb)	12,235	8,325
Net Heating Value (LHV) (Btu/lb)	11,709	7,869

*Fixed carbon and volatile matter combined

Table 2-2

SITE CONDITIONS

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

Table 2-3

WATER ANALYSIS
(Units are ppm unless otherwise noted)

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

ECONOMIC CRITERIA

A consistent set of criteria for estimating capital requirements and the cost of services was supplied by EPRI. These criteria are summarized in Tables 2-4, 2-5, 2-6, 2-7, and 2-8.

Operating labor requirements were determined after the plant design was completed and the associated costs were computed in accordance with rates shown in Table 2-6. 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 emerge which were not considered during the early stages of the development. 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 separate process contingency is applied to individual process units or equipment based on their state of development and accumulating the results. The process contingency allowances, shown as a percentage of the installed unit costs before any project contingency has been added, are listed in Table 2-8.

Table 2-4

CAPITAL INVESTMENT BASIS

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: • Process plant investment • General facilities costs • Process contingency • Project contingency These items are discussed below:
Process Plant Investment	Total constructed cost of all onsite processing and generating units, including all direct and indirect construction costs and all engineering and home office fees. All sales taxes (5 percent of total materials) are included.
General Facilities Cost	The capital cost of the general (offsite) facilities is explicitly accounted for in the computation of total plant investment. Offsite facilities include roads, buildings, railroad loading and unloading systems, electrical distribution and substations, cooling water systems, inerting systems, instrument air systems, laboratories, effluent water treatment

Item	Basis
	facilities, etc. All sales taxes (5 percent of total materials) and home office fees are included.
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 percent of the sum of the Process Plant Investment and the General Facilities Cost is used.
Total Capital Requirement	The total capital requirement includes all capital necessary to complete the entire project. These capital charges include the following: • Total plant investment • Prepaid royalties • Preproduction costs • Inventory capital • Initial chemical and catalyst charge • Allowance for Funds During Construction (AFDC) • Land These items are discussed below:
Prepaid Royalties	0.5 percent of the Process Plant Investment (excluding contingencies)
Preproduction Costs	The preproduction costs are intended to cover operator training, equipment checkout, major changes in plant equipment, extra maintenance, and inefficient use of coal and other materials during plant startup.

Item	Basis
	The preproduction costs are estimated as follows: • One month of fixed operating costs. (These costs consist of operating and maintenance labor, administrative and support labor, and maintenance materials.) • One month of variable operating costs at full capacity excluding coal. (These costs consist of catalysts, chemicals, and other consumables, water, and ash disposal charges.) • 25 percent of full capacity coal cost for one month. (This charge is to cover inefficient operation during the start-up period.) • 2 percent of Total Plant Investment. (This charge is to cover expected changes and modifications to equipment that will be needed to bring the plant up to full capacity.)
Inventory Capital	The value of inventories of coal and other consumables is capitalized and included in the inventory capital account. Inventory capital is the sum of the following: • Cost of a one-month supply of coal at full capacity operation • Cost of a one-month supply of catalysts, chemicals and other consumables (excluding water) based on full capacity operation.
Initial Catalyst and Chemicals Charge	The initial cost of the charge of catalysts or chemicals contained <u>within</u> the process equipment (but not in storage since this cost is covered in the Inventory Capital) is included.

Item	Basis
Allowance for Funds During Construction (AFDC)	The AFDC charges are computed from the Total Plant Investment (TPI) as follows:

$$\text{AFDC} = (0.1249) \times (\text{TPI})$$

The AFDC charge factor of 0.1249 is based on an annual interest rate of 8 percent with annual end-of-year compounding and the following construction expenditure schedule:

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

Expenditures in a given year are assumed to be uniform over that year. Since the AFDC charge is expressed in the same year dollars as the Total Plant Investment, cost escalation (inflation) is not included.

Land	Land costs are estimated at $5,000/acre.

Table 2-5

CAPACITY FACTOR

Item	Basis
Design Capacity Factor	A design capacity factor (operating load factor) of 70 percent is used in this study. The design capacity factor is assumed to be constant over the life of the plant. The plant capacity factor is the percent of the time the plant is on-line at baseload (100 percent plant design power output). This is not equivalent to plant operation in a turndown mode producing less than the baseload power output.
Capacity Factor Range	A capacity factor range of 60 to 90 percent is used to present two "Power Cost Sensitivity Curves."

Table 2-6

OPERATING COST BASIS

Item	Basis
Operating Costs	The operating costs are estimated and reported on a first-year basis. These same costs are also computed and reported on a 30-year levelized basis. Operating costs are divided into fixed and variable costs. Descriptions of these cost categories are found below:
Fixed Operating Costs	The fixed costs are essentially independent of the plant capacity factor and are composed of the following charges: • Operating labor • Maintenance • Overhead charges These items are discussed below:
Operating Labor	The operating labor charge (OLC) is computed using an average labor rate (ALR) of \$12.50/person hour. This ALR is based on a direct labor charge of \$9.25/person hour plus a 35 percent payroll burden. The formula for calculating the first-year OLC in \$/yr is as follows: OLC = (OJ) x (ALR) x (8760 hr/yr) where OJ is the number of operating jobs.
Maintenance Costs	Annual maintenance costs are estimated as a percentage of the installed capital

Item	Basis
	cost of the facilities. The percentage applied varies with the nature of the processing conditions and the type of design. The following percentages are used in all four cases:

Process Unit	Maintenance % of Installed Plant Section Cost/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 Expansion % Air Compression	3.0
Process Condensate Treating	3.0
Steam, Condensate & BFW	1.5
Support Facilities	1.5
Combined Cycle	1.5
Steag Cycle	2.0

Item	Basis
	The maintenance costs are divided into maintenance labor and maintenance materials. A maintenance labor/materials ratio of 40/60 is used.
Overhead Charges	The only overhead charge included in the fixed costs is a charge for administrative and support labor. This overhead charge is 30 percent of the sum of the operating and maintenance labor.

Item	Basis
Variable Operating Costs	The variable operating costs are dependent upon the plant capacity factor and are composed of the following charges: • Coal • Raw water • Catalysts and chemicals • Ash disposal These items are discussed below:
Coal	The first-year coal cost is \$1/MM Btu (HHV).
Raw Water	The first-year raw water acquisition cost is 40¢/1000 gallons. Treating costs and pumping costs are included in the operating and maintenance charges.
Catalysts and Chemicals	The first-year catalyst, chemicals and other consumable costs are based on pricing data from vendors.
Ash Disposal	The first-year ash disposal charge is \$4.00/ton.
Levelized Operating Costs	Inflation will tend to increase the operating costs (in current dollars) over the life of the plant. In this study, a long-term inflation rate of 6%/yr is assumed in estimating both the Cost of Capital and the life cycle revenue requirements for other expenses. The "present worth" concept of money is used to compute a single "levelized" value for the fixed and variable operating costs which represents the varying revenue requirements.

Item	Basis
	Using the following assumptions, Inflation rate = 6%/yr Discount rate = 10%/yr The computed 30-year levelization factor (LF) for both the fixed and variable operating and maintenance (O&M) costs, excluding coal cost, is 1.886. The computed 30-year levelized O&M cost, excluding coal, is as follows: 30-year levelized O&M = 1.886 x (1st year O&M) A 6.2 percent coal price escalation rate is used in this study. The resulting 30-year LF for coal cost is 1.932. The computed 30-year levelized coal cost (CC) is as follows: 30-year levelized CC = 1.932 x (1st year CC) Both first year and 30-year levelized operating costs are developed and reported in the economic analyses for each case.
By-product Credits	No credit is taken for by-product sulfur. The first-year credit for by-product ammonia is $100/ton.

Table 2-7

COST OF CAPITAL

Item	Basis
Basic Assumptions	The following assumptions form the basis for calculation of the Cost of Capital.

Inflation Rate	6%/yr
Debt/Equity Ratio	50/50
Debt Cost	8%/yr
Preferred Stock Ratio	15%
Preferred Stock Cost	8.5%/yr
Common Stock Ratio	35%
Common Stock Cost	13.5%/yr
Weighted Cost of Capital	10%/yr
Federal and State Income Tax Rate	50%
Property Taxes and Insurance	2%/yr
Investment Tax Credit	0
Book Life	30 yr
Tax Life	20 yr
Iowa Type S Retirement Dispersion	

Levelized Fixed Charges (30-year plant)	At a discount rate of 10%/yr, the 30-year levelized Fixed Charge Rate (FCR) based on the above assumptions is 18%/yr. The Levelized Fixed Charges (LFC) in \$/yr for a 30-year plant are calculated from the Total Capital Requirement (TCR) as follows:

$$LFC = (0.18) \times (TCR)$$

Table 2-8

PROCESS CONTINGENCIES

Case	EXTC 78 (%)	EXTS (%)	EATS (%)	MASW (%)
Coal Handling	0	0	0	0
Oxidant Feed	0	0	0	0
Gasification	15	15	15	5
Ash Handling	5	5	5	5
Gas Cooling	0-15*	0-30#	0-30#	0
Acid Gas Removal	0	0	0	5
Sulfur Recovery (Claus)	0	0	0	0
Tail Gas Treating	15	15	15	15
Process Condensate Treatment, Steam, Condensate and BFW	0	0	0	0-15**
Support Facilities	0	0	0	0
STEAG Combined Cycle	-	5##	5##	5##
U.S. Combined Cycle	0	-	-	-

*For Case EXTC78, 15 percent is applied to high-temperature waste heat boilers in the raw gas cooling section, 0 percent is applied to remaining low-temperature gas cooling equipment

**For Case MASW, 15 percent is applied to the by-products boiler, 0 percent is applied to all other equipment

#For Cases EXTS and EATS, 30 percent is applied to waste heat boilers in the raw gas cooling sections because of the 3100 psig steam generation pressure, 0 percent is applied to remaining low-temperature gas cooling equipment

##A five percent process contingency is applied to the pressurized boiler combined cycle because the 2814 psig high-pressure steam level assumed in this report is somewhat higher than the 1850 psig level used in the initial demonstration plant at Lünen

Section 3

DISCUSSION OF RESULTS

This report contains the results of a "screening type" comparative evaluation. Fluor did not perform a comprehensive analysis of the four individual GCC systems. Instead, the overall system designs contained herein were developed around the power block configurations and performance information supplied by STEAG and UTC. Within the budget and time limitations which applied to this study, it was not feasible for Fluor, STEAG or UTC to fully optimize these designs. Except for the Texaco gasification processes and the high temperature sections of their accompanying heat recovery systems, the four GCC systems are based on currently available equipment with a few extensions to sizes which are larger than the current state-of-the-art. The type of approach taken in this study often defines both the need to investigate additional changes in system configuration and the incentive for development of certain process equipment. With the foregoing comments in mind, the reader should guard against assuming that all comparisons contained in this study are either complete or final.

In performing process evaluations of relatively new or unfamiliar technology, a tendency exists for plant cost estimates to be somewhat optimistic. This tendency is always a hazard where full and complete mechanical definitions cannot be developed for each item in the plant. In an attempt to offset this tendency, we have applied both a process contingency and a project contingency to the plant cost estimates. These contingencies are 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. The accuracy of the plant investment estimates is judged to be ±25 percent.

The Lurgi and oxygen-blown Texaco gasification processes presented in this report are basically the same as presented in EPRI Report AF-642. However, the details of waste heat recovery and utilization have been changed to meet the requirements of the STEAG and the UTC combined-cycle generation systems. Similarly, the air-blown Texaco process is basically the same as presented in EPRI Report AF-753 (Economics of Texaco Gasification Combined-Cycle Systems), but with the waste heat utilization flowscheme changed to meet the requirements of the STEAG combined-cycle power generation system. As in the earlier reports, the material balances for the Texaco gasifiers, including feed coal slurry concentrations and gasifier oxygen consumption, are based on a Texaco extrapolation of the state-of-the-art to a period several years hence.

PERFORMANCE COMPARISONS

Table S-2 in the Summary section lists the major operating parameters for all four GCC systems. Final choices of some of the parameters in the combined-cycle designs were left to the discretion of STEAG and UTC. The major combined-cycle design criterion specified by EPRI for this study was the use of "commercially available" power generation equipment. Another general requirement in each of the four GCC plant designs was the utilization of configurations for gasification-power block heat integration which were directly analogous to those employed in previous studies as described in EPRI Reports AF-642 and AF-753.

As a result of the split responsibility for development of the combined-cycle designs, STEAG and UTC have adopted different specifications for certain important operating parameters. Examples of these important parameters are the gas turbine "combustor exit" temperature and the flue gas stack exit temperature. The differences in these parameters must be recognized when making side-by-side comparisons of the STEAG and U.S. combined-cycle-based systems contained in this study. Important aspects of the design and performance of each system are discussed below.

Case MASW: Air-Blown Lurgi, STEAG Combined Cycle

This case contains a combined cycle based on pressurized boilers as designed by STEAG which is heat integrated with a commercial air-blown Lurgi-based gasification process designed by Fluor. All equipment in both the gasification process and the pressurized boiler combined cycle represents current technology. The coal feed differs from the Texaco-based cases in that it is a Western U.S. coal rather than an Illinois No. 6. The Lurgi process is considered to be particularly suitable for gasification of Western U.S. coals.

Operating results for Case MASW previously presented in Table S-2 show that the power produced by gas turbines is nearly the same as that obtained in each of the two Texaco-based cases which utilize the STEAG combined cycle. However, the power produced by the steam turbine in Case MASW is significantly lower than in the Texaco-based cases. A design feature of the Texaco-based cases is the generation of high-pressure steam for use in the steam turbine cycle by cooling of the gasifier effluent. This high-pressure steam generated by cooling of the Texaco gasifier effluent supplements the high-pressure steam produced in the STEAG combined cycle and thereby increases the total steam turbine power output. The same ability to generate supplemental high-pressure steam does not exist in the Lurgi gasification process employed for Case MASW. As a result, Case MASW exhibits the lowest net power output (865 MW) and the lowest overall thermal efficiency (30.6 percent) of the three GCC systems based on the STEAG combined cycle.

Case EXTS: O_2-Blown Texaco, STEAG Combined Cycle

This case contains a STEAG pressurized boiler combined cycle which is heat-integrated with an oxygen-blown Texaco gasification process. As noted in the discussion of Case MASW, supplemental high-pressure saturated steam is generated by cooling of the Texaco gasifier effluent. This supplemental high-pressure steam is sent to the STEAG combined cycle where it is mixed with the larger flow of high-pressure steam produced in the pressurized boiler. The combined stream of saturated steam then flows to the superheating coil of the pressurized boiler. This method of integrating the gasification area and combined-cycle high-pressure steam systems is analogous to the approach used in U.S. combined-cycle-based systems. Specifically, the pressurized boiler in the STEAG combined cycle serves the same role in the integration of the high-pressure steam systems as that served by the HRSG in the U.S. combined cycle.

As shown in Table S-2, the bulk of the net power output from the STEAG combined cycle is produced by the steam turbine. For reasons of cycle efficiency, STEAG has based their design on a steam generation pressure level of approximately 3100 psig. The steam pressure at the outlet of the pressurized boiler superheater coil is 2814 psig. Since the saturated steam generated by cooling of the Texaco gasifier effluent is also sent to the pressurized boiler superheater, the required heat transfer equipment must operate with a steam generation pressure of 3100 psig. Such equipment is presently outside the realm of current technology for energy recovery from hot, particulate-bearing gas streams. An extensive development effort will be required to provide commercial heat transfer equipment which can handle both the high steam generation pressure and particulate-bearing gas on the process side. Estimates prepared for this study suggest that the equipment which is ultimately developed for Case EXTS-type service will be very expensive.

Case EXTS exhibits the second highest thermal efficiency (32.6 percent) and the lowest "first-year" cost of electricity (43.9 mills/kWh) of the three GCC systems which employ the STEAG combined cycle. A subsequent section of this chapter contains further discussion of the economic impact of the chosen method of heat integration in Case EXTS.

Case EATS: Air-Blown Texaco, STEAG Combined Cycle

This case contains a STEAG pressurized boiler combined cycle which is heat-integrated with an air-blown Texaco gasification process. The method of integration of the gasification area and combined-cycle high-pressure steam systems is directly analogous to the approach used in the U.S. combined-cycle-based study found in EPRI report AF-753. Supplemental superheated high-pressure steam is generated by cooling of the air-blown Texaco gasifier effluent. This supplemental superheated high-pressure steam is sent to the STEAG combined cycle where it is combined with the superheated steam produced in the pressurized boiler. The combined stream of superheated steam then flows to the steam turbine.

Due to the need for the combined capability to generate and superheat steam at 3100 psig, the heat transfer equipment required for gasifier effluent cooling in Case EATS is far outside the realm of present technology. The development effort to provide commercial equipment for this service is expected to be both extensive and costly. Present estimates indicate that the resulting heat transfer equipment will be even more expensive than the units required for Case EXTS. Furthermore, as discussed in EPRI Report AF-753, the state of development of air-blown Texaco gasification process is less advanced than that of the oxygen-blown process.

The overall thermal efficiency of Case EATS is the highest of the three GCC systems based on the STEAG combined cycle. However, the accompanying high plant investment for this case more than offsets the efficiency advantage and results in a "first-year" cost of power which is the highest (50.8 mills/kWh) of all four GCC systems in the study.

Case EXTC78: O_2-Blown Texaco, U.S. Combined Cycle

This case utilizes a U.S. combined-cycle configuration as designed by UTC for heat-integration with an oxygen-blown Texaco gasification process. In contrast to the STEAG design, the greater portion of the net power output from the U.S. combined cycle is produced by the gas turbines. UTC has employed current production model gas turbines with a 1980°F combustor exit temperature. This combustor exit temperature (upstream of the turbine) is 238°F higher than the temperature of the pressurized boiler exhaust (upstream of the turbine) used in each of the three STEAG-based cases. As discussed previously, this difference in temperature is a result of allowing both STEAG and UTC to independently specify certain operating parameters in the course of their design work.

The heat-integration design employed by UTC is based on generating saturated high-pressure steam by cooling the Texaco gasifier effluent. This supplemental saturated high-pressure steam is subsequently superheated in the combined-cycle HRSG and sent to the steam turbine. UTC has based their steam cycle design on a turbine with a 1450 psig inlet pressure at the high-pressure end. Steam generation at approximately 1500 psig is therefore required in the heat transfer equipment used to cool the gasifier effluent. Less development effort will be required to provide this equipment as compared to the units in the STEAG-based cases. In particular, generation of steam at 1500 psig is within present mechanical design capabilities for such heat recovery equipment. The remaining development problems are those associated with handling hot, particulate-bearing gasifier effluent.

Case EXTC78 exhibits the highest overall thermal efficiency (34.8 percent) of the four GCC systems. In Total Capital Requirement, Case EXTC78 is second lowest to the low efficiency, Lurgi-based Case MASW. As a net result of these cost and efficiency factors, Case EXTC78 yields the lowest cost of power of all four systems in the study (38.9 mills/kWh).

ECONOMIC COMPARISONS

Comparisons of Total Capital Requirements for all four study cases are presented in Table 3-1, and on a $/kW basis in Table 3-2. A percentage breakdown of the Total Plant Investment on a plant section basis is contained in Table 3-3 for each case. Cost of services comparisons are found in Table 3-4.

Reference to Table 3-1 shows that Case MASW has the lowest Total Capital Requirement in total dollars. This GCC system couples the Lurgi gasification process with the STEAG combined cycle. However, as a result of the low thermal efficiency and accompanying low power output, the comparision in Table 3-2 shows that the Total Capital Requirement for Case MASW becomes the second highest of the four cases on a $/kW basis. The adverse impact of the low thermal efficiency in Case MASW carries through to the cost of electricity presented in Table 3-4.

Cases EXTS and EATS, respectively, couple oxygen-blown and air-blown Texaco gasification processes to the STEAG pressurized boiler combined cycle. Table 3-3 shows that the cost of the gas cooling section constitutes a major fraction of the relatively high Total Plant Investment for each of these cases. In the air-blown, Texaco-based Case EATS, the gas cooling section is estimated to be the single most expensive system in the design. The total cost of the gas cooling section in both Case EXTS and Case EATS is dominated by the cost of the heat transfer equipment located immediately downstream of the gasifier. As discussed previously, the requirement for generation of either saturated or superheated steam at approximately 3100 psig places this heat recovery equipment well beyond the present state-of-the-art for commercially available units. The highly developmental nature of this equipment mandates application of a significant process contingency to the estimated cost. We have used a 30 percent process contingency for this equipment. The process contingencies used in this study are summarized in Table 2-8.

Table 3-3 also shows that the cost of the gas cooling section in the U.S. combined-cycle-based Case EXTC78 constitutes a much lower percentage of the Total Plant Investment than in Cases EXTS and EATS. Although the heat recovery equipment required for gasifier effluent cooling in Case EXTC78 is also developmental in nature, the requirement for generation of saturated steam at only 1500 psig leads to a lower estimated cost. In addition, the fact that this equipment is closer to the present designs for commercial units than the equipment in Cases EXTS and EATS permits the use of a lower process contingency. We have used a 15 percent process contingency for this equipment in Case EXTC78.

A side-by-side summary comparison of the estimated total gas cooling section costs and the accompanying process contingencies is shown below:

Case	Heat Recovery Steam Conditions	Gas Cooling Section Estimated Cost $1000**	Gas Cooling Section Process Contingency $1000
EXTS	3100 psig/Sat'd	139,562	32,524
EATS	2813 psig/995°F*	302,785	75,057
EXTC78	1500 psig/Sat'd	60,688	3,730

*Superheated steam

**Without process or project contingencies

It is obvious from the above comparison that reduction of the cost of the gas cooling section in Cases EXTS and EATS could lead to improvement in the economics for these two STEAG combined-cycle-based systems. Such reductions in cost must be accomplished without accompanying significant reductions in the overall thermal efficiency of these cases. The above comparison illustrates the strong impact of steam conditions on the cost of the heat recovery equipment used for gasifier effluent cooling. Since these steam conditions are the direct result of the method of heat integration with the combined-cycle systems, gas cooling section cost reductions will be linked to changes in this area of the GCC plant design.

A side-by-side comparison of the combined-cycle section costs for the three Texaco-based cases is presented below:

Case	Combined-Cycle Type	Power Generation Section Estimated Cost $1000*
EXTS	STEAG	268,614
EATS	STEAG	267,053
EXTC78	U.S.	342,213

*Without process or project contingencies

The foregoing comparison indicates that a GCC plant power generation section based on the STEAG combined cycle is less expensive than one based on the U.S. combined cycle. Some of the above-noted difference in cost is probably derived from the relatively compact nature of the STEAG combined-cycle components. In particular, fuel combustion at elevated pressure plus steam generation at approximately 3100 psig are factors which contribute to the comparatively small size of the pressurized boiler. However, it sould also be noted that the U.S. combined-cycle design provided by UTC for Case EXTC78 is based on 24 parallel gas turbines of approximately 30 MW and 12 parallel HRSG units. Total installed costs tend to rise with an increase in number of parallel items of equipment in a system. A U.S. combined-cycle design based on fewer gas turbines of higher output per unit will probably be less costly than the Case EXTC78 power generation system. Single gas turbines with an output of approximately 100 MW and an accompanying combustor exit temperature of approximately 2000°F are presently available from other U.S. power generation equipment vendors.

POTENTIAL IMPROVEMENTS

The following is a discussion of some possible improvements in the designs presented in this report:

Case MASW: Air-Blown Lurgi, STEAG Combined Cycle

Case MASW is based on system designs which utilize commercially available technologies and components. Therefore, improvements in Case MASW must necessarily involve design changes which increase the overall thermal efficiency as opposed to reducing the cost of or eliminating developmental-stage equipment. As noted in the previous discussion, the gas turbine specified by STEAG for Case MASW operates with a combustor exit (pressurized boiler exhaust) temperature which is 238°F lower than that employed in Case EXTC78. It is probable that the thermal efficiency of Case MASW would be improved by redesign of the STEAG combined cycle to provide for both a pressurized boiler exhaust temperature of 1980°F and concurrent use of a gas turbine which is similar in performance to the unit specified by UTC for Case EXTC78. At this time, Fluor has not attempted a redesign effort to define the impact of such a change on performance and economics.

STEAG is presently working on a new version of the air-blown Lurgi coal gasification process. This new version includes innovations such as recycle of both tar and CO_2 to the gasifier and utilization of a waste heat boiler on the gasifier effluent upstream of the present water quench system. Recycling of tar to the gasifier would eliminate the tar boiler unit now contained in the present Case MASW design. STEAG has successfully demonstrated the tar recycle modification in their unit at Lünen. Recycle of CO_2 to the gasifier would involve replacement of the present acid gas removal system in the Case MASW design with a combination of hot potassium carbonate and Stretford units plus a CO_2 recycle compressor. A design is now being prepared by STEAG for a pilot unit in Lünen which will test the CO_2 recycle and waste heat boiler modifications.

Based on in-house flowsheet development and cost estimating, STEAG claims that a GCC plant based on the new verson of the Lurgi process and their combined-cycle system will show a 38 percent thermal efficiency and an accompanying first-year cost of power of 34.2 mills/kWh. It should be noted that Fluor has not performed any checks on these STEAG estimates of efficiency and cost for their new Lurgi-based GCC system. Since STEAG has demonstrated an ability to successfully develop Lurgi gasification process modifications such as the tar recycle, progress in piloting the above-described additional changes should be monitored with a high degree of interest.

Case EXTS: O_2-Blown Texaco, STEAG Combined Cycle

Based on the factors discussed in the economic comparisons, improvements in Case EXTS must focus on changes in the heat integration design that lead to a reduced cost for the gas cooling section of the plant. Fluor has developed a rough design for a version of Case EXTS in which saturated, intermediate-pressure (595 psig) steam is generated by gasifier effluent cooling and subsequently superheated in the pressurized boiler reheating coil along with the exhaust steam from the high-pressure end of the steam turbine. An approximate energy balance shows an overall decrease in power output of 29 MW and an accompanying decrease in the overall thermal efficiency to 31.6 percent. Rough cost estimates indicate that the cost of the gas cooling section in this version of Case EXTS will decrease to approximately $50 million prior to the addition of contingencies. The estimated reduction in gas cooling section cost plus a minor overall decrease in cost of the STEAG combined cycle lowers the first-year power cost by 4.5 mills/kWh. This revised version of Case EXTS appears to become more competitive with Case EXTC78 in terms of power cost and less competitive on basis of thermal efficiency. It is to be emphasized that this version of Case EXTS has been developed to indicate one possible approach to further improvement in system economics. The design and cost estimates have not been prepared with the same degree of detail that characterizes the four fully-described cases in this report.

STEAG has also recognized the heat integration related problems in Case EXTS and has subsequently developed their own revised version of this GCC system. This STEAG-developed version reportedly features both a gas turbine which operates with a pressurized boiler exhaust temperature of 1895°F and a totally revised steam system in the combined cycle. Heat integration of the steam cycle with the gasification area is apparently accomplished by steam generation at approximately 1500 psig in the heat recovery equipment on the Texaco gasifier effluent. STEAG has quoted both an overall thermal efficiency of 34.97 percent and a first year power cost of 37.6 mills/kWh for this system. Fluor has not seen detailed flowsheets or cost estimates for this STEAG-developed revision of Case EXTS and has, therefore, performed no independent checks of either the thermal efficiency or the accompanying power cost.

As noted previously, definition of important areas for design optimization is often obtained by preparation of flowsheets and cost estimates of the type contained in this report. Case EXTS serves as an illustration of this point. Alternate versions of Case EXTS developed by both Fluor and STEAG indicate that several options may exist for improvement of this system.

Case EATS: Air-Blown Texaco, STEAG Combined Cycle

The economics of Case EATS could probably be improved by adoption of the same type of changes proposed for Case EXTS. It was noted in EPRI Report AF-753 that steam generation plus superheating in the gasifier effluent heat recovery equipment is necessary to achieve closure of the overall steam balance in an air-blown, Texaco-based GCC system. Reduction of the steam pressure level to 1500 psig or less in this heat recovery equipment would have a marked beneficial effect in terms of reducing the Case EATS gas cooling section cost.

Since the development of the air-blown Texaco gasification process is not as advanced as the development of the O_2-blown process, modifications to Case EATS have not been addressed by either Fluor or STEAG at this time.

Case EXTC78: O_2-Blown Texaco, U.S. Combined Cycle

Economics for Case EXTC78 can be improved by reducing the cost of the combined-cycle section of the plant. The use of large, single shaft gas turbines of approximately 100 MW output each would reduce the required number of gas turbines and accompanying HRSG's to seven. Based on the cost estimates prepared for the power generation section of Case EXTC (slurry feed) in EPRI Report AF-642, it is possible that such a reduction in the required number of gas turbines and HRSG's in the Case EXTC78 combined cycle could decrease the installed cost of this plant section by approximately 10 percent. A 10 percent decrease in the Case EXTC78 combined-cycle cost will reduce the first-year cost of power by approximately 1 mill/kWh.

EPRI is currently sponsoring additional studies of Texaco-based GCC systems which employ commercially available gas turbines (2000°F combustor exit temperature) in the combined-cycle designs. The focus of these additional studies is to determine the impact of various heat integration design options on the overall thermal efficiency. Some cases in this ongoing study of heat integration design options are expected to yield thermal efficiencies closer to 36 percent as opposed to the 34.8 percent obtained for Case EXTC78. Future economic evaluations will incorporate the design refinements identified by this heat integration study. Since many of these heat integration refinements appear to be specific to the gas cooling section of a Texaco-based plant, incorporation of the indicated design changes in both STEAG combined-cycle-based and U.S. combined-cycle-based GCC systems will probably not alter the relative cost and performance ranking indicated for the four cases contained in this study.

GENERAL CONCLUSIONS

The following general conclusions can be stated by considering the results of both this study and past Fluor evaluations such as those contained in EPRI Reports AF-642 and AF-753.

- The overall thermal efficiency of a Texaco-based GCC system tends to be improved by heat integration designs which directly couple the high-pressure steam systems of the gasification and combined-cycle sections of the plant.

- In terms of cost, heat-integrated, Texaco-based GCC systems tend to show more favorable economics if relatively mild steam conditions can be used in the gasifier effluent heat recovery equipment. High steam generation pressures and/or requirements for superheating high-pressure steam to approximately 1000°F are examples of design conditions which can markedly increase the cost of the gas cooling section of a Texaco-based plant.

- Design of certain Texaco-based GCC systems may involve some compromises between the highest obtainable thermal efficiency and acceptable costs. The steam-cycle intensive STEAG combined cycle is an example of a system where significant cost versus efficiency compromises may be required for integration with the Texaco gasification process.

- The three STEAG-based GCC systems supplied for this study provide no incentive for their development and subsequent use by the U.S. electric power industry. Of these three systems, the O_2-blown Texaco-based Case EXTS shows the most promise for possible improvements in terms of power costs.

- The final results of this study plus the tentative results of further examinations by both STEAG and Fluor provide no evidence to indicate that the STEAG combined cycle is superior to the U.S. combined cycle.

- Of the various refinements suggested as ways to improve the three STEAG-based GCC systems, the proposed modifications to the air-blown Lurgi coal gasification process are probably the most interesting. Since the Lurgi process has already been proven in commercial applications, STEAG's progress in demonstrating their proposed modifications should be followed.

Table 3-1

SUMMARY OF CAPITAL INVESTMENT ESTIMATES

Gasification Process	Lurgi	Texaco	Texaco	Texaco
Oxidant	Air	O_2	Air	O_2
Combined-Cycle Type	STEAG	STEAG	STEAG	U.S.
Case Designation	MASW	EXTS	EATS	EXTC78
PLANT INVESTMENT - $1000				
Coal Handling	26,039	22,591	23,040	22,591
Oxidant Feed	20,544	118,190	46,555	119,400
Gasification and Ash Handling	103,269	25,087	61,901	25,087
Gas Cooling	20,765	139,562	302,785	60,688
Acid Gas Removal and Sulfur Recovery	39,204	29,963	44,051	29,337
Lurgi Process Condensate Treating	60,213	--	--	--
Steam, Condensate and BFW	37,078*	1,347	2,382	2,021
Combined Cycle	244,946	268,614	267,053	342,213
General Facilities	60,542	59,371	59,378	61,966
Process Contingencies	24,092	50,083	98,333	7,858
Project Contingencies	91,890	99,709	121,072	99,495
TOTAL PLANT INVESTMENT	728,582	814,520	1,026,550	770,656
CAPITAL CHARGES				
Prepaid Royalties	2,760	3,024	3,739	3,007
Preproduction Costs	19,203	20,625	25,668	19,317
Inventory Capital	7,088	7,481	7,487	7,476
Initial Catalyst and Chemical Charges	1,472	714	1,762	714
Allowance for Funds During Construction	91,000	101,734	128,216	96,255
Land	1,000	1,000	1,000	1,000
TOTAL CAPITAL CHARGES	122,523	134,578	167,872	127,769
TOTAL CAPITAL REQUIREMENTS	851,105	949,098	1,194,422	898,425

*Cost includes tar boiler

Table 3-2

SUMMARY OF CAPITAL INVESTMENT ESTIMATES - $/KW

Gasification Process	Lurgi	Texaco	Texaco	Texaco
Oxidant	Air	O_2	Air	O_2
Combined-Cycle Type	STEAG	STEAG	STEAG	U.S.
Case Designation	MASW	EXTS	EATS	EXTC78
NET POWER PRODUCTION - MW	865.00	973.00	1,013.00	1,039.00
PLANT INVESTMENT - $/kW				
Coal Handling	30.10	23.22	22.74	21.74
Oxidant Feed	23.75	121.47	45.96	114.92
Gasification and Ash Handling	119.39	25.78	61.11	24.14
Gas Cooling	24.01	143.44	298.90	58.41
Acid Gas Removal and Sulfur Recovery	45.32	30.79	43.48	28.24
Lurgi Process Condensate Treating	69.61	--	--	--
Steam, Condensate, and BFW	42.86*	1.38	2.35	1.95
Combined Cycle	283.18	276.07	263.63	329.37
General Facilities	69.99	61.02	58.62	59.64
Process Contingencies	27.85	51.47	97.07	7.56
Project Contingencies	106.23	102.48	119.52	95.76
TOTAL PLANT INVESTMENT	842.29	837.12	1,013.38	741.73
CAPITAL CHARGES				
Prepaid Royalties	3.19	3.11	3.69	2.89
Preproduction Costs	22.20	21.20	25.34	18.59
Inventory Capital	8.19	7.69	7.39	7.20
Initial Catalyst and Chemical Charges	1.70	0.73	1.74	0.69
Allowance for Funds During Construction	105.20	104.55	126.57	92.64
Land	1.16	1.03	0.99	0.96
TOTAL CAPITAL CHARGES	141.64	138.31	165.72	122.97
TOTAL CAPITAL REQUIREMENTS	983.93	975.43	1,179.10	864.70

*Cost includes tar boiler

Table 3-3

PERCENTAGE CONTRIBUTION OF PLANT SUBSECTIONS
INVESTMENT TO TOTAL PLANT INVESTMENT

Gasification Process	Lurgi	Texaco	Texaco	Texaco
Oxidant	Air	O_2	Air	O_2
Combined-Cycle Type	STEAG	STEAG	STEAG	U.S.
Case Designation	MASW	EXTS	EATS	EXTC78
PLANT SUBSECTION INVESTMENT (including contingencies)				
Coal Handling	4.11	3.19	2.57	3.37
Oxidant Feed	3.24	16.68	5.22	17.82
Gasification	17.01	3.97	7.81	4.19
Gas Cooling	3.28	23.70	41.23	9.54
Acid Gas Removal and Sulfur Recovery	6.48	4.32	5.03	4.46
Lurgi Process Condensate Treating	9.50	--	--	--
Steam, Condensate, and BFW	6.48*	0.19	0.27	0.30
Combined Cycle	40.34	39.57	31.22	51.07
Support Facilities	9.56	8.38	6.65	9.25
TOTAL PLANT INVESTMENT	100.00	100.00	100.00	100.00

*Cost includes tar boiler

Table 3-4

SUMMARY OF BUSBAR POWER COSTS AT 70 PERCENT CAPACITY FACTOR

Gasification Process	Lurgi		Texaco		Texaco		Texaco	
Oxidant	Air		O_2		Air		O_2	
Combined-Cycle Type	STEAG		STEAG		STEAG		U.S.	
Case Designation	MASW		EXTS		EATS		EXTC78	
NET PRODUCTION*								
Net Power	865		973		1,013		1,039	
By-product Ammonia, ST/D	128		-0-		-0-		-0-	
By-product Sulfur, ST/D	63		301		309		301	
TOTAL CAPITAL REQUIREMENT, $1000/YEAR	851,105		949,098		1,194,422		898,425	
	First Year Cost	30 Year Levelized Cost	First Year Cost	30 Year Levelized Cost	First Year Cost	30 Year Levelized Cost	First Year Cost	30 Year Levelized Cost
FIXED OPERATING COST, $1000/YEAR								
Operating Labor	3,942	7,435	3,066	5,782	3,395	6,403	3,066	5,783
Maintenance Labor	7,405	13,966	7,599	14,331	10,885	20,529	5,995	11,307
Maintenance Materials	11,108	20,950	11,398	21,497	16,328	30,795	8,992	16,959
Administrative and Support Labor	3,404	6,420	3,199	6,034	4,284	8,079	2,718	5,127
Total Fixed O&M Costs	25,859	48,771	25,262	47,644	34,892	65,806	20,771	39,176
VARIABLE OPERATING COSTS (EXCLUDING COAL), $1000/YEAR								
Raw Water	1,779	3,355	1,801	3,397	1,753	3,306	1,361	2,568
Catalysts and Chemicals	408	769	319	602	365	689	279	526
Ash Disposal	3,826	7,216	981	1,850	981	1,850	981	1,850
Total Variable O&M Costs	6,013	11,340	3,101	5,849	3,099	5,845	2,621	4,944
COAL COST, $1000/YEAR	59,132	114,243	62,521	120,790	62,521	120,790	62,521	120,790
BY-PRODUCT CREDITS, $1000/YEAR								
By-product Ammonia	(3,270)	(6,168)	-0-	-0-	-0-	-0-	-0-	-0-
By-product Sulfur	-0-	-0-	-0-	-0-	-0-	-0-	-0-	-0-
Total By-product Credits	(3,270)	(6,168)	-0-	-0-	-0-	-0-	-0-	-0-
TOTAL OPERATING COSTS, $1000/YEAR	87,734	168,186	90,884	174,283	100,512	192,441	85,913	164,910
LEVELIZED FIXED CHARGES, $1000/YEAR	153,199	153,199	170,838	170,838	214,996	214,996	161,716	161,716
TOTAL COST OF ELECTRICITY**								
$1000/year	240,933	321,385	261,722	345,121	315,508	407,437	247,629	326,626
mills/kWh	45.42	60.59	43.87	57.84	50.79	65.59	38.87	51.28

*At 100 percent plant design power output
**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

Section 4

PLANT DESCRIPTION - CASE MASW

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 MASW-1-1. This plant consumes 13,900 ST/day of western coal (New Mexico coal, see Table 2-1).

The Case MASW plant differs from the Lurgi-based plant, presented as Case MACW in EPRI Report AF-642, primarily in that the STEAG pressurized boiler combined cycle is used for generation of electric power. The coal is the same as in Case MACW and, as with Case MACW, the main fuel gas 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 intent is to limit the impact of an upset condition to the train in which the upset occurs.

In addition to the main processing trains, the complete plant includes offsite, utility and environmental facilities. Hydrogen sulfide is removed from the crude fuel gas and processed through sulfur recovery facilities to produce elemental sulfur. Other facilities in the plant include raw water treating, cooling water, process condensate treating and effluent water treating.

The fuel gas produced in the fuel processing units then flows to two identical parallel pressurized boiler combined-cycle systems for electric power generation.

Table 4-1 summarizes major equipment sections in the plant and shows the numbers of operating and spare trains. The train count listed is the same as in Case MACW except as noted.

The following sections of this report contain descriptions of the major process units in the Case MASW plant. Emphasis is placed on describing the differences in process design between the Case MASW units and the units represented in Case MACW in EPRI Report AF-642.

Table 4-1

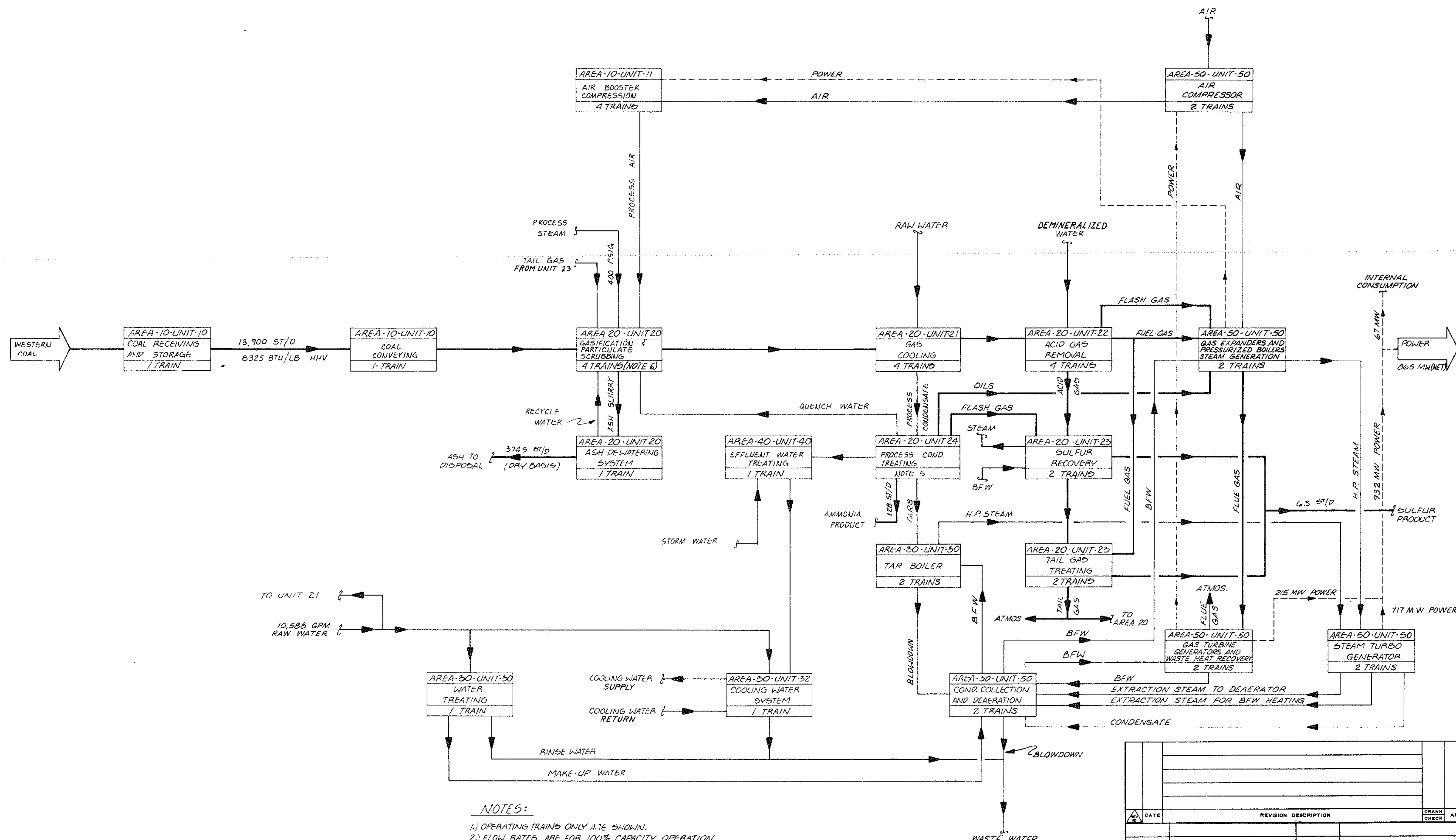
TRAINS OF EQUIPMENT IN MAIN PROCESSING UNITS - CASE MASW

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	2	1
23	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		
	Tar Boiler	2**	1
	Condensate Collection and Deaeration	2**	0
	Water Treating	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0
50	STEAG Power Generation System	2#	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

**Train count increased from the one train in Case MACW to the two listed to match the two STEAG Power generation trains

#These two STEAG power generation trains replace the six trains of gas-turbine-heat-recovery-steam-generators of Case MACW



NOTES:

1.) OPERATING TRAINS ONLY A.;E 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, ONE PHENOL EXTRATION UNIT, AND ONE AMMONIA RECOVERY UNIT.
6.) EACH TRAIN CONSISTS OF FIVE OPERATING GASIFIERS.

REV NO	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
RELEASE DATE
SUPERVISING ENGR: CRAWFORD
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

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

SCALE: NONE
DRAWING NUMBER: MASW-1-1
REVISION: 01 11-8-79

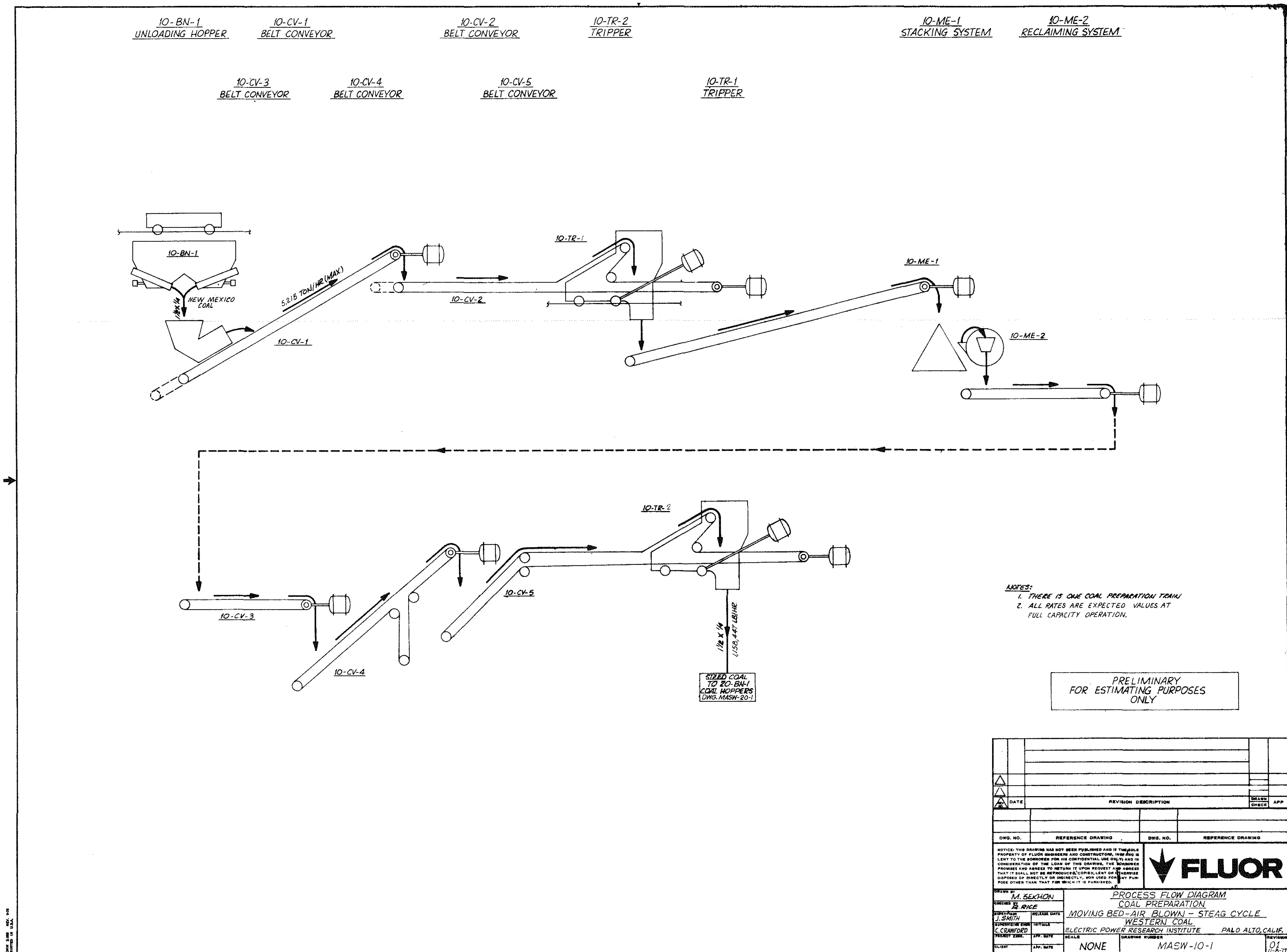
REDUCED PRINT SCALES

COAL PREPARATION

Process Flow Diagram MASW-10-1 depicts the process arrangement of equipment in this section. Coal receiving, storage, and conveying is accomplished in a single train to minimize space and operating labor requirements. This section is identical to the coal preparation section for Case MACW. Refer to EPRI Report AF-642, Case MACW, for the detailed process description of this section.

Equipment Notes

All equipment in this section 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
10-BN-1
NEW MEXICO COAL
1½x¼
5,215 TON/HR (MAX.)
10-CV-1
10-CV-2
10-TR-1
10-ME-1
10-ME-2
10-TR-2
10-CV-3
10-CV-4
10-CV-5
1½ x ¼
(158,447 LB/HR)
SIZED COAL TO 20-BN-1 COAL HOPPERS DWG. MASW-20-1
NOTES:
1. THERE IS ONE COAL PREPARATION TRAIN
2. ALL RATES ARE EXPECTED VALUES AT FULL CAPACITY OPERATION.
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
DATE
REVISION DESCRIPTION
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CHECK
APP
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PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE
PROCESS FLOW DIAGRAM
COAL PREPARATION
MOVING BED-AIR BLOWN - STEAG CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER MASW-10-1
REVISION 01 11-8-79

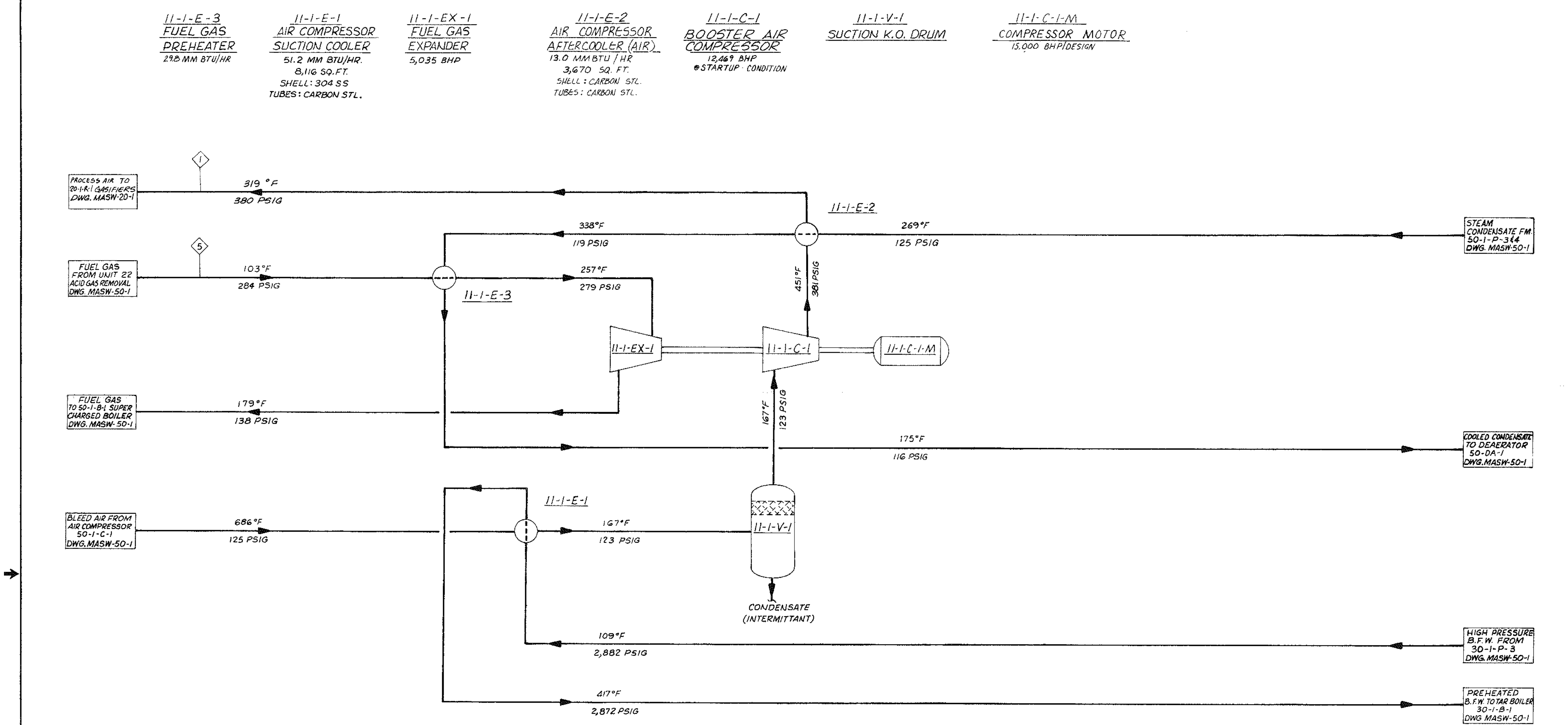
OXIDANT FEED SYSTEM

Process Flow Diagram MASW-11-1 depicts one of the four parallel trains for Case MASW. The STEAG cycle design allows the use of a simpler process for oxidant feed system heat integration than was employed in Case MACW. The steam turbine drivers for the Booster Air Compressor 11-1-C-1 have been replaced by the Fuel Gas Expander 50-1-EX-1 plus a helper motor. Also, the Therminol loop used in Case MACW for heat exchange between the combustion air and fuel gas has been eliminated.

Air for the gasifier is obtained as a bleed stream from the discharge of the combustion air compressor in the STEAG cycle. Hot bleed air at 125 psig, 686°F, exchanges heat with boiler feedwater in 11-1-E-1 and is then compressed to 381 psig in Booster Air Compressor 11-1-C-1. Hot air at 381 psig and 451°F from 11-1-C-1 is cooled to 319°F in 11-1-E-2 by heat exchange with steam condensate before flowing to the gasifier.

Equipment Notes

The design proposed for the Booster Air Compressor 11-1-C-1 in Case MASW produces a higher air discharge temperature (451°F) than is usual for centrifugal air compressors and would require an extension of the present state of the art. This compressor design should be available commercially, but will have prototype aspects.



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ONLY

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	

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NOTES:
1. THERE ARE FOUR PARALLEL OPERATING TRAINS
2. ALL RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

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

SCALE NONE
DRAWING NUMBER MASW-11-1
REVISION 01 11-8-79

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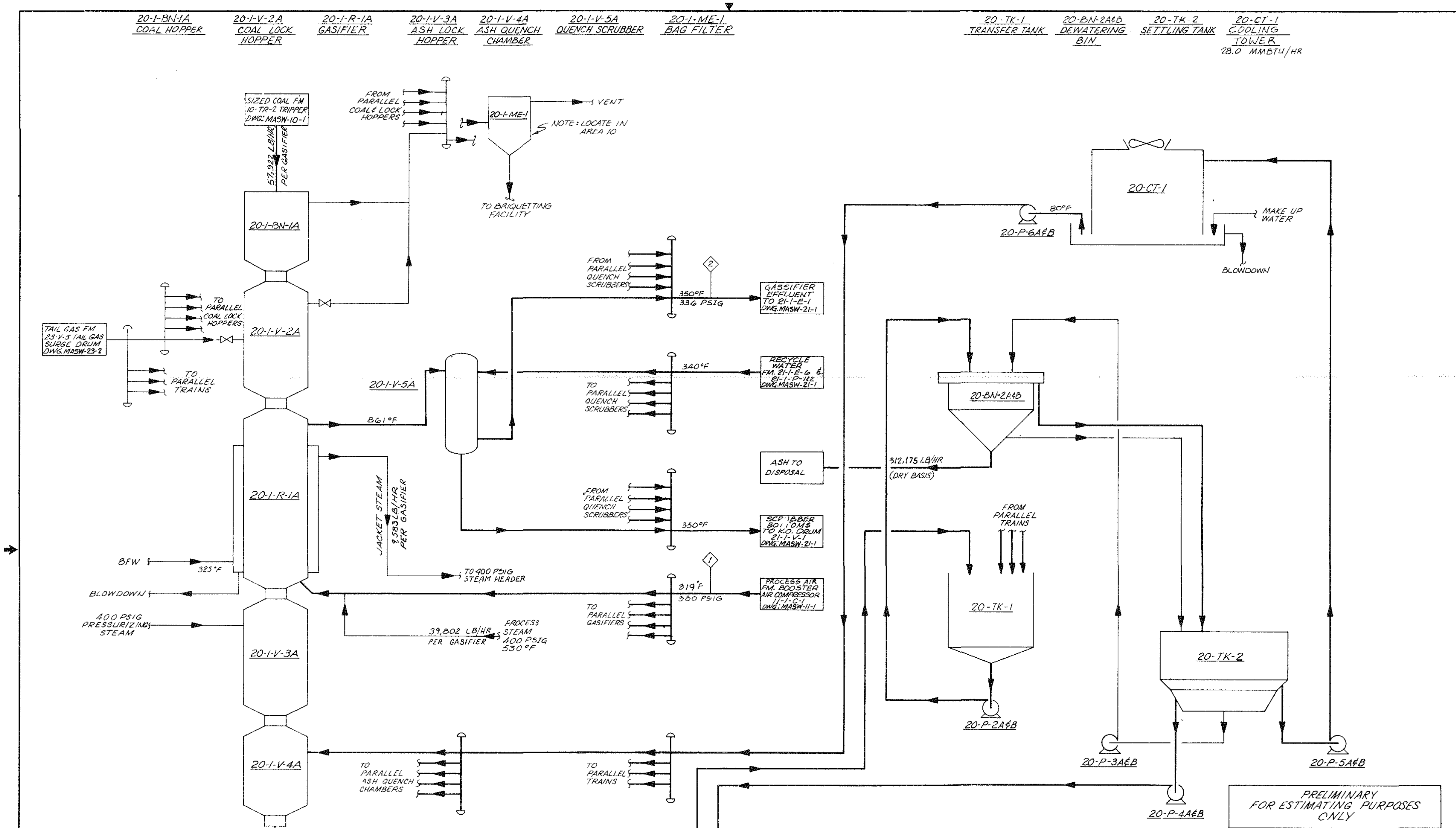
GASIFICATION AND ASH HANDLING

Process Flow Diagram MASW-20-1 shows the gasification system. There are four parallel trains, each train having five parallel gasifiers. Two additional spare gasifiers are provided. The gasification and ash handling section is similar to that provided for Case MACW. Refer to EPRI Report AF-642, Case MACW, for a detailed process description of the gasification system.

Equipment Notes

The Lurgi moving-bed air-blown 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.



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 RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.

MATERIAL BALANCE

STREAM NUMBER	1 PROCESS AIR		2 GASIFIER EFFLUENT	
COMPONENT	MPH	MOL.%	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	

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PROCESS FLOW DIAGRAM
GASIFICATION & ASH HANDLING
MOVING BED-AIR BLOWN- STEAG CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIFORNIA

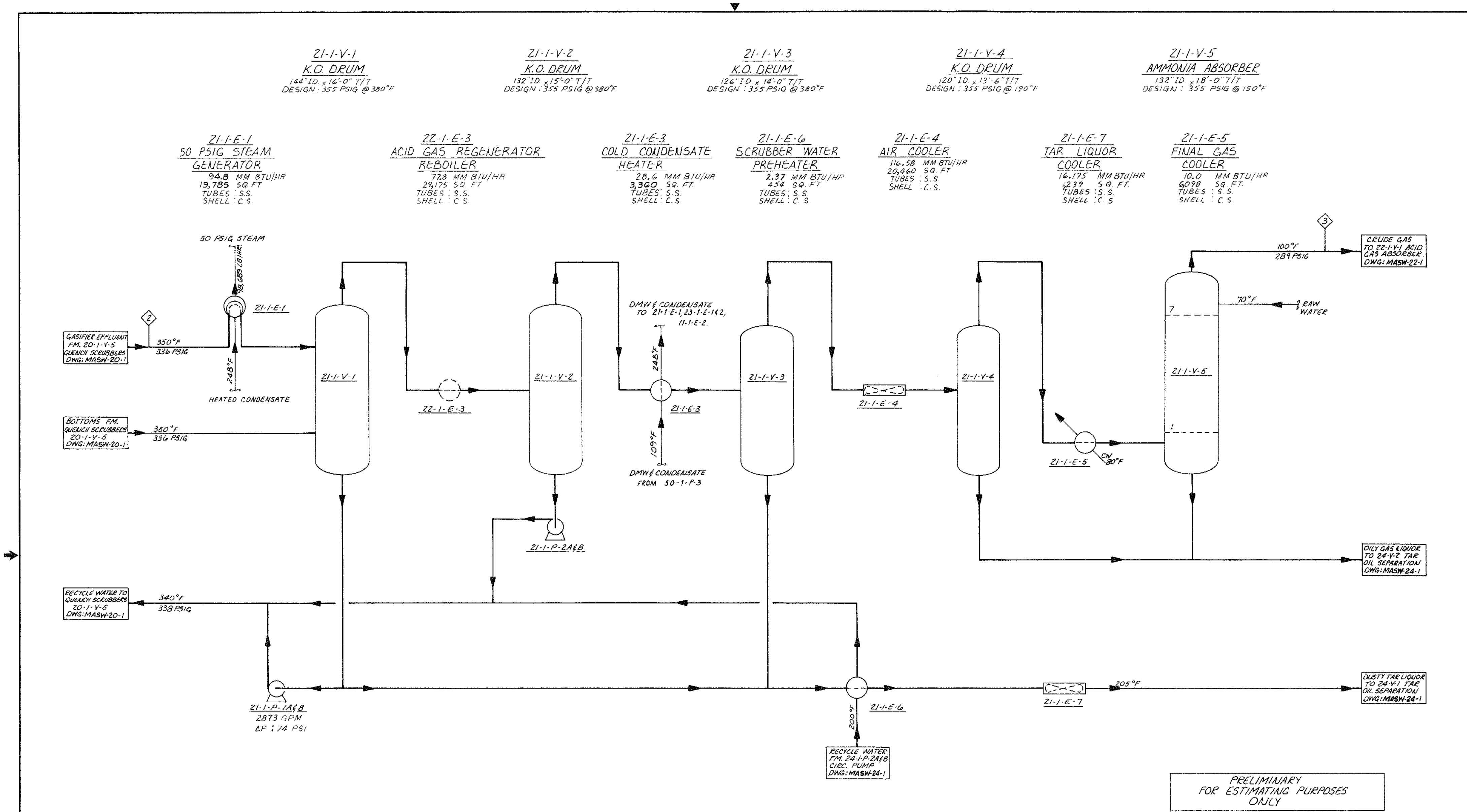
SCALE NONE
DRAWING NUMBER MAS W-20-1
REVISION 02 11-8-79

GAS COOLING

Process Flow Diagram MASW-21-1 depicts one of the four parallel gas cooling trains. The process details for each gas cooling train are similar to those described for Case MACW in EPRI Report AF-642.

Equipment Notes

All equipment is commercially available.



MATERIAL BALANCE				
STREAM NUMBER	② GASIFIER EFFLUENT		③ CRUDE GAS	
COMPONENT	MPH	MOL.%	MPH	MOL.%
CH_4	1,270.0	3.68	1270.0	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-306 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 RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.

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FLUOR

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SUPERVISOR: J. SMITH
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PROCESS FLOW DIAGRAM
GAS COOLING
MOVING BED-AIR BLOWN - STEAG CYCLE
WESTERN COAL
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.

SCALE: NONE
DRAWING NUMBER: MASW-21-1
REVISION: 02 11-8-79

REDUCED PRINT SCALES

ACID GAS REMOVAL

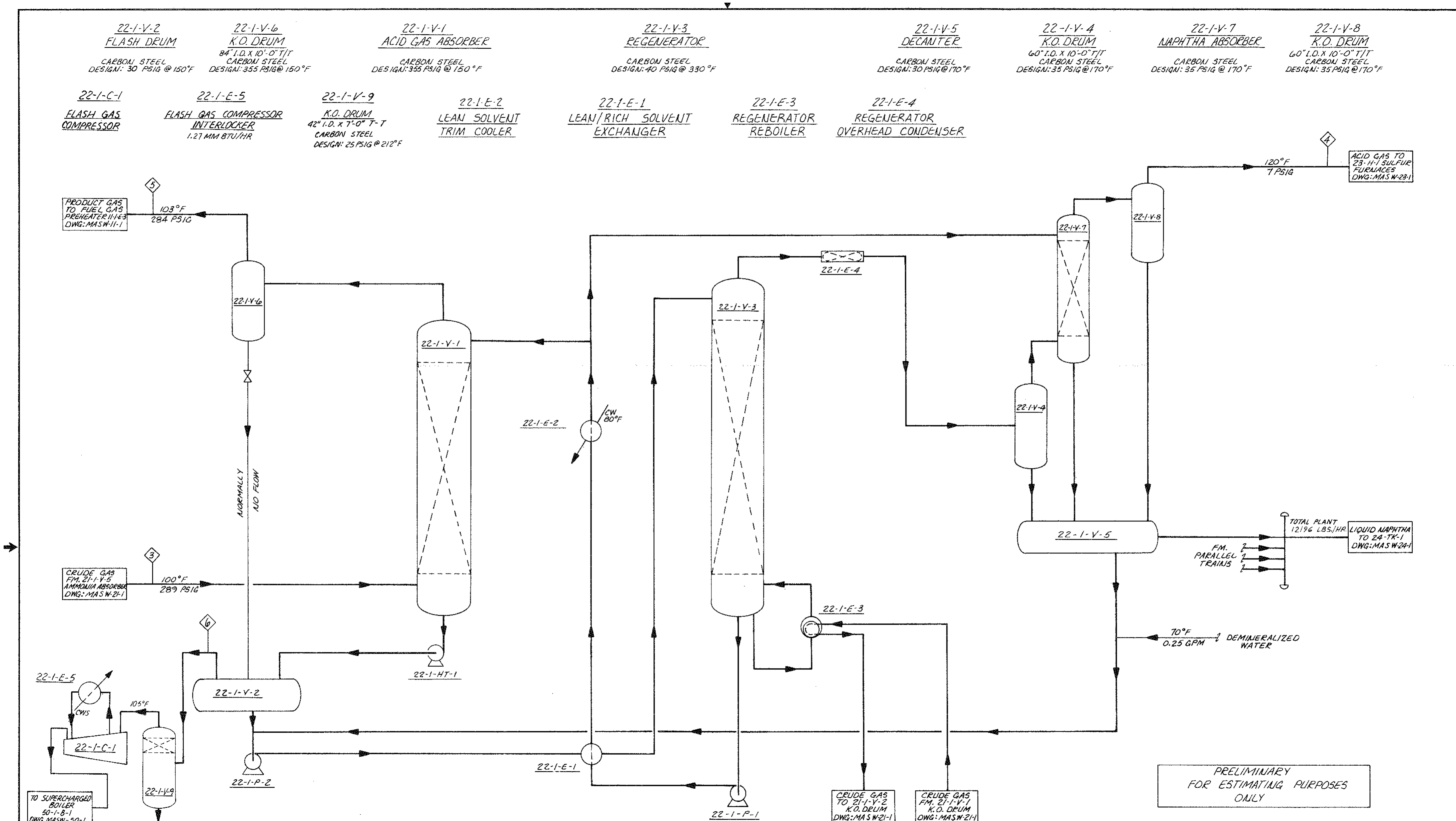
Process Flow Diagram MASW-22-1 depicts one of the four parallel acid gas removal trains. Each of these trains is similar to its counterpart designed for Case MACW and described in EPRI Report AF-642. See Acid Gas Removal in Section 5 for general comments on Selexol.

Since 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 MASW-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.

Equipment Notes

The equipment in this unit is generally 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 type of equipment.



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 RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.

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PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
MOVING BED-AIR BLOWN- STEAG CYCLE
WESTERN COAL

EPRI PALO ALTO, CALIF.

SCALE NONE | DRAWING NUMBER MASW-22-1 | REVISION 02 11-8-79

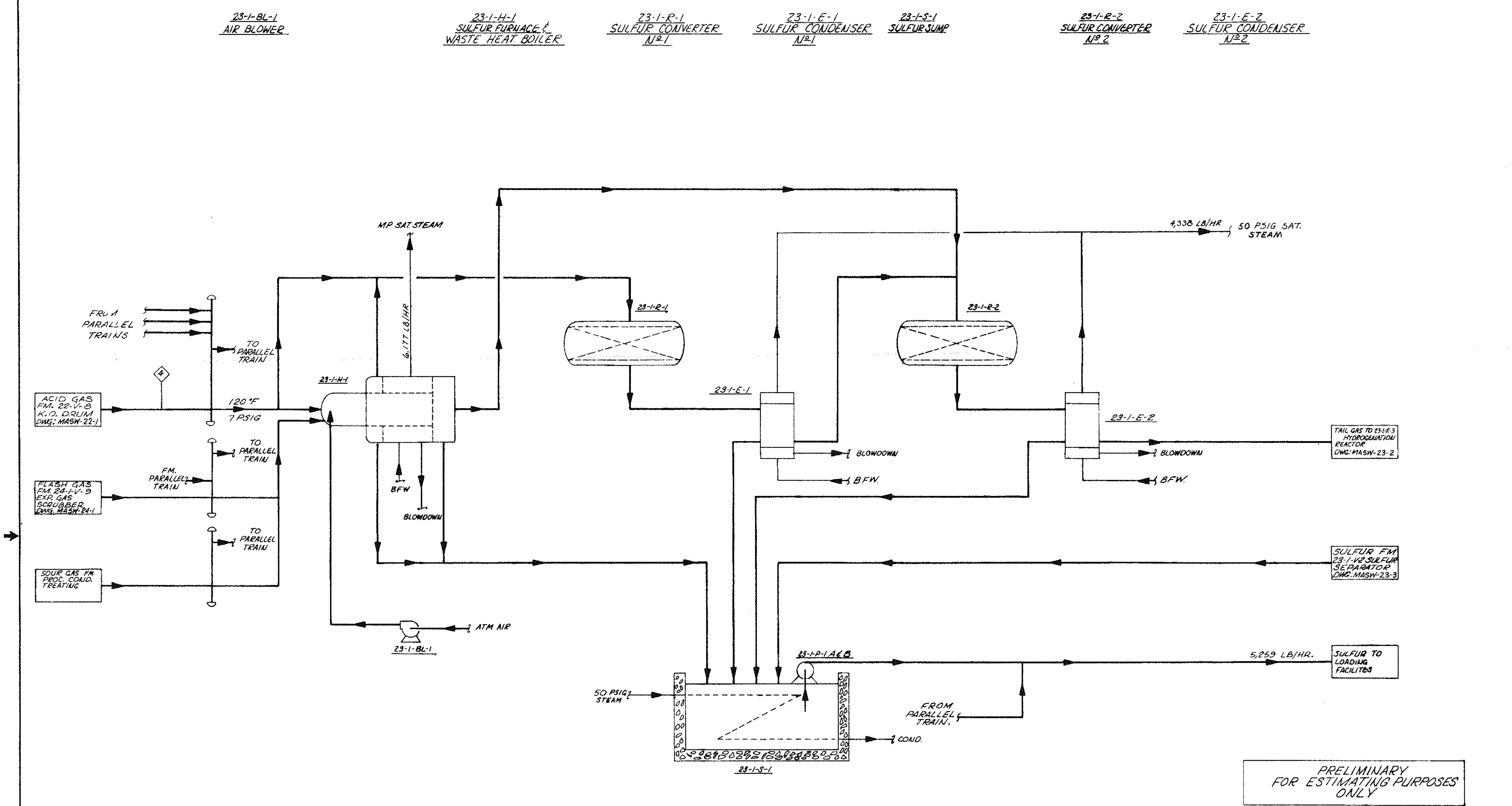
SULFUR RECOVERY AND TAIL GAS TREATING

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

There are two 50 percent 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 emission requirements fulfilled by these units. Process details for these units are similar to those described for Case MACW in EPRI Report AF-642.

Equipment Notes

The Claus sulfur process depicted in Process Flow Diagram MASW-23-1 is an established commercial process and consequently the equipment requirements are well known. Burning ammonia in the sulfur furnace 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 successfully in many commercial size plants.



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MATERIAL BALANCE

STREAM NUMBER / COMPONENT	◇4 ACID GAS 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 RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

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

SCALE NONE
DRAWING NUMBER MASW-23-1
01 11-8-74

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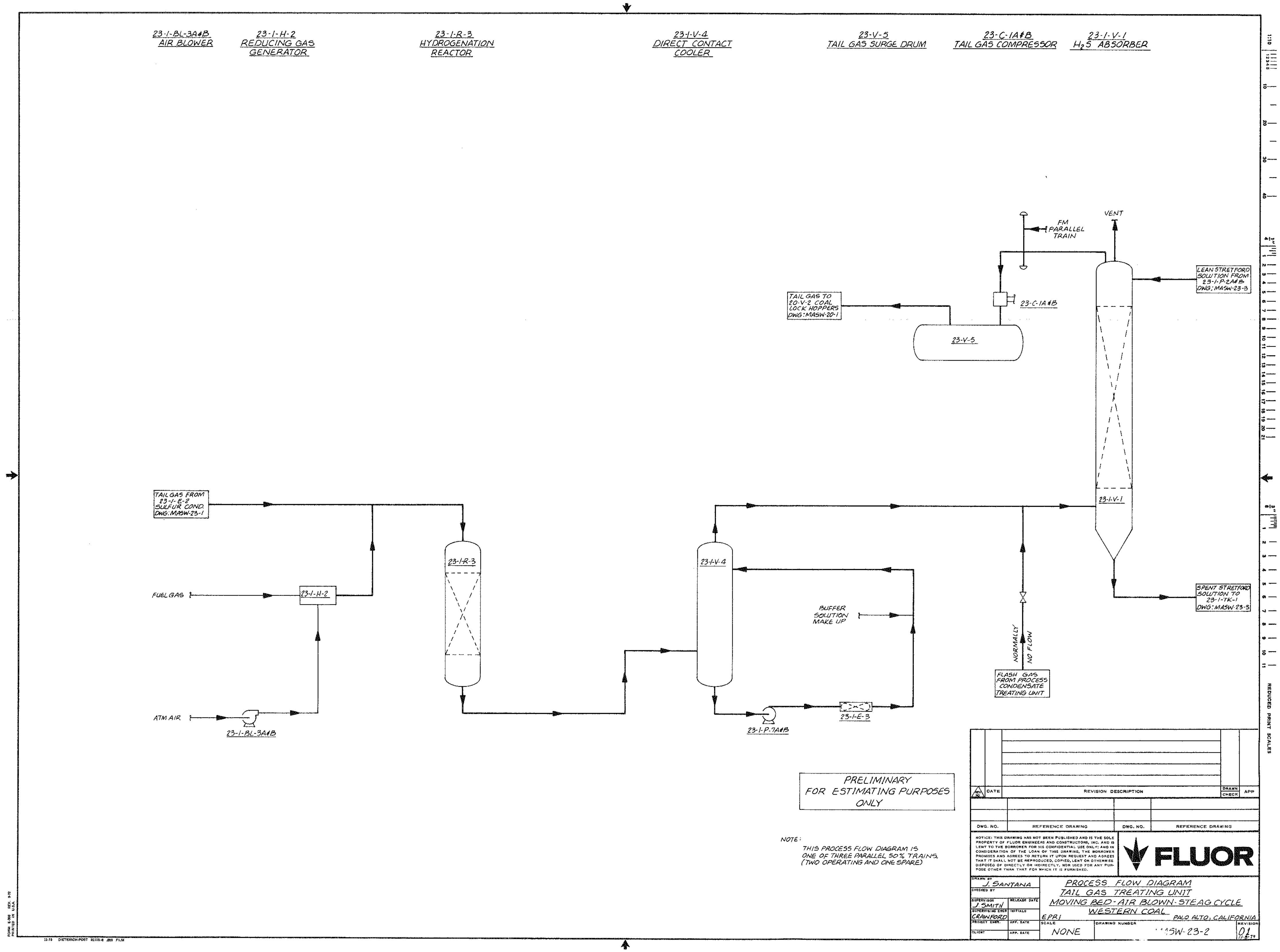
23-1-BL-3A&B AIR BLOWER
23-1-H-2 REDUCING GAS GENERATOR
23-1-R-3 HYDROGENATION REACTOR
23-1-V-4 DIRECT CONTACT COOLER
23-V-5 TAIL GAS SURGE DRUM
23-C-1A&B TAIL GAS COMPRESSOR
23-1-V-1 H_2S ABSORBER
TAIL GAS FROM 23-1-E-2 SULFUR COND. DWG: MASW-23-1
FUEL GAS
ATM AIR
23-1-H-2
23-1-BL-3A&B
23-1-R-3
23-1-V-4
BUFFER SOLUTION MAKE UP
23-1-P-7A&B
23-1-E-3
TAIL GAS TO 20-V-2 COAL LOCK HOPPERS DWG: MASW-20-1
23-V-5
23-C-1A&B
FM PARALLEL TRAIN
VENT
LEAN STRETFORD SOLUTION FROM 23-1-P-2A&B DWG: MASW-23-3
23-1-V-1
SPENT STRETFORD SOLUTION TO 23-1-TK-1 DWG: MASW-23-3
NORMALLY NO FLOW
FLASH GAS FROM PROCESS CONDENSATE TREATING UNIT
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
NOTE:
THIS PROCESS FLOW DIAGRAM IS ONE OF THREE PARALLEL 50% TRAINS. (TWO OPERATING AND ONE SPARE)
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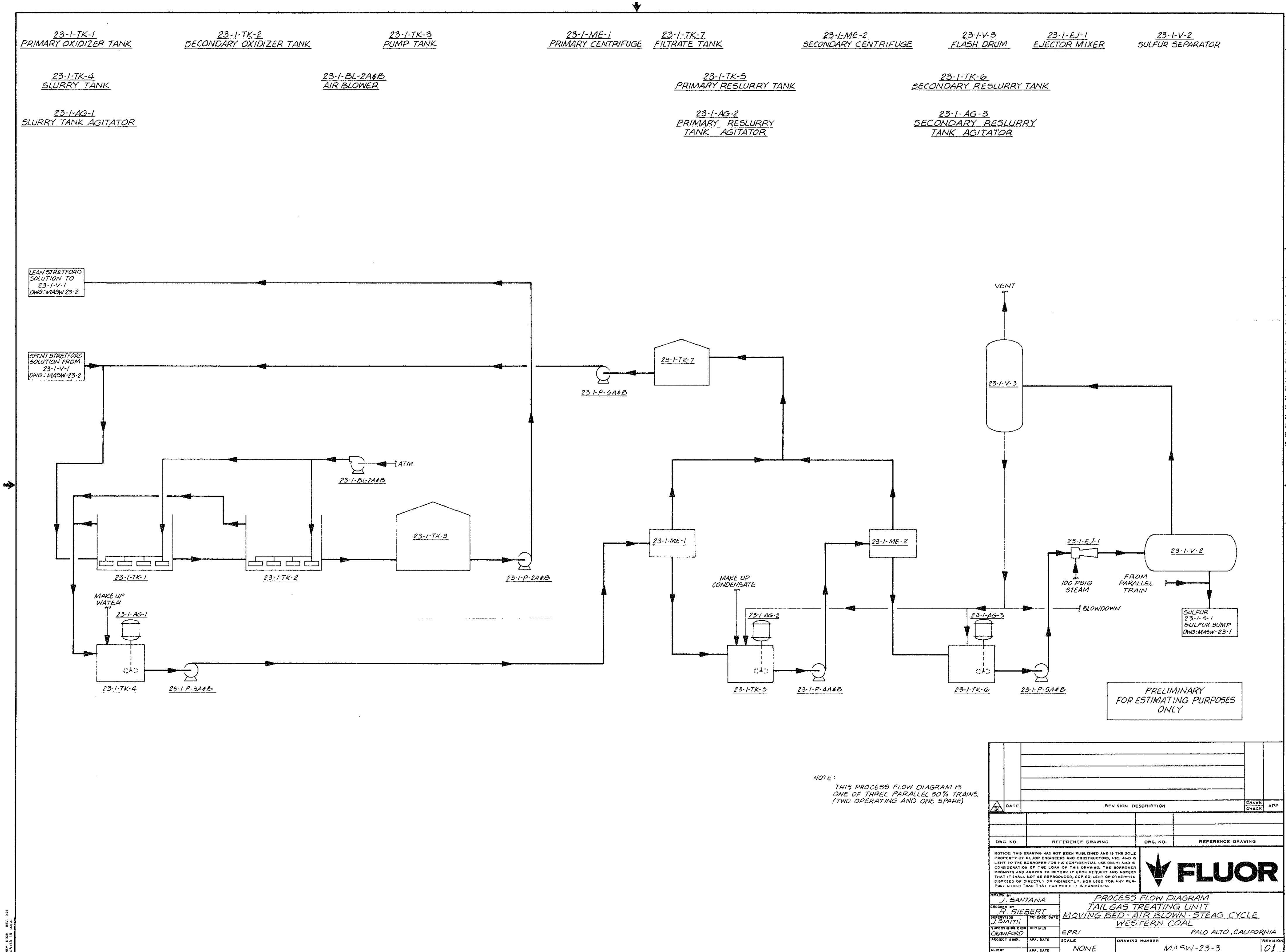
PROCESS FLOW DIAGRAM
TAIL GAS TREATING UNIT
MOVING BED - AIR BLOWN - STEAG CYCLE
WESTERN COAL
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER
SW-23-2
REVISION 01 11-8-79



23-1-TK-1
PRIMARY OXIDIZER TANK
23-1-TK-2
SECONDARY OXIDIZER TANK
23-1-TK-3
PUMP TANK
23-1-ME-1
PRIMARY CENTRIFUGE
23-1-TK-7
FILTRATE TANK
23-1-ME-2
SECONDARY CENTRIFUGE
23-1-V-3
FLASH DRUM
23-1-EJ-1
EJECTOR MIXER
23-1-V-2
SULFUR SEPARATOR
23-1-TK-4
SLURRY TANK
23-1-BL-2A&B
AIR BLOWER
23-1-TK-5
PRIMARY RESLURRY TANK
23-1-TK-6
SECONDARY RESLURRY TANK
23-1-AG-1
SLURRY TANK AGITATOR
23-1-AG-2
PRIMARY RESLURRY TANK AGITATOR
23-1-AG-3
SECONDARY RESLURRY TANK AGITATOR
LEAN STRETFORD SOLUTION TO 23-1-V-1 DWG: MASW-23-2
SPENT STRETFORD SOLUTION FROM 23-1-V-1 DWG: MASW-23-2
VENT
23-1-TK-7
23-1-P-6A&B
23-1-V-3
ATM.
23-1-BL-2A&B
23-1-TK-3
23-1-ME-1
23-1-ME-2
23-1-EJ-1
23-1-V-2
23-1-TK-1
23-1-TK-2
23-1-P-2A&B
100 PSIG STEAM
FROM PARALLEL TRAIN
MAKE UP CONDENSATE
BLOWDOWN
MAKE UP WATER
23-1-AG-1
23-1-AG-2
23-1-AG-3
SULFUR 23-1-S-1 SULFUR SUMP DWG: MASW-23-1
23-1-TK-4
23-1-P-3A&B
23-1-TK-5
23-1-P-4A&B
23-1-TK-6
23-1-P-5A&B
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
NOTE:
THIS PROCESS FLOW DIAGRAM IS ONE OF THREE PARALLEL 50% TRAINS. (TWO OPERATING AND ONE SPARE)
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APP. DATE
PROCESS FLOW DIAGRAM
TAIL GAS TREATING UNIT
MOVING BED - AIR BLOWN - STEAG CYCLE
WESTERN COAL
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER MASW-23-3
REVISION 01 11-8-79
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PROCESS CONDENSATE TREATING

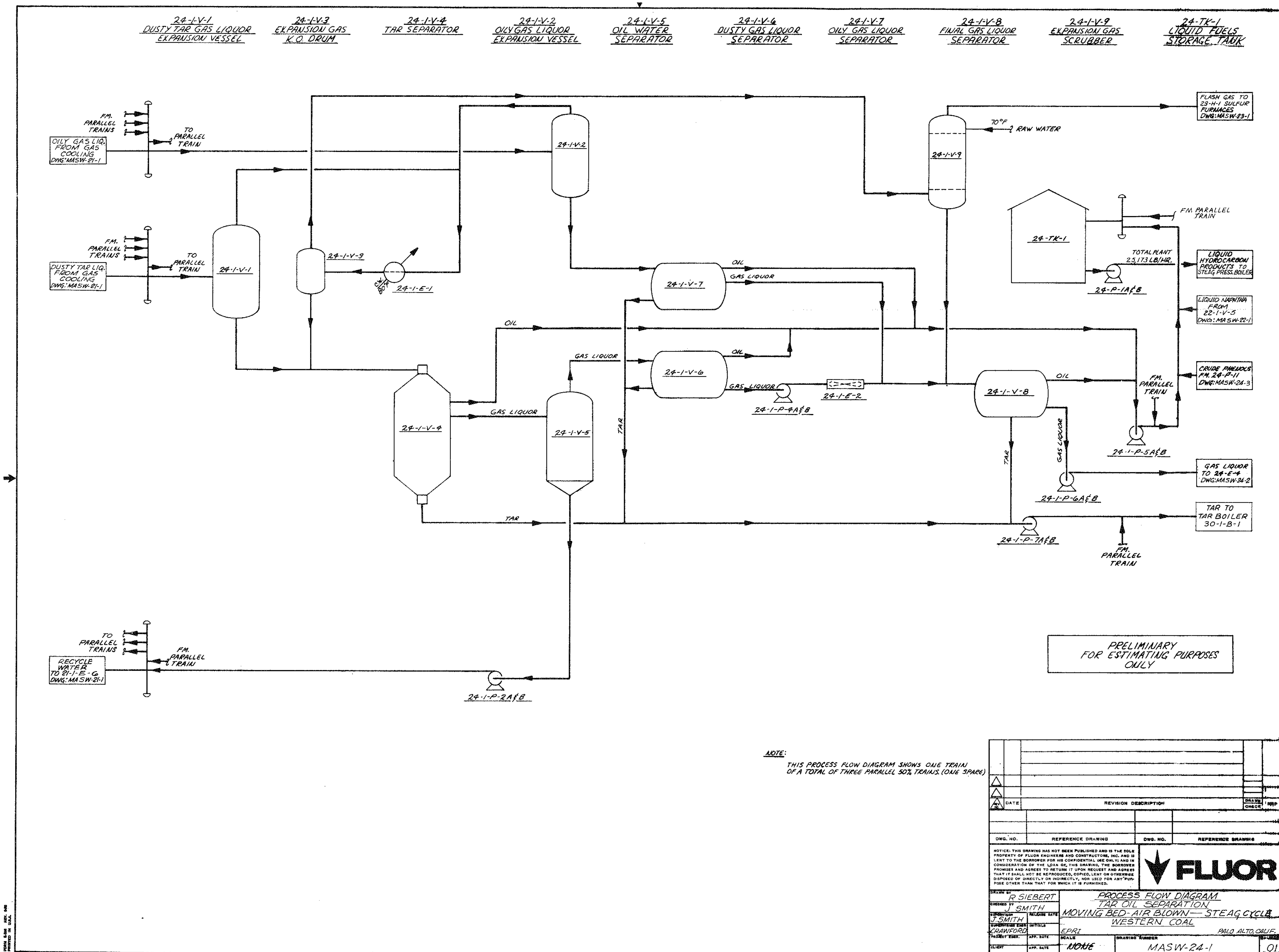
Process Flow Diagrams MASW-24-1, MASW-24-2, MASW-24-3 and MASW-24-4 schematically represent the flow through the tar oil separation unit and phenol extraction plant, respectively, as described in applications to the Federal Power Commission by Transwestern Pipeline Company (1) and El Paso Natural Gas Company (2). As shown on Process Flow Diagram MASW-24-1, the process differs slightly from Case MACW in that after the tar and oil have been separated, only the tar is combusted as fuel in the Tar Boiler 30-1-B-1. The separated oil is used as a fuel in the pressurized boiler in the STEAG combined cycle system. In contrast, in the system for Case MACW, both streams are combusted as fuel in the Tar Boiler 30-1-B-1. This design change for Case MASW substantially reduces the cost of the Tar Boiler 30-1-B-1 with its associated stack gas scrubbing systems without significantly increasing the cost of the STEAG pressurized boiler. This ability to utilize the separated oil by-product as fuel in the pressurized boiler is one of the advantages of the STEAG cycle.

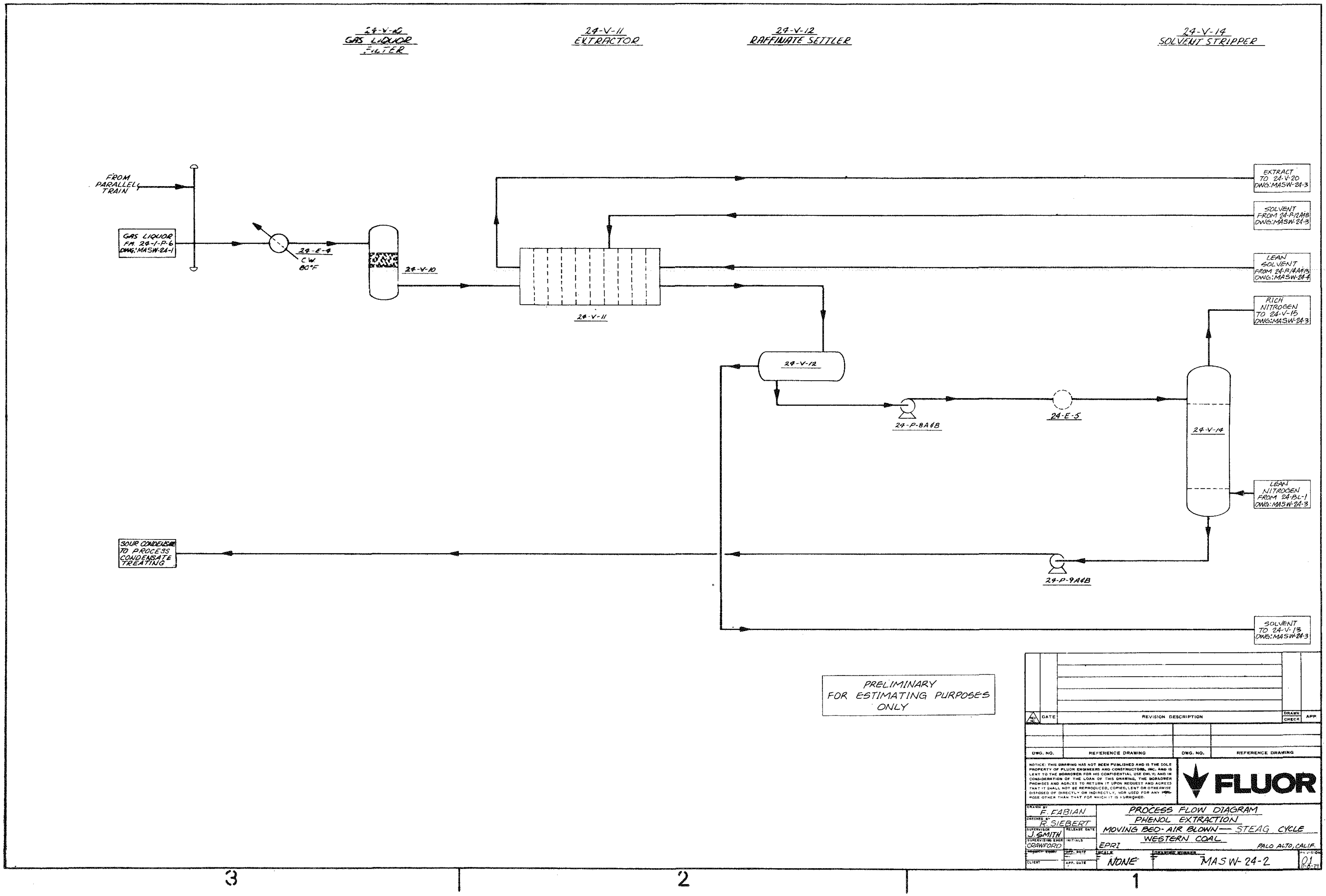
Other features of the tar oil separation unit and phenol extraction plant remain the same as described for Case MACW in EPRI Report AF-642.

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.


24-1-V-1
DUSTY TAR GAS LIQUOR EXPANSION VESSEL
24-1-V-3
EXPANSION GAS K.O. DRUM
24-1-V-4
TAR SEPARATOR
24-1-V-2
OILY GAS LIQUOR EXPANSION VESSEL
24-1-V-5
OIL WATER SEPARATOR
24-1-V-6
DUSTY GAS LIQUOR SEPARATOR
24-1-V-7
OILY GAS LIQUOR SEPARATOR
24-1-V-8
FINAL GAS LIQUOR SEPARATOR
24-1-V-9
EXPANSION GAS SCRUBBER
24-TK-1
LIQUID FUELS STORAGE TANK
FM. PARALLEL TRAINS
TO PARALLEL TRAIN
OILY GAS LIQ. FROM GAS COOLING DWG: MASW-21-1
DUSTY TAR LIQ. FROM GAS COOLING DWG: MASW-21-1
TO PARALLEL TRAINS
FM. PARALLEL TRAIN
RECYCLE WATER TO 21-1-E-6 DWG: MASW-21-1
24-1-E-1
24-1-E-2
70°F RAW WATER
FLASH GAS TO 23-H-1 SULFUR FURNACES DWG: MASW-23-1
TOTAL PLANT 23,173 LB/HR.
LIQUID HYDROCARBON PRODUCTS TO STEAG PRESS. BOILER
LIQUID NAPHTHA FROM 22-1-V-5 DWG: MASW-22-1
CRUDE PHENOLS FM. 24-P-11 DWG: MASW-24-3
GAS LIQUOR TO 24-E-4 DWG: MASW-24-2
TAR TO TAR BOILER 30-1-B-1
OIL
GAS LIQUOR
TAR
24-P-1A&B
24-1-P-2A&B
24-1-P-4A&B
24-1-P-5A&B
24-1-P-6A&B
24-1-P-7A&B
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
NOTE:
THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE PARALLEL 50% TRAINS. (ONE SPARE)
DATE
REVISION DESCRIPTION
DWG. NO.
REFERENCE DRAWING
FLUOR
R SIEBERT
J SMITH
J. SMITH
CRAWFORD
PROCESS FLOW DIAGRAM
TAR OIL SEPARATION
MOVING BED-AIR BLOWN— STEAG CYCLE
WESTERN COAL
EPRI
PALO ALTO, CALIF.
NONE
MASW-24-1
01



24-V-10
GAS LIQUOR FILTER
24-V-11
EXTRACTOR
24-V-12
RAFFINATE SETTLER
24-V-14
SOLVENT STRIPPER
FROM PARALLEL TRAIN
GAS LIQUOR FM 24-I-P-6 DWG: MASW-24-1
24-E-4
CW 80°F
24-V-10
24-V-11
EXTRACT TO 24-V-20 DWG: MASW-24-3
SOLVENT FROM 24-P-12A&B DWG: MASW-24-3
LEAN SOLVENT FROM 24-P-4A&B DWG: MASW-24-4
RICH NITROGEN TO 24-V-15 DWG: MASW-24-3
24-V-12
24-P-8A&B
24-E-5
24-V-14
LEAN NITROGEN FROM 24-BL-1 DWG: MASW-24-3
SOUR CONDENSATE TO PROCESS CONDENSATE TREATING
24-P-9A&B
SOLVENT TO 24-V-13 DWG: MASW-24-3
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
REV NO
DATE
REVISION DESCRIPTION
DRAWN
CHECK
APP.
DWG. NO.
REFERENCE DRAWING
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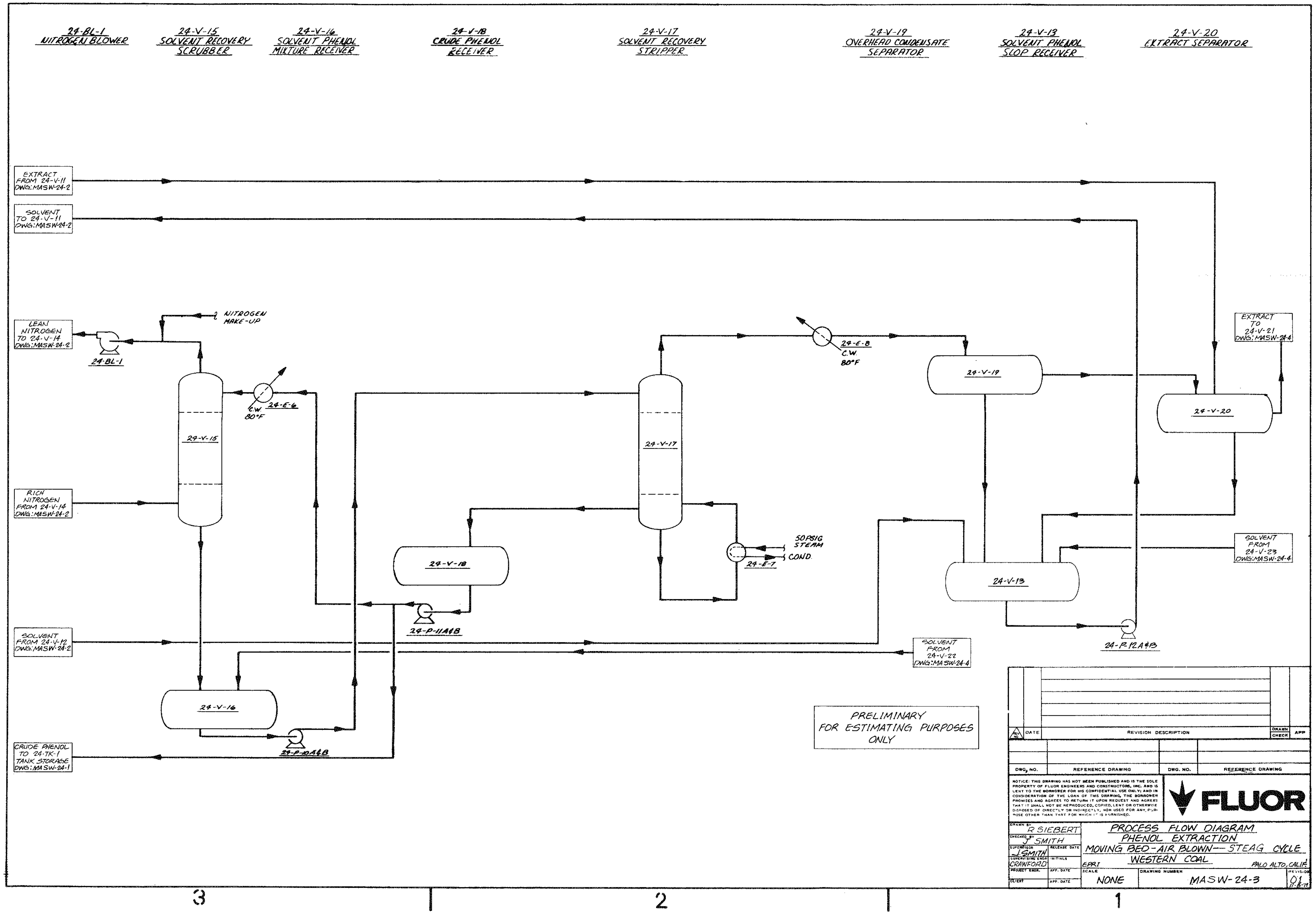
DRAWN BY F. FABIAN
CHECKED BY R. SIEBERT
SUPERVISOR J. SMITH
RELEASE DATE
SUPERVISING ENGR CRAWFORD
INITIALS
PROCESS FLOW DIAGRAM
PHENOL EXTRACTION
MOVING BED-AIR BLOWN — STEAG CYCLE
WESTERN COAL
EPRI
PALO ALTO, CALIF.
CLIENT
APP. DATE
SCALE NONE
MASW-24-2
REVISION 01 11-8-79
3
2
1


24-BL-1 NITROGEN BLOWER
24-V-15 SOLVENT RECOVERY SCRUBBER
24-V-16 SOLVENT PHENOL MIXTURE RECEIVER
24-V-18 CRUDE PHENOL RECEIVER
24-V-17 SOLVENT RECOVERY STRIPPER
24-V-19 OVERHEAD CONDENSATE SEPARATOR
24-V-13 SOLVENT PHENOL SLOP RECEIVER
24-V-20 EXTRACT SEPARATOR
EXTRACT FROM 24-V-11 DWG: MASW-24-2
SOLVENT TO 24-V-11 DWG: MASW-24-2
LEAN NITROGEN TO 24-V-14 DWG: MASW-24-2
NITROGEN MAKE-UP
RICH NITROGEN FROM 24-V-14 DWG: MASW-24-2
SOLVENT FROM 24-V-12 DWG: MASW-24-2
CRUDE PHENOL TO 24-TK-1 TANK STORAGE DWG: MASW-24-1
C.W. 80°F
24-E-6
24-E-7
24-E-8
50 PSIG STEAM
COND.
24-P-11A&B
24-P-10A&B
24-P-12A&B
SOLVENT FROM 24-V-22 DWG: MASW-24-4
SOLVENT FROM 24-V-23 DWG: MASW-24-4
EXTRACT TO 24-V-21 DWG: MASW-24-4
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
DATE
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CHECK
APP
DWG. NO.
REFERENCE DRAWING
FLUOR
DRAWN BY R SIEBERT
CHECKED BY J SMITH
J SMITH
CRAWFORD
PROCESS FLOW DIAGRAM
PHENOL EXTRACTION
MOVING BED-AIR BLOWN—STEAG CYCLE
WESTERN COAL
EPRI
PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER MASW-24-3
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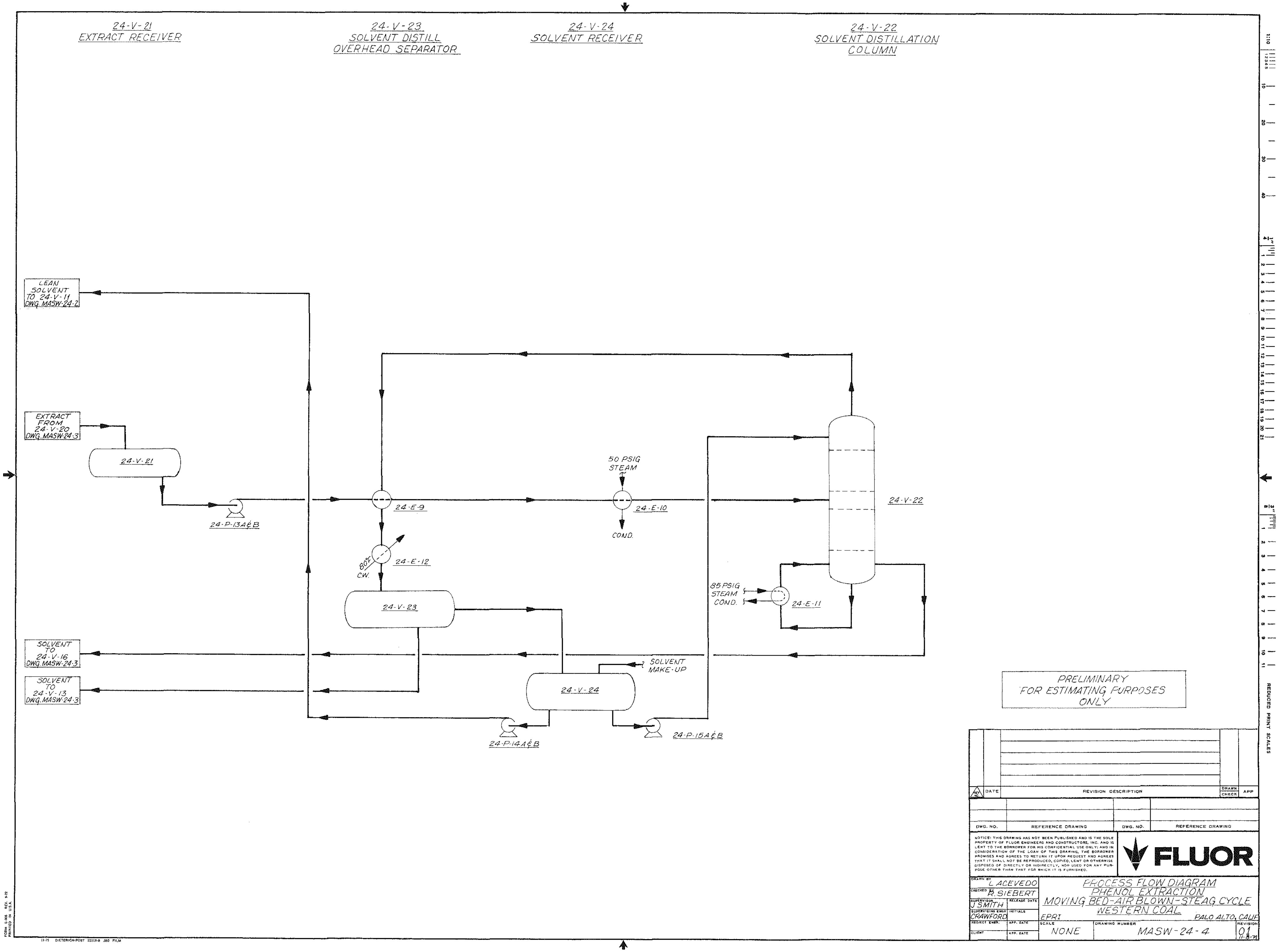

24-V-21
EXTRACT RECEIVER
24-V-23
SOLVENT DISTILL
OVERHEAD SEPARATOR
24-V-24
SOLVENT RECEIVER
24-V-22
SOLVENT DISTILLATION
COLUMN
LEAN SOLVENT TO 24-V-11 DWG.MASW-24-2
EXTRACT FROM 24-V-20 DWG.MASW-24-3
24-V-21
24-P-13A&B
24-E-9
50 PSIG STEAM
24-E-10
COND.
80°F CW.
24-E-12
24-V-23
24-V-22
85 PSIG STEAM
COND.
24-E-11
SOLVENT TO 24-V-16 DWG.MASW-24-3
SOLVENT TO 24-V-13 DWG.MASW-24-3
SOLVENT MAKE-UP
24-V-24
24-P-14A&B
24-P-15A&B
PRELIMINARY
FOR ESTIMATING PURPOSES
ONLY
DATE
REVISION DESCRIPTION
DRAWN
CHECK
APP
DWG. NO.
REFERENCE DRAWING
DWG. NO.
REFERENCE DRAWING
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CHECKED BY R. SIEBERT
SUPERVISOR J SMITH
RELEASE DATE
SUPERVISING ENGR CRAWFORD
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE
PROCESS FLOW DIAGRAM
PHENOL EXTRACTION
MOVING BED-AIR BLOWN-STEAG CYCLE
WESTERN COAL
EPRI
PALO ALTO, CALIF
SCALE NONE
DRAWING NUMBER MASW-24-4
REVISION 01 11-8-79
REDUCED PRINT SCALES

STEAM, BOILER FEEDWATER, AND CONDENSATE

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

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

Turbine Inlet	- 2669 psig (185 bar), 977°F
Intermediate Pressure (IP)	- 595 psig (42 bar) (at reheater inlet)
Low Pressure	- 86 psig (6.9 bar)
Low Pressure	- 50 psig (4.4 bar)

The major high pressure steam generation is carried out in the STEAG combined-cycle system in the pressurized boiler 50-1-B-1. There are two parallel identical trains of equipment, each train contains a pair of pressurized boilers. The STEAG system is described in detail in Appendix A. Additional generation of high-pressure superheated steam, about 14 of the total HP superheated steam, is obtained in the tar boiler, 30-1-B-1. All high-pressure superheated steam is used to drive the steam turbine, 50-1-T-1.

The high-pressure (HP) end of the turbine, 50-1-T-1, exhausts steam at 595 psig (42 bar). The majority of this steam is reheated to 977°F in the pressurized boiler 50-1-B-1. The reheated intermediate pressure (IP) steam is then used in the IP end of turbine 50-1-T-1. A small portion of the 595 psig (42 bar) turbine exhaust steam is used for boiler feedwater heating. The balance is reduced in pressure to 400 psig and flows to the gasification area. Additional 400 psig steam is generated in the sulfur furnace (23-1-H-1) and in the jackets of the gasifiers 20-1-R-1. All these 400 psig sources are used to supply process steam to the gasifiers, 20-1-R-1.

The IP end of turbine 50-1-T-1 exhausts steam at 86 psig (6.94 bar). A portion of this is used as stripping steam in the deaerator, 50-1-DA-1. Additional 86 psig turbine 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 86 psig exhaust steam is used in the condensing turbine 50-1-T-2. This turbine exhausts at 2.5 inches Hg abs.

Condensed steam from the main power turbine surface condenser, 50-1-E-5, is pumped by condensate pump 50-1-P-3 from 2.5 inches Hg abs., to 131 psig (10 bar) in order to produce flow to the deaerator. A sidestream of this condensate is pumped by booster pump 30-1-P-3 as feedwater to the tar boiler, 30-1-B-1, for generation of high pressure steam. The balance returns to the deaerator. This condensate stream is preheated to 288°F in Exchangers 50-1-E-1, 50-1-E-2, 50-1-E-3, and 50-1-E-4. These exchangers are heated with low-pressure extraction steam from turbine 50-1-T-2 combined with 50 psig process steam. Part of this stream is routed through Stack Gas Recuperator 50-1-E-10 for additional heating before returning to the deaerator.

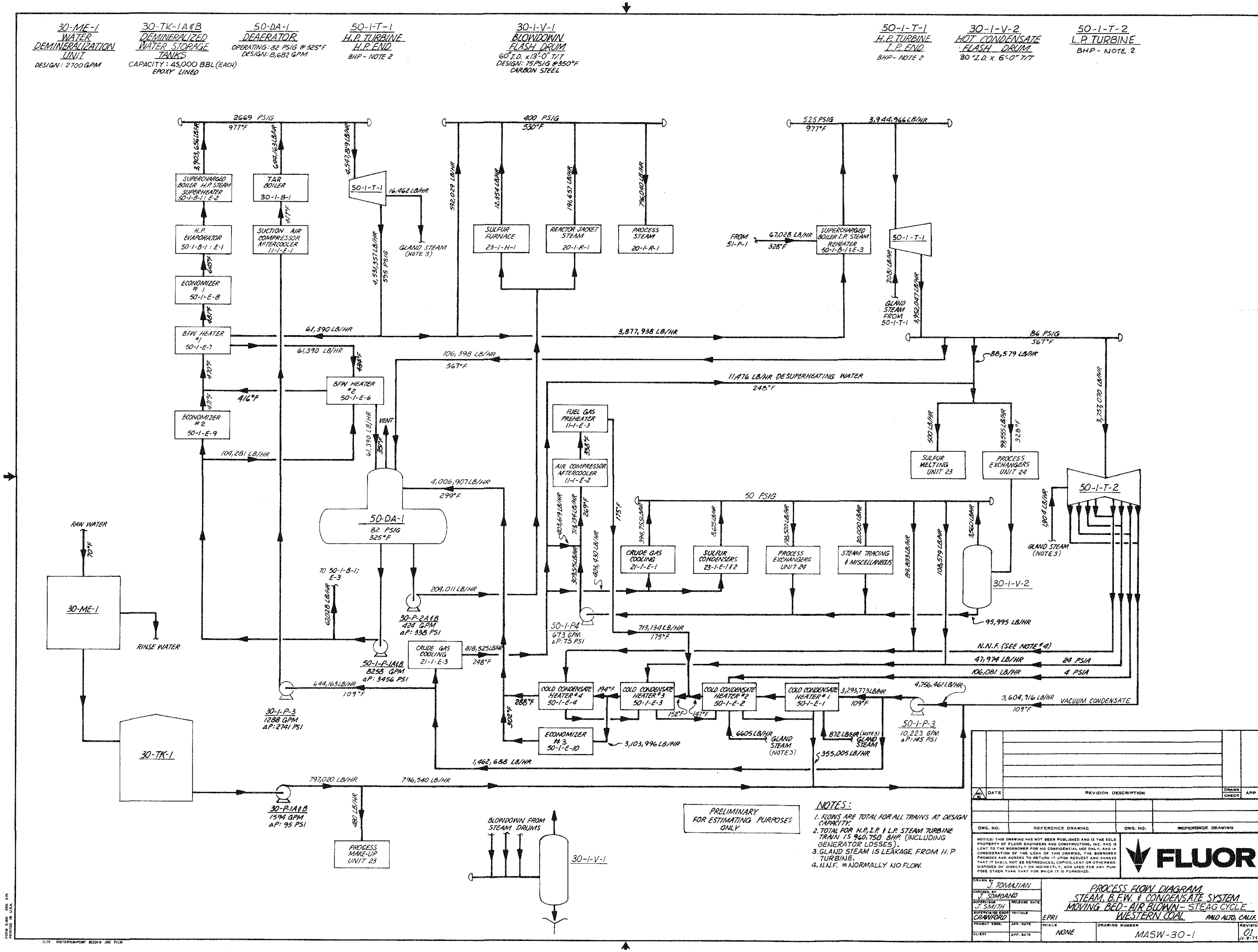
The deaerator, 50-1-DA-1, in the STEAG cycle operates at 82 psig. Stripping steam is supplied from turbine extraction. The majority of the boiler feedwater from the deaerator is pumped by high-pressure boiler feedwater pumps (50-1-P-1) through feedwater preheaters to the pressurized boiler 50-1-B-1 for generation of high-pressure steam. A smaller stream of boiler feedwater from the deaerator is pumped by medium-pressure boiler feedwater pumps (30-1-P-2) to the 400 psig steam generators: the sulfur furnace 23-1-H-1 in the Claus plant and the jackets of the Lurgi gasifiers.

The recirculating boiler feedwater is treated continuously in a Benson-type water treating system for removal of dissolved salts as well as dissolved oxygen. This water treatment minimizes the makeup and blowdown requirements.

Raw water is treated in a semiautomatic, resin bed demineralization unit 30-ME-1 to produce demineralized water suitable for use in the STEAG supercharged boiler. Demineralized water is stored in a tank, 30-TK-1. Storage equivalent to 24 hours of demineralized water production is provided. Demineralized water from the storage tank is transported to the deaerator through Pumps 30-1-P-1A&B. A small quantity of the makeup water is withdrawn from the discharge of 30-1-P-1A&B and transported to Unit 22. The balance of the demineralized water combines with the vacuum condensate from the turbine 50-1-T-2, and after preheating in 50-1-E-1, 50-1-E-2, 50-1-E-3, 50-1-E-4, and 50-1-E-10, flows to the deaerator.

Equipment Notes

The design of the tar boilers 30-1-B-1 is conventional. The emission controls for the tar boilers include electrostatic precipitators and stack gas cleanup facilities to meet the total plant specified emission limit of 1.2 lbs SO_2/MM Btu.



30-ME-1 WATER DEMINERALIZATION UNIT
DESIGN: 2700 GPM
30-TK-1A&B DEMINERALIZED WATER STORAGE TANKS
CAPACITY: 45,000 BBL (EACH) EPOXY LINED
50-DA-1 DEAERATOR
OPERATING: 82 PSIG @ 325°F
DESIGN: 8,682 GPM
50-1-T-1 H.P. TURBINE H.P. END
BHP - NOTE 2
30-1-V-1 BLOWDOWN FLASH DRUM
60" I.D. x 13'-0" T/T
DESIGN: 75 PSIG @ 350°F
CARBON STEEL
50-1-T-1 H.P. TURBINE L.P. END
BHP - NOTE 2
30-1-V-2 HOT CONDENSATE FLASH DRUM
30" I.D. x 6'-0" T/T
50-1-T-2 L.P. TURBINE
BHP - NOTE 2
2669 PSIG
977°F
400 PSIG
530°F
525 PSIG
977°F
3,944,966 LB/HR
SUPERCHARGED BOILER H.P. STEAM SUPERHEATER 50-1-B-1: E-2
TAR BOILER 30-1-B-1
H.P. EVAPORATOR 50-1-B-1: E-1
SUCTION AIR COMPRESSOR AFTERCOOLER 11-1-E-1
ECONOMIZER #1 50-1-E-8
BFW HEATER #1 50-1-E-7
BFW HEATER #2 50-1-E-6
ECONOMIZER #2 50-1-E-9
SULFUR FURNACE 23-1-H-1
REACTOR JACKET STEAM 20-1-R-1
PROCESS STEAM 20-1-R-1
GLAND STEAM (NOTE 3)
FROM 51-P-1
SUPERCHARGED BOILER I.P. STEAM REHEATER 50-1-B-1: E-3
GLAND STEAM FROM 50-1-T-1
86 PSIG
567°F
11,476 LB/HR DESUPERHEATING WATER
FUEL GAS PREHEATER 11-1-E-3
AIR COMPRESSOR AFTERCOOLER 11-1-E-2
SULFUR MELTING UNIT 23
PROCESS EXCHANGERS UNIT 24
50-1-T-2
50 PSIG
50-DA-1
82 PSIG
325°F
CRUDE GAS COOLING 21-1-E-1
SULFUR CONDENSERS 23-1-E-1 & 2
PROCESS EXCHANGERS UNIT 24
STEAM TRACING & MISCELLANEOUS
30-1-V-2
RAW WATER
30-ME-1
RINSE WATER
TO 50-1-B-1; E-3
30-P-2A&B
424 GPM
ΔP: 338 PSI
50-1-P4
673 GPM
ΔP: 75 PSI
CRUDE GAS COOLING 21-1-E-3
50-1-P-1A&B
8258 GPM
ΔP: 3456 PSI
N.N.F. (SEE NOTE #4)
24 PSIA
4 PSIA
VACUUM CONDENSATE
COLD CONDENSATE HEATER #4 50-1-E-4
COLD CONDENSATE HEATER #3 50-1-E-3
COLD CONDENSATE HEATER #2 50-1-E-2
COLD CONDENSATE HEATER #1 50-1-E-1
ECONOMIZER #3 50-1-E-10
30-1-P-3
1288 GPM
ΔP: 2741 PSI
50-1-P-3
10,223 GPM
ΔP: 145 PSI
30-TK-1
30-P-1A&B
1594 GPM
ΔP: 95 PSI
PROCESS MAKE-UP UNIT 23
BLOWDOWN FROM STEAM DRUMS
30-1-V-1
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
NOTES:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. TOTAL FOR H.P., I.P. & L.P. STEAM TURBINE TRAIN IS 960,750 BHP. (INCLUDING GENERATOR LOSSES).
3. GLAND STEAM IS LEAKAGE FROM H.P. TURBINE.
4. N.N.F. = NORMALLY NO FLOW.
FLUOR
PROCESS FLOW DIAGRAM
STEAM, B.F.W. & CONDENSATE SYSTEM
MOVING BED-AIR BLOWN-STEAG CYCLE
WESTERN COAL
EPRI
PALO ALTO, CALIF.
DRAWN BY J. TOMAJIAN
CHECKED BY J. SOMOANO
SUPERVISOR J. SMITH
CRAWFORD
SCALE NONE
DRAWING NUMBER MASW-30-1
REVISION 01

COMBINED-CYCLE SYSTEM

Process Flow Diagram MASW-50-1 depicts the combined-cycle system for Case MASW. This schematic diagram was prepared by STEAG and shows both total power block flows and various sections of the gasification plant which are described on other flow diagrams. These sections include the gasification, oxidant supply, and steam systems.

The power block is divided into two independent parallel trains of identical capacity. Each train contains two pressurized boilers, 50-1-B-1, one gas turbine, 50-1-GT-1, one steam turbine, 50-1-T-1&2, and one steam turbine generator, 50-1-G-2. Each train is designated a "Kombi-block." Refer to Appendix A for a detailed description of the pressurized boiler combined cycle system, including detailed performance information for the power block components, i.e., pressurized boilers, steam turbines, and gas turbines.

The steam system is described separately in connection with Process Flow Diagram MASW-30-1.

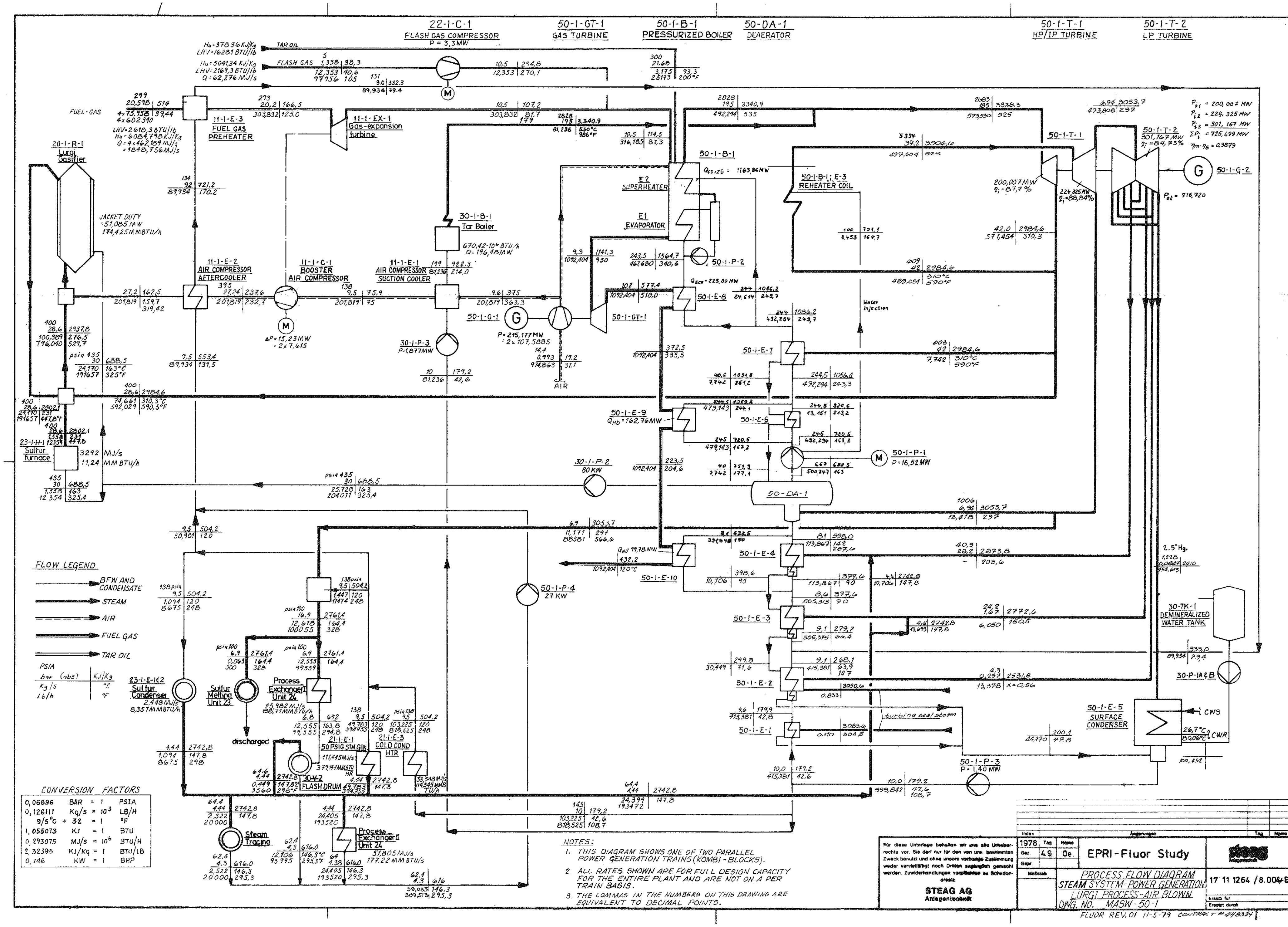
The following is a brief description of the process flow of fuel gas received from the gasification area. Fuel gas enters the power block at 138 psig and 179°F from the exhaust of gas expansion turbine 11-1-EX-1 in Unit 11. The fuel gas is then combined with a small amount of flash gas from Unit 22 and the combined gas flows to the pressurized boiler 50-1-B-1. Tar oil from Unit 24 also flows to pressurized boiler 50-1-B-1.

The fuel gas, flash gas, and tar oil streams are combusted in the pressurized boiler 50-1-B-1 to produce superheated high pressure steam. Flue gases exhaust from the pressurized boiler at 1742°F and 120 psig and enter gas turbine 50-1-GT-1 for expansion to atmospheric pressure. The hot turbine exhaust gases then flow through economizers 50-1-E-8, 50-1-E-9, and 50-1-E-10 to preheat boiler feedwater flowing to the pressurized boiler 50-1-B-1, and finally to the atmosphere through a tall stack.

Gas turbine 50-1-GT-1 drives air compressor 50-1-C-1 which is integral to the turbine, and also electric generator 50-1-G-1. Air from 50-1-C-1 flows to the pressurized boiler for combustion of fuel gas. A portion of this air is split-off and flows to the oxidant supply system in Unit 11.

Equipment Notes

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


22-1-C-1
FLASH GAS COMPRESSOR
P = 3,3MW
50-1-GT-1
GAS TURBINE
50-1-B-1
PRESSURIZED BOILER
50-DA-1
DEAERATOR
50-1-T-1
HP/IP TURBINE
50-1-T-2
LP TURBINE
TAR OIL
FLASH GAS
FUEL-GAS
11-1-E-3
FUEL GAS PREHEATER
11-1-EX-1
Gas-expansion turbine
20-1-R-1
Lurgi Gasifier
JACKET DUTY = 51,085 MW
174,425 MMBTU/h
30-1-B-1
Tar Boiler
E2
SUPERHEATER
E1
EVAPORATOR
50-1-P-2
50-1-B-1; E-3
REHEATER COIL
11-1-E-2
AIR COMPRESSOR AFTERCOOLER
11-1-C-1
BOOSTER AIR COMPRESSOR
11-1-E-1
AIR COMPRESSOR SUCTION COOLER
50-1-E-8
50-1-G-1
50-1-GT-1
Water Injection
30-1-P-3
AIR
50-1-E-7
50-1-E-9
50-1-E-6
50-1-P-1
P = 16,52 MW
23-1-H-1
Sulfur Furnace
30-1-P-2
80 KW
50-DA-1
50-1-E-10
50-1-E-4
50-1-E-3
50-1-E-2
50-1-E-1
50-1-P-4
27 KW
50-1-G-2
FLOW LEGEND
BFW AND CONDENSATE
STEAM
AIR
FUEL GAS
TAR OIL
23-1-E-1(2
Sulfur Condenser
Sulfur Melting Unit 23
Process Exchanger I Unit 26
discharged
21-1-E-1
50 PSIG STM.GEN.
21-1-E-3
COLD COND HTR
FLASH DRUM
Steam Tracing
Process Exchanger II Unit 26
turbine seal steam
30-TK-1
DEMINERALIZED WATER TANK
30-P-1A&B
50-1-E-5
SURFACE CONDENSER
CWS
CWR
50-1-P-3
P = 1,40 MW
2.5" Hg.
CONVERSION FACTORS
0,06896 BAR = 1 PSIA
0,126111 Kg/s = 10³ LB/H
9/5°C + 32 = 1 °F
1,055073 KJ = 1 BTU
0,293075 MJ/s = 10⁶ BTU/H
2,32395 KJ/Kg = 1 BTU/LB
0,746 KW = 1 BHP
NOTES:
1. THIS DIAGRAM SHOWS ONE OF TWO PARALLEL POWER GENERATION TRAINS (KOMBI - BLOCKS).
2. ALL RATES SHOWN ARE FOR FULL DESIGN CAPACITY FOR THE ENTIRE PLANT AND ARE NOT ON A PER TRAIN BASIS.
3. THE COMMAS IN THE NUMBERS ON THIS DRAWING ARE EQUIVALENT TO DECIMAL POINTS.
STEAG AG
Anlagentechnik
EPRI - Fluor Study
PROCESS FLOW DIAGRAM
STEAM SYSTEM - POWER GENERATION
LURGI PROCESS - AIR BLOWN
DWG. NO. MASW-50-1
17 11 1264 / 8.004 B
FLUOR REV. 01 11-5-79

PROCESS DISCUSSION - CASE MASW

The table below summarizes pertinent heat and material balance results.

Table 4-2

SUMMARY OF SYSTEM PERFORMANCE - CASE MASW

GASIFICATION AND GAS CLEANING SYSTEM	
Coal Feed Rate, lbs/hr (m.f.)	1,014,814
Oxidant*/Coal Ratio, lbs/lb m.f.	1.562
Oxidant Temperature, °F	319
Steam/Coal Ratio, lbs/lb m.f.	0.784
Gasifier Exit Pressure, psig	336
Crude Gas Temperature, °F	861
Crude Gas HHV (dry basis). Btu/SCF**	189.1
Temperature of Fuel Gas to Power System, °F	179
POWER SYSTEM	
Gas Turbine Inlet Temperature, °F (at pressurized boiler exit)	1,742
Turbine Air Compressor Pressure Ratio	9.7:1
Gas Turbine Exhaust Temperature, °F	950
Steam Cycle Conditions, psig/°F/°F (at turbine inlet)	2,669/977/977
Condenser Pressure, inches Hg abs.	2.5
Power Block Stack Temperature, °F	248
Gas Turbine Power, MW#	215
Steam Turbine Power, MW#	717
Power Consumed, MW	67
Net System Power, MW	865
OVERALL SYSTEM	
Process and Deaerator Makeup Water, gpm/1000 MW, Net	2,750
Cooling Tower Makeup Water, gpm/1000 MW, Net	9,491
Cooling Water Circulation Rate, gpm/MW	389
Cooling Tower Heat Rejection, % of Coal HHV	39.4
Air Cooler Heat Rejection, % of Coal HHV	6.2
Net Heat Rate, Btu/kWh	11,149
Overall System Efficiency (Coal→Power), % of Coal HHV	30.59

*Oxidant air on dry basis

**Excluding HHV of H_2S, COS and NH_3

#At generator terminals

GASIFIER MATERIAL BALANCE

Table 4-3 presents the gasifier material balance for Case MASW.

This case is based on using U.S. western coal in Lurgi type gasifiers. U.S. western coal is particularly suitable for the Lurgi process, 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 note that the other cases studied are based on an Illinois No. 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 MASW 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 tar is burned in the by-product boilers (equipped with stack gas scrubbers and electrostatic precipitators) to generate high-pressure steam which is integrated with the steam power systems. Oils are separated from the tar and burned separately as fuel in the supercharged boilers. For a final design it would be necessary to evaluate this approach compared to recycling all by-product liquid hydrocarbons into the gasifiers. STEAG has practiced this latter approach at their gasification plant at Lünen, West Germany. This method is practiced in a way that cleans heavy tars and particulates from the gasifier effluent with the recycled by-product liquids. This latter approach minimizes fouling of downstream equipment, but would require testing and confirmation of suitable results with U.S. coals.

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 4-3

MATERIAL BALANCE - CASE MASW

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	319		
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	530		
Jacket		191,657	10,638.1
Makeup		604,383	33,546.9
TOTAL STEAM		796,040	44,185.0
TOTAL FEEDS		3,554,815	

EFFLUENTS

	psig	T (°F)	lb/hr	lb mol/hr	mol % (wet)
Gasifier Effluent	340	861			
CH_4			81,498	5,080.0	3.64
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	17.33
CO			423,069	15,104.0	10.82
CO_2			747,160	16,977.20	12.10
H_2S			10,535	309.2	0.22
COS			-	-	-
NH_3			10,701	628.4	0.45
N_2			1,216,408	43,422.4	31.10
H_2O			599,605	33,283.1	23.83
Subtotal			3,156,172	139,627.5	100.00
N+T+O*			81,153		
P+O**			5,315		
TOTAL GASIFIER EFFLUENT			3,215,640		
Ash		900			
Carbon			15,642		
Ash			296,533		
Subtotal			312,175		
TOTAL EFFLUENTS			3,554,815		

*Naphtha, tars, oils
**Phenols, others

	Wt%*	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

PROCESS ENERGY BALANCES

Table 4-4 presents the overall process energy balance at 100 percent capacity operation. The boundary for the balance encompasses the entire plant, exclusive of the cooling tower heat balance. 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 0.5 percent. The discrepancies result from approximations used for some process units and for calculating some heat loads.

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

As shown in Table 4-4, total power generation of 3179 MM Btu/hr is obtained in this case. Coal charge rate based on high heating value (HHV) is 9644 MM Btu/hr. This results in a conversion of 32.96 percent on a gross basis.

As shown in Table 4-2, if all the power consumed in the plant is debited, the system efficiency (net power at 3413 Btu/kWh), as a percent of coal HHV, is 30.60 percent. The heat rate based on net power produced and coal HHV input is 11,149 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: some lean Selexol solvent is used in the naphtha absorber, resulting in slightly higher circulation rates as compared to operation without naphtha in the feed gas.

The major energy losses from the plant are the heat rejected to cooling water in the surface condensers and the heat existing in the power block stack gases. These losses amount to 35.91 percent and 12.46 percent of the coal HHV, respectively.

The liquid hydrocarbon by-product tars are combusted in tar boilers to generate high-pressure superheated steam resulting in a loss of 3.04 percent of coal HHV with the boiler stack gases. The steam from the tar boiler has been added to the other plant steam and used to generate power in the steam turbine generator.

Table 4-4

ENERGY BALANCE - CASE MASW

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		52	126			178
Demineralized Water		12				12
Auxiliary Power Inputs					229	229
TOTAL	9,644	71	126	0	229	10,070
HEAT OUT						
Ash	226	28				254
Gasifier Heat Loss				88		88
Sulfur Product	21					21
Ammonia Product	85					85
Generated Power					3,179	3,179
Power Block Losses*	22			14	53	89
Power Surface Condensers			3,463			3,463
Power Block Stack Loss		427	775			1,202
Gasifier Effluent Cooling			506			506
Selexol Solvent Cooling		177				177
Regenerator Overhead Cooling		113	29			142
Process Condensate Cooling		65				65
Steam Heat Losses		18	2			20
Motor Losses (Air Cooler Fans, etc.)					132	132
Process Condensate Treating Unit			351			351
Waste Water Effluent		24				24
Tar Boiler Stack Loss		260	33			293
Spent Tail Gas Loss		10	8			18
TOTAL	354	1,122	5,167	102	3,364	10,109

$$\frac{\text{Output} - \text{Input}}{\text{Input}} = 0.39\%$$

*Includes mechanical and electrical losses

Table 4-5

ENERGY BALANCE AS PERCENT COAL HHV - CASE MASW

	MM Btu/hr	Percent
IN		
Coal HHV	9,644	100.0
OUT		
Net Power	2,950	30.59
Sulfur Product, HHV	21	0.22
Ammonia Product, HHV	85	0.88
Selexol Sensible and Latent	319	3.31
Ash	254	2.63
Power Block Stack Losses	1,202	12.46
Rejected at Condensers	3,463	35.91
Other Sensible Losses, Net	46	0.48
Other Latent Losses, Net	741	7.68
Gasifier Heat Losses	88	0.91
Power Block Losses	89	0.92
Tar Boiler Stack Loss	293	3.04
Motor Losses	132	1.37
	9,683	100.40

ECONOMICS - CASE MASW

Table 4-6 summarizes the economics of Case MASW.

Table 4-6

SUMMARY OF ECONOMIC RESULTS - CASE MASW

PRODUCTION AT DESIGN CAPACITY	
Net Power, MW*	865
Overall Plant Heat Rate, Btu/kWh*	11,149
TOTAL CAPITAL REQUIREMENT**	
Total Capital Requirement, $1,000	851,105
Total Capital Requirement, $/kW	983.93
AVERAGE COSTS OF SERVICES**	
First Year Cost, $1000/year	240,933
First Year Cost, mills/kWh	45.42
Thirty Year Levelized Cost, $1000/year	321,385
Thirty Year Levelized Cost, mills/kWh	60.59

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

TOTAL PLANT INVESTMENT

Table 4-7 gives a detailed breakdown of the Total Plant Investment required for Case MASW. The accuracy of the plant investment estimate is judged to be ±25 percent. Since other capital charges are keyed to elements of plant investment, this accuracy is reflected in these other capital figures as well. However, due to the similar nature of the cases in this evaluation, the estimates for all

Table 4-7

TOTAL PLANT INVESTMENT - CASE MASW

Plant Section	Cost Breakdown Without Contingencies: Direct Field Material#	Direct Field Labor##	Engrng. & Sprt. Costs§	Total Sales Tax	Total Cost $1,000*	Cost $/kW**	Per-cent	Contingencies: Process $1,000*	Project $1,000*	Total Plant Investment: $1,000*	$/kW**
Coal Handling	8,860	3,510	13,215	454	26,039	30.10	4.25	0	3,906	29,945	34.62
Oxidant Feed	10,005	3,768	6,263	508	20,544	23.75	3.35	0	3,082	23,626	27.31
Gasification and Ash Handling	64,493	11,348	24,260	3,168	103,269	119.39	16.86	5,163	15,490	123,922	143.26
Gas Cooling	9,429	4,161	6,688	487	20,765	24.01	3.39	0	3,115	23,880	27.61
Acid Gas Removal and Sulfur Recovery	19,708	6,737	11,512	1,247	39,204	45.32	6.40	2,107	5,881	47,192	54.56
Process Condensate Treating	29,543	12,179	17,345	1,146	60,213	69.61	9.83	0	9,032	69,245	80.05
Steam, Condensate and B&W	19,997	6,856	10,069	156	37,078	42.86	6.05	4,574	5,562	47,214	54.58
STEAG Cycle	128,579	39,367	70,374	6,626	244,946	283.18	39.99	12,248	36,741	293,935	339.81
General Facilities	29,168	13,279	17,177	918	60,542	69.99	9.88	0	9,081	69,623	80.49
Subtotal	319,782	101,205	176,903	14,710	612,600	708.21	100.00	24,092	91,890	728,582	842.29

TOTAL PLANT INVESTMENT SUMMARY

	$1,000*	$/kW**
Process Plant Investment and General Facilities	612,600	708.21
Process Contingency	24,092	27.85
Project Contingency	91,890	106.23
Total Plant Investment	728,582	842.29

*Mid-1976 dollars
**Based on 100 percent plant design power output (865 MW)
#All materials and equipment that become a part of the plant facility
##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)
§Includes: a) Indirect field costs including all labor, supervision and expenses required to support field construction; b) Home office costs including all salaries and expenses required for engineering design and procurement; and c) Contractor's fee

cases should reflect about the same accuracy. When these estimates are used comparatively, the effect of individual accuracies should be minimal.

Two contingencies are included in the Total Plant Investment shown for each plant section. 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 innovative technology in an effort to quantify the uncertainty in the design, performance, and cost of the commercial scale equipment. Process contingencies are generally minimal for Case MASW because nearly all the process technology has been demonstrated by commercial scale operations.

TOTAL CAPITAL REQUIREMENT

Table 4-8 gives a breakdown of the Total Capital Requirement to place the plant into initial operation at 100 percent capacity. Starting with the total plant investment, capital charges are added for allowance for funds during construction, initial catalyst and chemicals charge, prepaid royalties, preproduction costs, the inventory capital, and the land required for the plant. Specific items included in each of these capital charges are described under the economic criteria.

COST OF SERVICES

Table 4-9 gives a cost of services breakdown for Case MASW. The costs are busbar power costs based on plant operation at a 70 percent capacity factor with a $1/MM Btu coal cost. Table 4-9 shows both first year power costs and 30 year levelized power costs. Since the power costs vary with the plant capacity factor, this additional relationship is shown in Figure 4-1 on both a first-year and 30-year levelized basis. The plant capacity factor presented in Figure 4-1 is the percent plant time at baseload (100 percent plant power output of 865 MW). This capacity factor is not equivalent to plant operation in a turndown mode producing less than baseload power output.

The largest single component of the power cost is the levelized fixed charges. These charges represent repayment from operating revenue of the financing originally used to supply the total capital required for construction and startup of the plant. These levelized fixed charges amount to 64 percent of the first year power cost at a 70 percent capacity factor. The charges shown are fixed annual expenses which are independent of the plant capacity factor.

Table 4-8

TOTAL CAPITAL REQUIREMENT - CASE MASW

	$1000*	$/kW**
TOTAL PLANT INVESTMENT	728,582	842.29
CAPITAL CHARGES		
Prepaid Royalties	2,760	3.19
Preproduction Costs	19,203	22.20
Inventory Capital	7,088	8.19
Initial Catalyst and Chemical Charge	1,472	1.70
Allowance for Funds During Construction	91,000	105.20
Land	1,000	1.16
Total Capital Charges	122,523	141.64
TOTAL CAPITAL REQUIREMENT	851,105	983.93

*Mid-1976 dollars

**Based on 100 percent plant design power output

Coal constitutes the major operating charge. This charge amounts to 25 percent of the first-year power cost at a 70 percent capacity factor.

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

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

Maintenance labor and material costs are calculated for each major plant section as a percentage of the Total Plant Investment for that section. These percentages are listed in the Economic Criteria section.

Table 4-9

BUSBAR POWER COST AT 70 PERCENT CAPACITY FACTOR - CASE MASW

	Air-Blown Lurgi MASW	
NET PRODUCTION*		
Net Power, MW	865	
By-product Ammonia, ST/D	128	
By-product Sulfur, ST/D	63	
TOTAL CAPITAL REQUIREMENT, $1,000	851,105	
	First Year Cost	30 Year Levelized Cost
FIXED OPERATING COST, $1,000/YEAR		
Operating Labor	3,942	7,435
Maintenance Labor	7,405	13,966
Maintenance Materials	11,108	20,950
Administrative and Support Labor	3,404	6,420
Total Fixed O&M Costs	25,859	48,771
VARIABLE OPERATING COSTS (EXCLUDING COAL), $1,000/YEAR		
Raw Water	1,779	3,355
Catalysts and Chemicals	408	769
Ash Disposal	3,826	7,216
Total Variable O&M Costs	6,013	11,340
COAL COST, $1,000/YEAR	59,132	114,243
BY-PRODUCT CREDITS, $1,000/YEAR		
By-product Ammonia	(3,270)	(6,168)
By-product Sulfur	-0-	-0-
Total By-product Credits	(3,270)	(6,168)
TOTAL OPERATING COSTS, $1,000/YEAR	87,734	168,186
LEVELIZED FIXED CHARGES, $1,000/YEAR	153,199	153,199
TOTAL COST OF ELECTRICITY**		
$1,000/year	240,933	321,385
mills/kWh	45.42	60.59

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

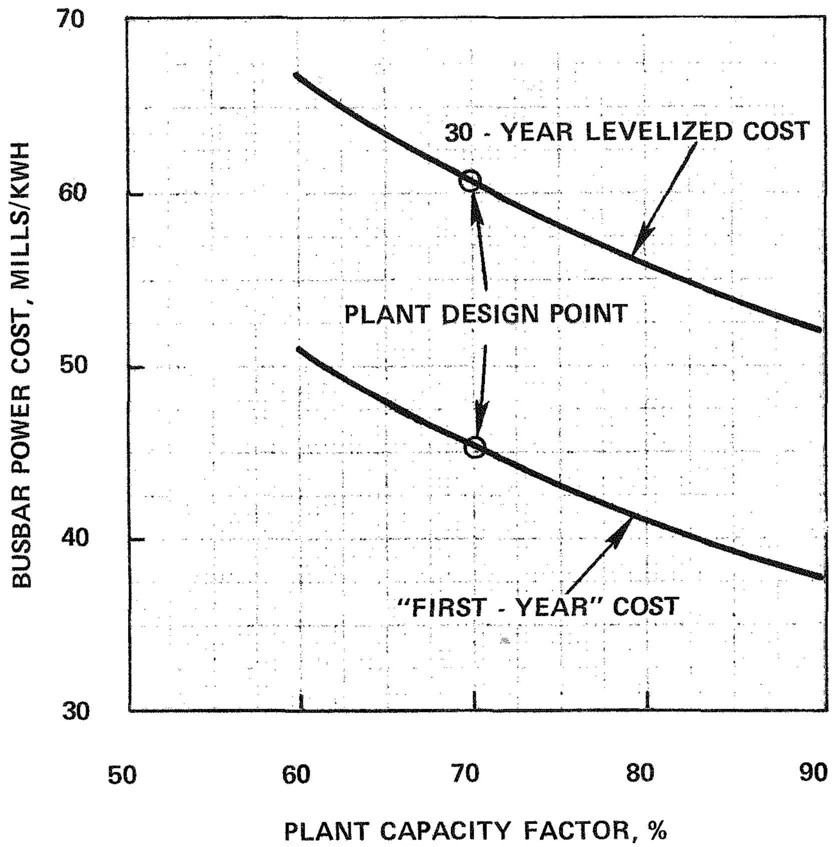


Figure 4-1. Power Cost Sensitivity Curve for Case MASW

Catalyst and chemical costs are associated with those chemicals consumed in the demineralizer, cooling tower, and boiler feedwater treating, plus the 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.

Although no by-product credit is assigned for sulfur, Case MASW also produces 128 ST/D of by-product ammonia. The ammonia by-product is credited at $100/ST for the first year's operations.

REFERENCES

1. Transwestern Pipeline Company, et al., "Amended 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.

Section 5

PLANT DESCRIPTION - CASE EXTS

GENERAL

A grass roots plant for electric power generation in a STEAG cycle integrated with single stage, entrained-bed, oxygen-blown gasifiers of the Texaco type is shown schematically on the Block Flow Diagram EXTS-1-1. This plant consumes 10,000 ST/day of Illinois No. 6 coal.

The Case EXTS plant differs from the Texaco-based GCC plant, presented as Case EXTC (Slurry Feed) in EPRI Report AF-642, primarily in that the STEAG pressurized boiler combined cycle is used for generation of electric power. A STEAG combined-cycle system is composed of the following major units:

- A clean fuel gas expander
- A pressurized boiler for high-pressure steam generation, super-heating and reheating
- An air compressor which serves the pressurized boiler
- A pressurized boiler flue gas expander (turbine) which drives both the boiler air compressor and a generator
- A system for heat recovery from the flue gas leaving the expander
- A steam turbine-generator

As indicated on Block Flow Diagram EXTS-1-1, the Case EXTS power generation section contains two identical STEAG combined-cycle systems in parallel. This power block process arrangement is different from the Case EXTC configuration which used seven parallel gas turbines and accompanying Heat Recovery Steam Generators (HRSG's) plus a single common steam turbine-generator unit. The remaining process units in Case EXTS are relatively similar to their counterparts in Case EXTC.

Table 5-1 summarizes major equipment sections in the plant and shows the numbers of operating and spare trains. The train count listed is the same as in Case EXTC except as noted.

The following sections of this report contain descriptions of the major process units in the Case EXTS plant. Emphasis is placed on describing the differences in process design between the Case EXTS units and the units shown for Case EXTC (Slurry Feed) in EPRI Report AF-642.

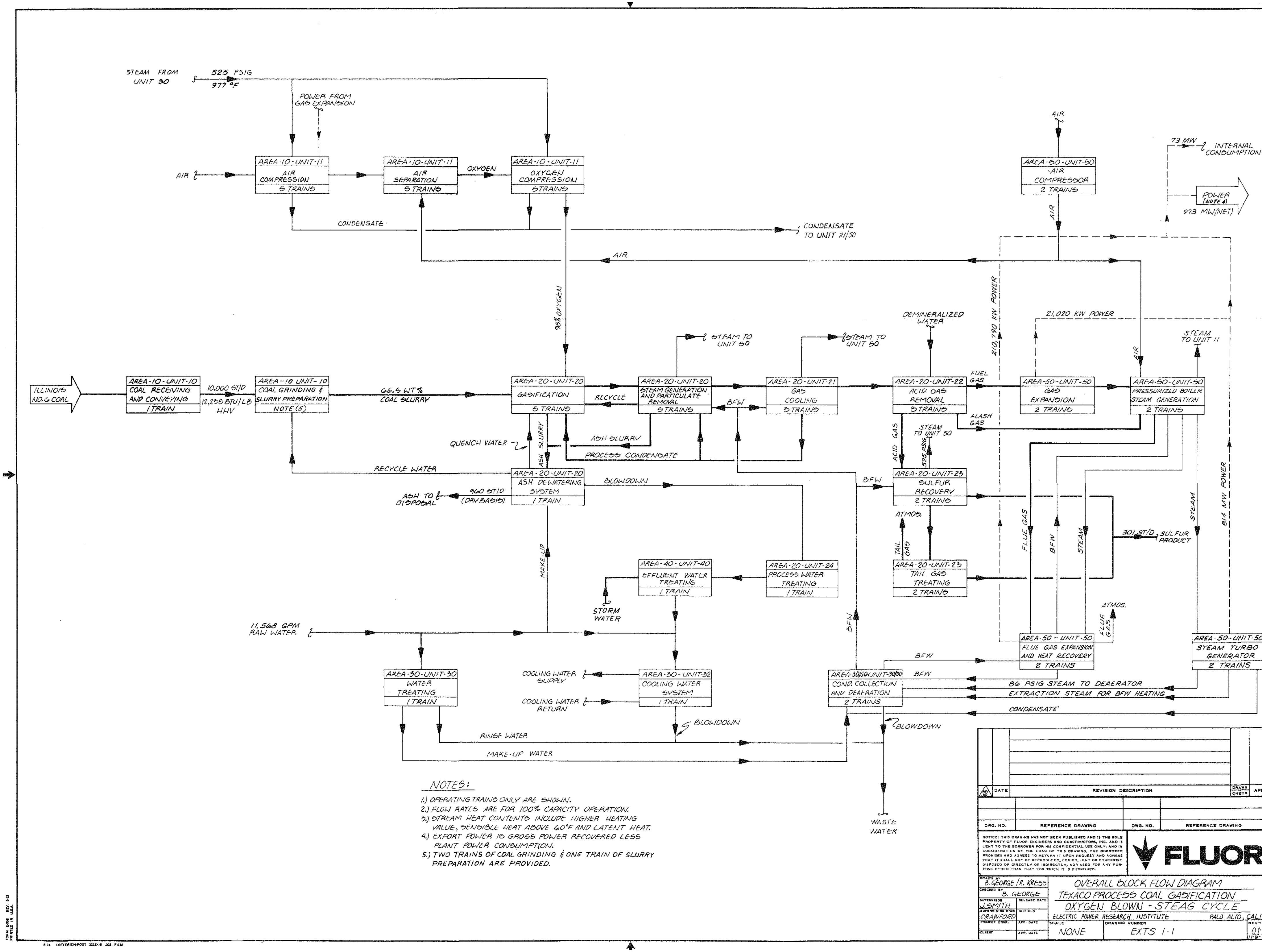
Table 5-1

TRAINS OF EQUIPMENT IN MAIN PROCESSING UNITS - CASE EXTS

Unit No.	Unit Name	Case EXTS Operating	Case EXTS Spare
10	Coal Handling	1	0
11	Oxidant Feed	5	0
20	Wet Coal Grinding	2	0
20	Slurry Preparation	1	0
20	Gasification	5	1
20	Ash Handling	1	0
20	Particulate Scrubbing	5	1
21	Gas Cooling	3	0
22	Acid Gas Removal	3	0
23	Sulfur Recovery	2	1
23	Tail Gas Treating	2	1
30	Steam, BFW and Condensate System		
	• Condensate Collection and Deaeration	2*	0
	• Water Treating	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0
50	STEAG Power Generation System	2**	0

*The train count increased from the one train in Case EXTC to the two listed to match the two STEAG power generation trains

**These two STEAG power generation trains replace the seven gas-turbine-heat-recovery-steam-generators of Case EXTC


STEAM FROM UNIT 50
525 PSIG
977 °F
POWER FROM GAS EXPANSION
AIR
AREA-10-UNIT-11 AIR COMPRESSION 5 TRAINS
AREA-10-UNIT-11 AIR SEPARATION 5 TRAINS
OXYGEN
AREA-10-UNIT-11 OXYGEN COMPRESSION 5 TRAINS
CONDENSATE
CONDENSATE TO UNIT 21/50
AIR
98% OXYGEN
AREA-50-UNIT-50 AIR COMPRESSOR 2 TRAINS
73 MW INTERNAL CONSUMPTION
POWER (NOTE 4)
973 MW(NET)
ILLINOIS NO.6 COAL
AREA-10-UNIT-10 COAL RECEIVING AND CONVEYING 1 TRAIN
10,000 ST/D
12,255 BTU/LB HHV
AREA-10 UNIT-10 COAL GRINDING & SLURRY PREPARATION NOTE (5)
66.5 WT % COAL SLURRY
AREA-20-UNIT-20 GASIFICATION 5 TRAINS
RECYCLE
STEAM TO UNIT 50
AREA-20-UNIT-20 STEAM GENERATION AND PARTICULATE REMOVAL 5 TRAINS
BFW
AREA-20-UNIT-21 GAS COOLING 5 TRAINS
DEMINERALIZED WATER
AREA-20-UNIT-22 ACID GAS REMOVAL 5 TRAINS
FUEL GAS
FLASH GAS
210,790 KW POWER
21,020 KW POWER
AREA-50-UNIT-50 GAS EXPANSION 2 TRAINS
STEAM TO UNIT 11
AREA-50-UNIT-50 PRESSURIZED BOILER STEAM GENERATION 2 TRAINS
QUENCH WATER
ASH SLURRY
PROCESS CONDENSATE
RECYCLE WATER
AREA-20-UNIT-20 ASH DE-WATERING SYSTEM 1 TRAIN
BLOWDOWN
ASH TO DISPOSAL
960 ST/D (DRY BASIS)
ACID GAS
STEAM TO UNIT 50
525 PSIG
AREA-20-UNIT-23 SULFUR RECOVERY 2 TRAINS
ATMOS.
TAIL GAS
AREA-20-UNIT-23 TAIL GAS TREATING 2 TRAINS
301 ST/D SULFUR PRODUCT
FLUE GAS
BFW
STEAM
814 MW POWER
MAKE-UP
AREA-40-UNIT-40 EFFLUENT WATER TREATING 1 TRAIN
AREA-20-UNIT-24 PROCESS WATER TREATING 1 TRAIN
STORM WATER
ATMOS.
FLUE GAS
11,568 GPM RAW WATER
AREA-50 - UNIT-50 FLUE GAS EXPANSION AND HEAT RECOVERY 2 TRAINS
AREA-50-UNIT-50 STEAM TURBO GENERATOR 2 TRAINS
AREA-30-UNIT-30 WATER TREATING 1 TRAIN
COOLING WATER SUPPLY
COOLING WATER RETURN
AREA-30-UNIT-32 COOLING WATER SYSTEM 1 TRAIN
AREA-30/50-UNIT-30/50 COND. COLLECTION AND DEAERATION 2 TRAINS
BFW
86 PSIG STEAM TO DEAERATOR
EXTRACTION STEAM FOR BFW HEATING
CONDENSATE
BLOWDOWN
RINSE WATER
MAKE-UP WATER
WASTE WATER
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.
REV NO
DATE
REVISION DESCRIPTION
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CHECK
APP
DWG. NO.
REFERENCE DRAWING
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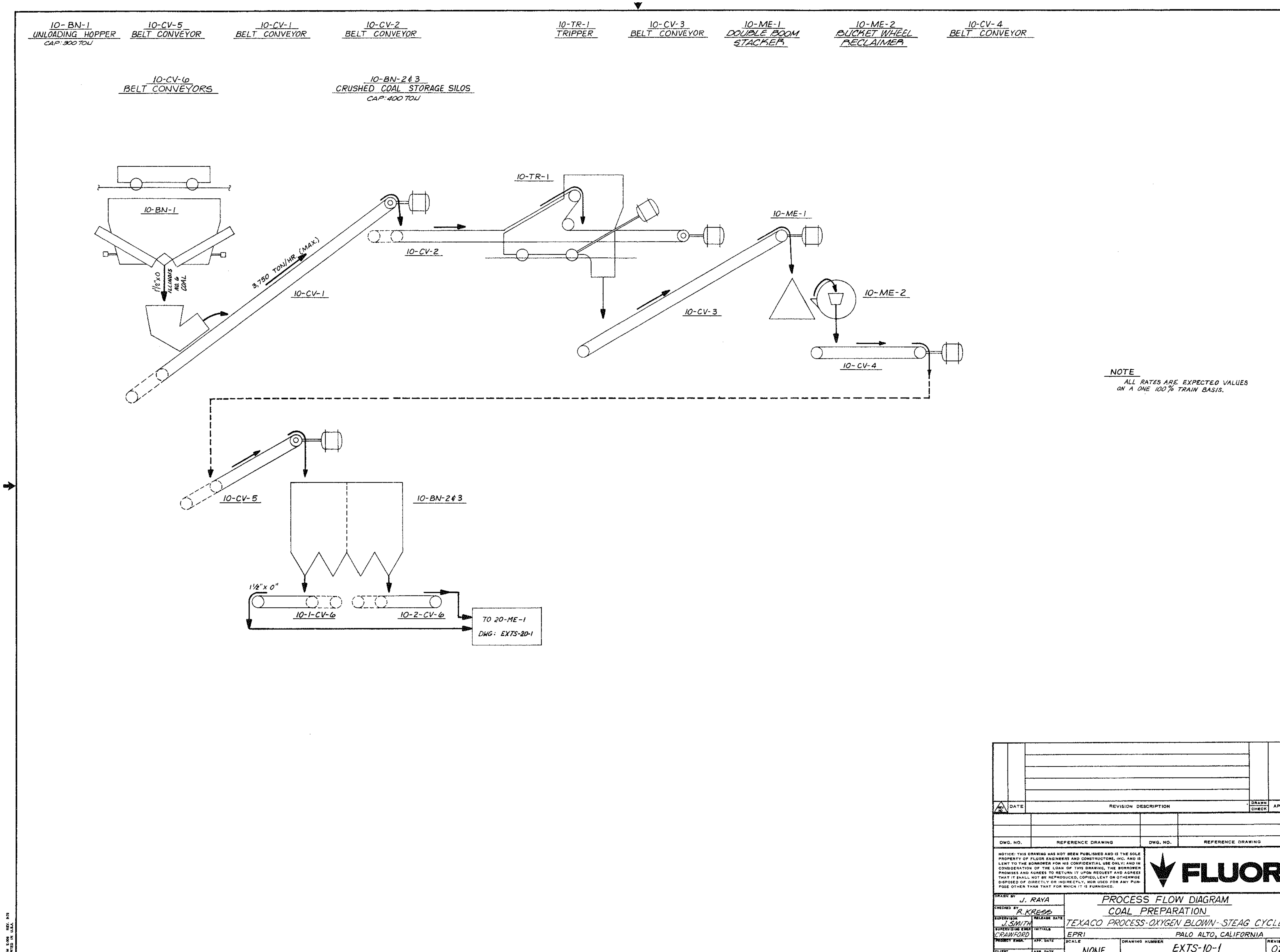
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE
OVERALL BLOCK FLOW DIAGRAM
TEXACO PROCESS COAL GASIFICATION
OXYGEN BLOWN - STEAG CYCLE
ELECTRIC POWER RESEARCH INSTITUTE
PALO ALTO, CALIF.
SCALE NONE
DRAWING NUMBER EXTS 1-1
REVISION 01 11-6-79

COAL PREPARATION

Process Flow Diagram EXTS-10-1 depicts the process arrangement of equipment in this section for Case EXTS. Coal preparation section process details are essentially the same as those described for Case EXTC (Slurry Feed) in EPRI Report AF-642 (January 1978, page 304).

Equipment Notes

All the equipment in this section 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
10-BN-1
1½"x0 ILLINOIS NO. 6 COAL
3,750 TON/HR (MAX)
10-CV-1
10-CV-2
10-TR-1
10-ME-1
10-ME-2
10-CV-3
10-CV-4
NOTE
ALL RATES ARE EXPECTED VALUES ON A ONE 100% TRAIN BASIS.
10-CV-5
10-BN-2 & 3
1½" x 0"
10-1-CV-6
10-2-CV-6
TO 20-ME-1
DWG: EXTS-20-1
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RELEASE DATE
SUPERVIZING ENGR CRAWFORD
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE
PROCESS FLOW DIAGRAM
COAL PREPARATION
TEXACO PROCESS-OXYGEN BLOWN-STEAG CYCLE
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EXTS-10-1
REVISION 02 11-6-79

OXIDANT FEED

Process Flow Diagram EXTS-11-1 shows the gasifier oxidant feed system for Case EXTS. Two major differences exist between the Case EXTS system and the system provided for Case EXTC (Slurry Feed) and described in EPRI Report AF-642 (January 1978, page 307).

Case EXTS differs from Case EXTC in that a part of the oxidant feed system air compression requirement is supplied in Power Generation Unit 50 for Case EXTS. The Power Generation Unit exports compressed air to Unit 11 because the Combustion Air Compressor 50-1-C-1 associated with Gas Turbine 50-1-GT-1 in Unit 50 is a standard model with excess capacity above the requirements of Pressurized Boiler 50-1-B-1. The excess air flow is therefore exported to the gasifier oxidant feed system in Case EXTS and serves to reduce the size of Air Compressor 11-1-C-1 compared to its counterpart in Case EXTC.

The second difference between Cases EXTC and EXTS is that the Fuel Gas Expander 50-1-EX-1 in Case EXTS is not coupled to Air Compressor 11-1-C-1, but instead drives an electric generator. Thus, the total compressor power requirement will be supplied by Air Compressor Turbine 11-1-T-1. This results in a simpler drive train.

Equipment Notes

All equipment is commercially available.

11-1-E-2
AIR INTERCOOLER
34.23 MM BTU/HR
15,000 SQ. FT. BARE
TUBES: CARBON STL.

11-1-E-3
AIR AFTERCOOLER
35.0 MM BTU/HR
9840 SQ FT. BARE
TUBES: CARBON STL.

11-1-C-1
AIR COMPRESSOR
30,725 BHP

11-1-E-1
SURFACE CONDENSER
161 MM BTU/HR
35,860 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

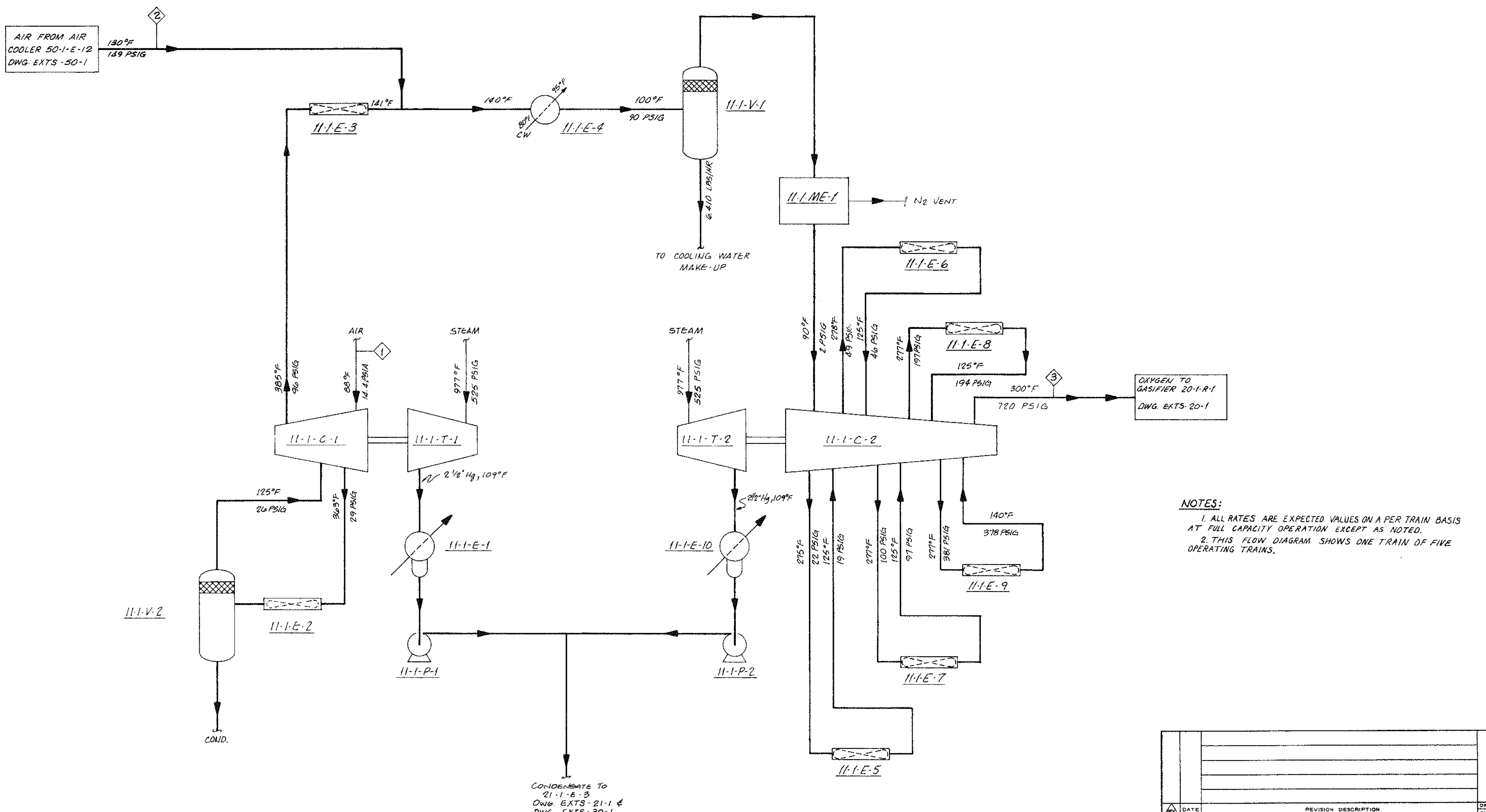
11-1-E-9
5TH STAGE O_2 COOLER
4.4 MM BTU/HR
1,126 SQ. FT. BARE
TUBES: 316 SS

11-1-V-2
K.O. DRUM
156" I.D. X 12'0" T/T
DESIGN : 50 PSIG @ 175°F
CARBON STEEL

11-1-T-1
AIR COMP TURBINE
30,725 BHP

11-1-E-10
SURFACE CONDENSER
61.4 MM BTU/HR
10,750 SQ. FT.
TUBES: CARBON STEEL
SHELL : CARBON STEEL

11-1-T-2
OXYGEN COMP. TURBINE
11,700 BHP



NOTES:
1. ALL RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.
2. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF FIVE OPERATING TRAINS.

MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER: 1 AIR MPH	MOL %	2 AIR MPH	MOL %	3 OXIDANT MPH	MOL %
O_2	4167.0	20.48	459.4	20.48	4279.5	98.00
N_2	15,477.6	76.07	1706.6	76.07	66.0	1.50
AR	199.4	0.98	22.0	0.98	21.7	0.50
H_2O	502.4	2.47	55.5	2.47	——	——
TOTAL MPH	20,346.5	100.00	2,243.5	100.00	4,367.2	100.00
LB/HR	583,924		64,386		139,650	
MOL. WT.	28.70		28.70		31.98	

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CHECKED BY J. RAYA
SUPERVISOR J. SMITH
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SUPERVISING ENGR CRAWFORD
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PROJECT ENGR.
APP. DATE
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PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
TEXACO PROCESS-OXYGEN BLOWN-STEAG CYCLE
EPRI PALO ALTO, CALIFORNIA

SCALE NONE
DRAWING NUMBER EXTS-11-1
REVISION 02 11-6-79

GASIFICATION AND ASH HANDLING

Process Flow Diagram EXTS-20-1 shows the gasification step for Case EXTS. The details of gasification and ash handling are similar to Case EXTC (Slurry Feed) described in EPRI Report AF-642 (January 1978, pages 312 and 313). The steam generation system, however, is different from Case EXTC in that higher steam pressure levels are used.

Raw hot gas from the gasifier is first cooled in a gas cooling unit, 20-1-ME-3. The Gas Cooling Unit 20-1-ME-3 generates high-pressure (HP) steam at 3100 psig, saturated intermediate-pressure (IP) steam at 595 psig and saturated medium pressure (MP) steam at 100 psig. The HP steam and IP steam both flow to steam turbines in the STEAG power generation unit for production of electric power. The MP steam is used both as stripping steam in the deaerator and for various process heating services. Preheated boiler feedwater streams are supplied from the STEAG power generation unit at temperatures approaching saturation for each steam pressure level.

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: EXTS-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 shown in Flow Diagram EXTS-21-1.

Equipment Notes

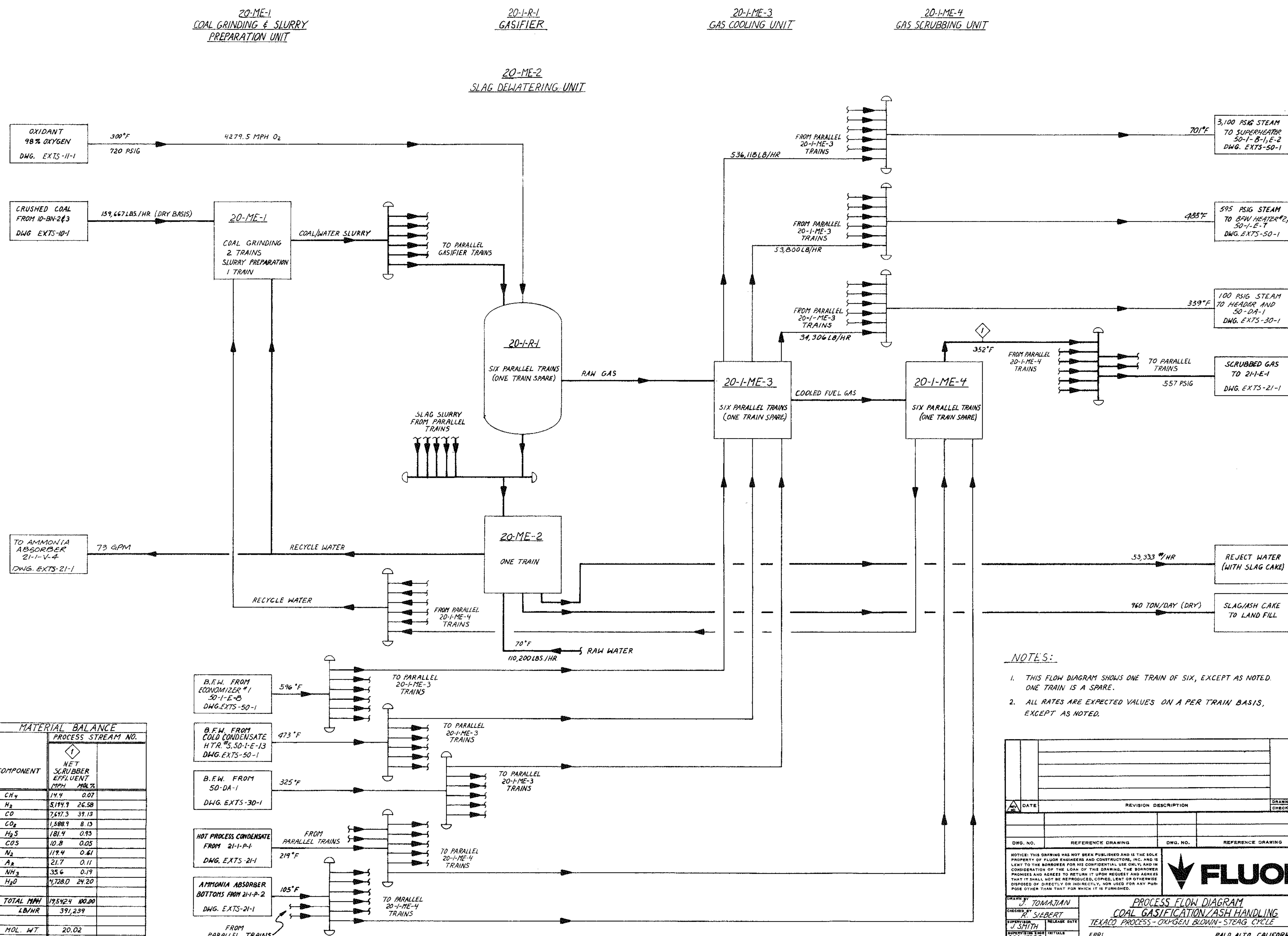
In Case EXTS, the high-pressure steam generation system in the 20-1-ME-3 gas cooling unit has been integrated with the steam cycle in the STEAG power block. This approach to heat integration with the power generation section is analogous to that employed in U.S. combined-cycle-based designs such as Case EXTC78 in this report and Case EXTC of EPRI Report AF-642. Since the saturated steam generated in 20-1-ME-3 is sent to the superheat coil in the STEAG pressurized boiler, the required heat transfer equipment must operate at a steam generation pressure of 3100 psig. Development of this heat transfer equipment will involve a significant extension of the present state of the art for such gas cooling services.

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 currently undergoing test runs in Germany.

The slag dewatering unit is commercially proven.

The gas scrubbing unit equipment is commercially available.



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	
A_R	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		

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NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF SIX, EXCEPT AS NOTED. ONE TRAIN IS A SPARE.
2. ALL RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS, EXCEPT AS NOTED.

DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

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DRAWN BY J. TOMAJIAN
CHECKED BY R. SIEBERT
SUPERVISOR J. SMITH
CRAWFORD

PROCESS FLOW DIAGRAM
COAL GASIFICATION/ASH HANDLING
TEXACO PROCESS - OXYGEN BLOWN - STEAG CYCLE
EPRI — PALO ALTO, CALIFORNIA

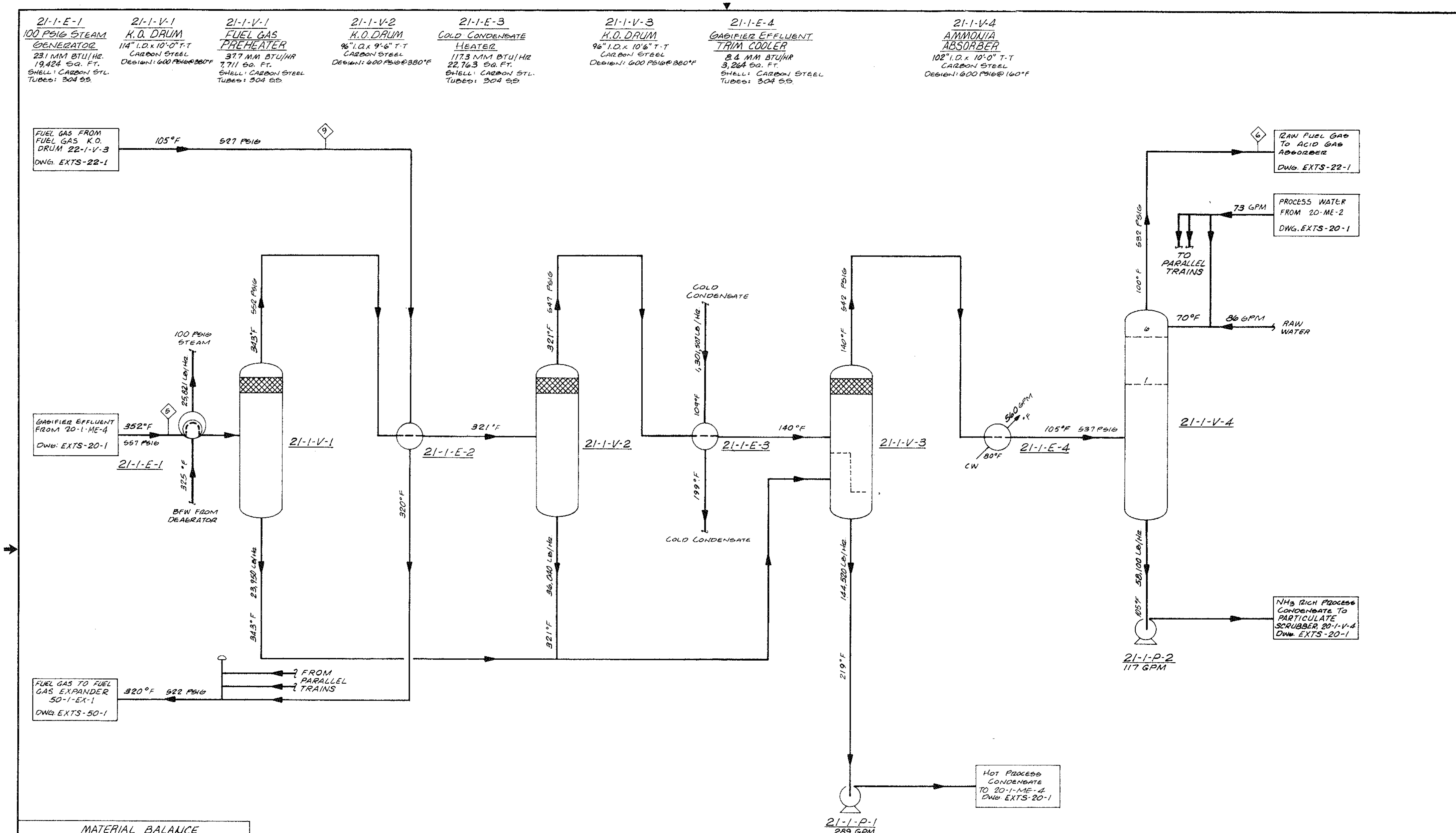
SCALE: NONE | DRAWING NUMBER: EXTS-20-1 | REVISION: 0

GAS COOLING

The process details for each gas cooling train, shown in Process Flow Diagram EXTS-21-1, are similar to those described for Case EXTC (Slurry Feed) in EPRI Report AF-642 (January 1978, page 317). The service for 21-1-E-1 has been changed from 50 psig steam generation to 100 psig steam generation for purposes of heat integration with the Power Generation Unit 50 of Case EXTS. The flow rate of cold condensate through 21-1-E-3 has also been revised slightly to accommodate the integration with the STEAG pressurized boiler combined cycle.

Equipment Notes

All equipment is commercially available.



MATERIAL BALANCE						
	PROCESS STREAM NUMBER					
COMPONENT	5 PARTICULATE SCRUBBER OVHD MPH	MOL%	6 AMMONIA SCRUBBER OVHD MPH	MOL%	9 FUEL GAS MPH	MOL%
CH_4	24.0	0.07	24.0	0.10	23.6	0.10
H_2	8658.2	26.35	8658.2	35.08	8632.4	36.26
CO	12,745.5	38.78	12,745.5	51.64	12,662.5	53.19
CO_2	2,648.2	8.06	2648.2	10.73	2,196.0	9.22
H_2S	303.0	0.92	302.3	1.22	24.0	0.10
COS	18.0	0.06	18.0	0.07	11.6	0.05
N_2	199.0	0.60	199.0	0.81	197.6	0.83
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	24,683.0		23,806.3	100.00
H_2O LIQ	—		—		—	
TOTAL LB/HR	658,100		509,481		477,796	
MOL WT VAPOR	20.02		20.64		20.07	

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11-75 DIETERICH-POST 2111X-8 FILM

NOTES:

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

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DRAWN BY P. STONE
CHECKED BY B. GEORGE
SUPERVISOR J. SMITH
SUPERVISING ENGR CRAWFORD

PROCESS FLOW DIAGRAM
GAS COOLING
TEXACO PROCESS-OXYGEN BLOWN-STEAG CYCLE
EPRI PALO ALTO, CALIFORNIA

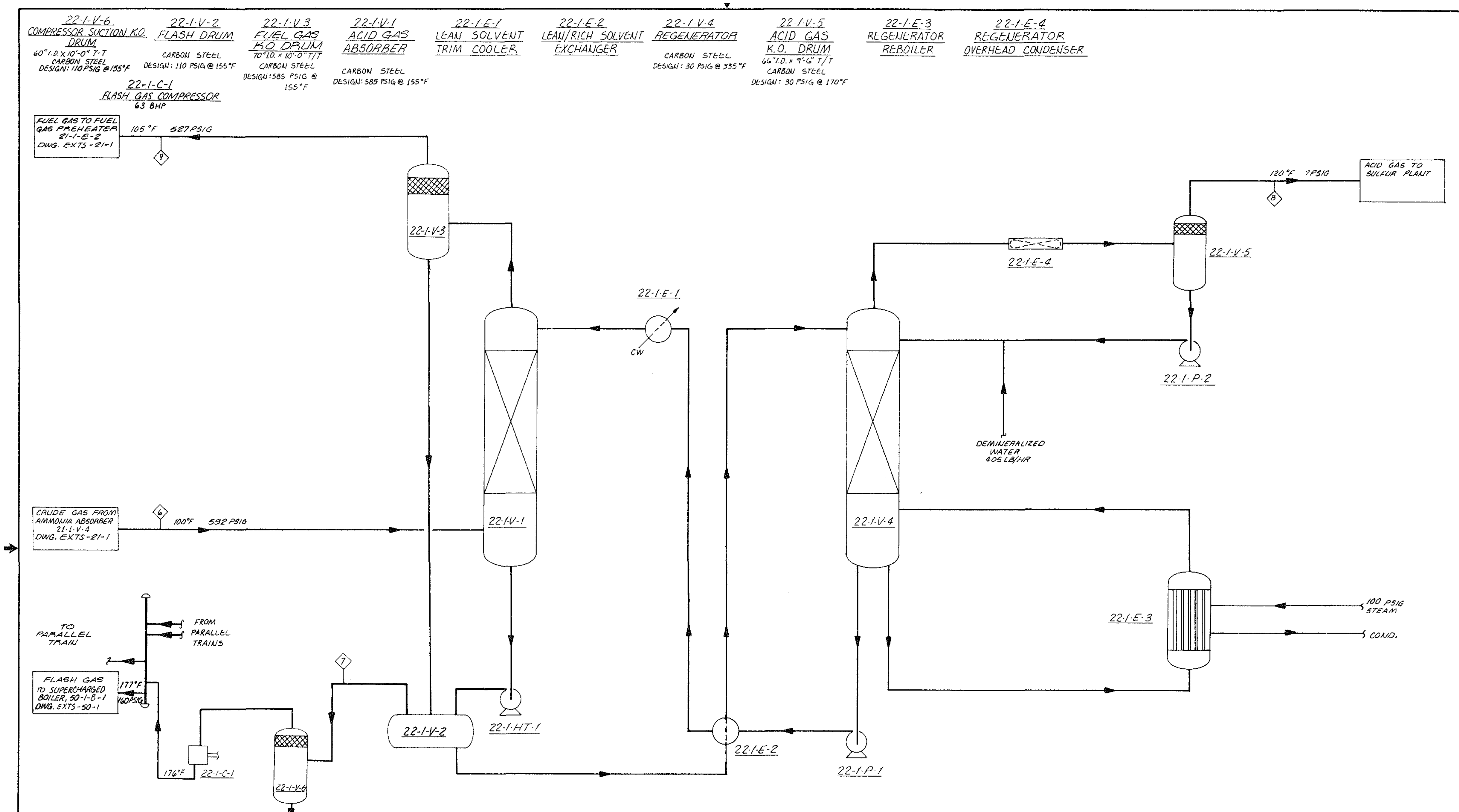
SCALE NONE | DRAWING NUMBER EXTS-21-1 | REVISION 01 11-7-79

ACID GAS REMOVAL

Process Flow Diagram EXTS-22-1 depicts one of the three parallel acid gas removal trains. This unit is similar to its counterpart designed for Case EXTC (Slurry Feed) and described in EPRI Report AF-642 (January 1978, page 321). However, for this case, the flash gas from the Flash Drum 22-1-V-2 is compressed and then combusted with the fuel gas in the STEAG pressurized boiler 50-1-B-1. 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.



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER 6 CRUDE GAS MPH	6 CRUDE GAS MOL%	7 FLASH GAS MPH	7 FLASH GAS MOL%	8 ACID GAS MPH	8 ACID GAS MOL%	9 FUEL GAS MPH	9 FUEL GAS MOL%
CH_4	24.0	0.10	0.3	0.16	0.1	0.01	23.6	0.10
H_2	8658.2	35.08	24.1	12.59	1.7	0.25	8632.4	36.26
CO	12,745.1	51.64	71.9	37.55	10.7	1.59	12,662.5	53.19
CO_2	2648.2	10.73	81.2	42.40	337.2	49.98	2196.0	9.22
H_2S	302.3	1.22	10.9	5.69	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	199.0	0.81	1.3	0.62	0.1	0.01	197.6	0.83
A_R	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	24,683.0	100.00	191.5	100.00	674.7	100.00	23,806.3	100.00
TOTAL LB/HR	510,279		6110		26539		477,796	
MOL WT	20.67		31.90		39.33		20.07	

NOTES:

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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CHECKED BY R. SIEBERT
SUPERVISOR J. SMITH
RELEASE DATE
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PROCESS FLOW DIAGRAM
ACID GAS REMOVAL SYSTEM
TEXACO PROCESS-OXYGEN BLOWN-STEAG CYCLE

EPRI — PALO ALTO, CALIFORNIA

SCALE: NONE
DRAWING NUMBER: EXTS-22-1
REVISION: 02 11-7-79

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SULFUR RECOVERY AND TAIL GAS TREATING

The processes used in these units are the same as for Case MASW. Refer to Case MASW and Process Flow Diagrams MASW-23-1, MASW-23-2 and MASW-23-3. The detailed process descriptions of these units are contained in EPRI Report AF-642 (January 1978, pages 74 through 84).

Equipment Notes

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

PROCESS CONDENSATE TREATING

Most of the sour process condensate generated is used for the preparation of the coal slurry feed to the gasifiers. A small stream of ammonia (as ammonium salts) contaminated effluent is produced in the gasification area (Flow Diagram: EXTS-20-1). This effluent is small and treated in the effluent water treating unit (Unit 40).

A unit for the recovery of by-product ammonia is therefore not provided. The system for treatment of process condensate is similar to that provided for Case EXTC (Slurry Feed) as described in EPRI Report AF-642 (January 1978, page 327).

Equipment Notes

All equipment is commercially available.

STEAM, BOILER FEEDWATER, AND CONDENSATE

Process Flow Diagram EXTS-30-1 schematically represents the steam, boiler feedwater, and condensate systems. This system is significantly different from that presented in EPRI Report AF-642 and is accordingly described in detail below. The description is typical for each of two parallel trains.

Process plant steam generation is closely integrated with the STEAG power generation unit. The integrated system operates at four steam pressure levels:

Turbine Inlet	- 2669 psig (185 bar), 977°F (525°C)
Intermediate Pressure (IP)	- 595 psig (42 bar), at Reheater Inlet
Medium Pressure (MP)	- 100 psig (8 bar)
Low Pressure (LP)	- 50 psig (4.5 bar)

Boiler feedwater is pumped to the STEAG pressurized boiler 50-1-B-1 from the Deaerator 50-DA-1 by High Pressure Boiler Feedwater Pumps 50-1-P-1A&B. A small slipstream of high-pressure boiler feedwater is withdrawn from the discharge of the pumps for steam desuperheating service. The remainder of the boiler feedwater stream is then preheated to 596°F before entering the boiler. This heating occurs in a series of boiler feedwater heaters, 50-1-E-6, 50-1-E-7, 50-1-E-8, and 50-1-E-9. Two boiler feedwater heaters, 50-1-E-7 and E-6 are heated by a slipstream of steam from the 595 psig steam system. The heaters 50-1-E-8 and 50-1-E-9 are economizers which are heated by hot exhaust gas from the turbine 50-1-GT-1. The preheated boiler feedwater at 596°F is then divided into two streams. One stream flows to the evaporator coils of the STEAG pressurized boiler 50-1-B-1; E-1 for generation of saturated 3100 psig steam from the firing of fuel gas. The other stream flows to the Gas Cooling Unit 20-1-ME-3 in the gasification area for the generation of saturated 3100 psig steam from gasifier effluent heat. These two streams of 3100 psig steam then combine in the Pressurized Boiler 50-1-B-1 and are superheated to 986°F in the Superheating Coil 50-1-B-1; E-2. The superheated HP steam then flows to the HP end of Turbine 50-1-T-1 for the generation of electric power. After pressure losses in the Superheating Coil 50-1-B-1; E-2, plus line losses in the interconnecting piping, HP steam enters 50-1-T-1 at 2669 psig and 977°F.

The HP end of the turbine, 50-1-T-1, exhausts steam at 595 psig (42 bar). A portion of this turbine exhaust steam then combines with saturated 595 psig steam produced in the process area. The saturated 595 psig steam from the process area is generated in the Claus plant waste heat boiler 23-1-H-1 and the gas cooling units 20-1-ME-3. Additional 595 psig saturated steam is generated in the air coolers 50-1-E-12 in the combined-cycle area. A large stream of saturated 595 psig steam is diverted for boiler feedwater heating in 50-1-E-7. The remaining 595 psig steam is reheated to 977°F in the reheat coils of the STEAG Pressurized Boiler 50-1-B-1; E-3. The reheated intermediate-pressure (IP) steam then flows to the IP end of turbine 50-1-T-1.

The IP end of turbine 50-1-T-1 exhausts at 85 psig (6.9 bar) and flows to low-pressure turbine 50-1-T-2 for generation of electric power. The LP end of 50-1-T-1 is a condensing turbine exhausting at 2.5 inches Hg abs.

The 100 psig steam header is supplied by steam generation in the gas cooling units, 20-1-ME-3 and 21-1-E-1. This 100 psig steam is used in the Selexol Regenerator Reboiler, 22-1-E-3, and in the sulfur heater, 23-1-EJ-1. The remainder flows to the Deaerator 50-DA-1 as stripping steam.

The 50 psig steam header is supplied from the 50 psig Steam Generator in the Claus plant (Unit 23). The 50 psig steam is mainly used for steam tracing and also in the sulfur pit located in Unit 23.

Condensate from steam users throughout the plant is returned to the Deaerator, 50-DA-1. This includes vacuum condensate (2.5 inches Hg abs.) from the surface condensers coupled to condensing steam turbines 11-1-T-1, 11-1-T-2 and 50-1-T-2. Condensate pumps located at the discharge of each surface condenser pump a part of the vacuum condensate through Condensate Heaters 50-1-E-1, 50-1-E-2, 50-1-E-3, and 50-1-E-4 and subsequently back to the deaerator. Part of the vacuum condensate is heated in 21-1-E-3 before returning to the deaerator. The remainder of the vacuum condensate is pumped as feedwater to the Air Recuperators 50-1-E-11 and 50-1-E-12, and the 50 psig steam generaters in the Unit 23 Claus Plant.

The Deaerator, 50-DA-1, in the STEAG cycle operates at 82 psig. Stripping steam is supplied from the 100 psig steam header. Boiler feedwater from the deaerator is pumped by High-Pressure Boiler Feedwater Pumps (50-P-1A&B) to generation of HP steam as described above. Additional boiler feedwater from the Deaerator 50-DA-1 is pumped by low-pressure boiler feedwater pumps (30-1-P-3A&B) to 20-1-E-3 and 21-1-E-1 for generation of 100 psig steam. A portion of this water flows to

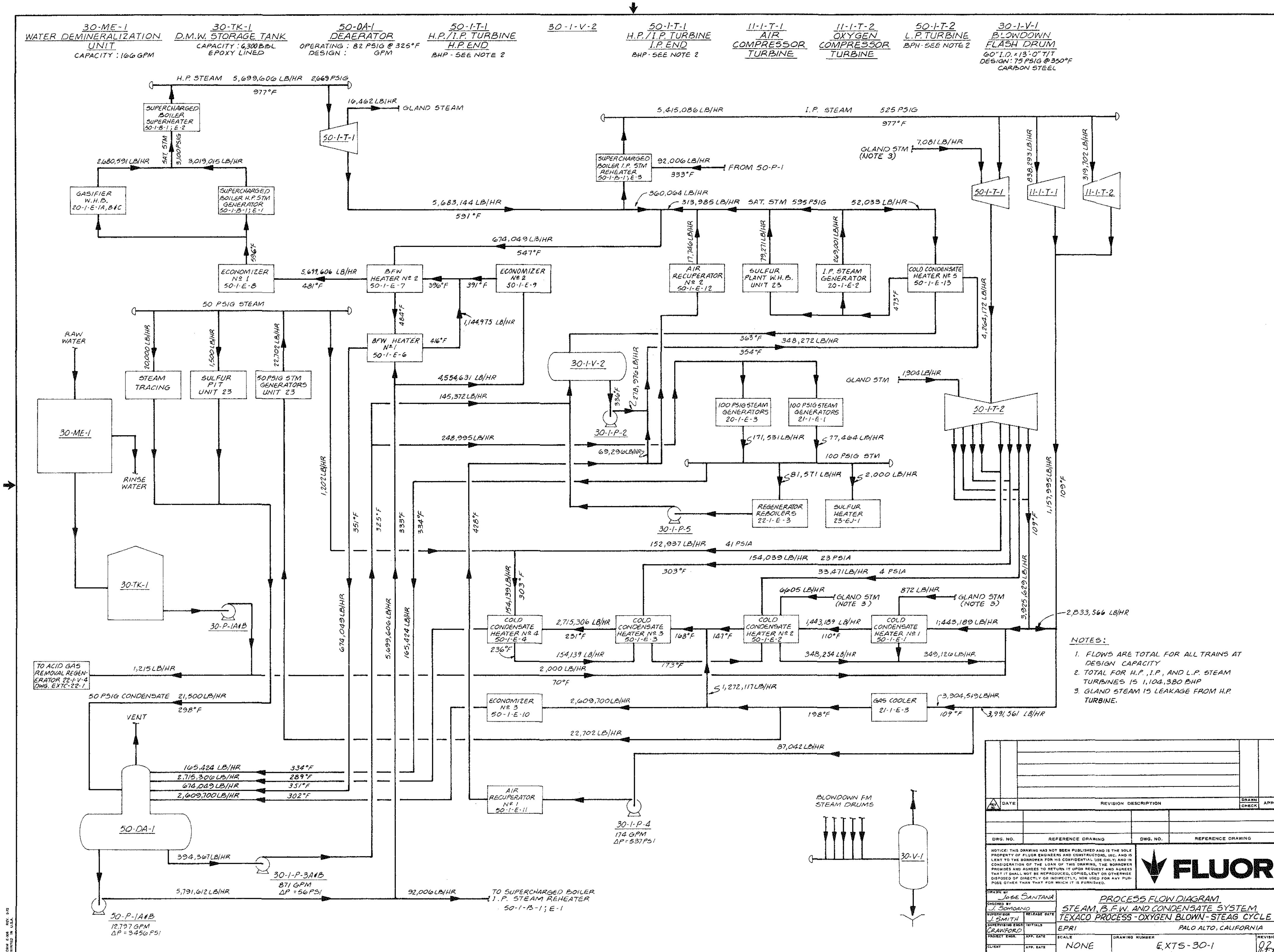
Feedwater Collection Vessel, 30-1-V-2, where it is combined with condensate from the Selexol Regenerator Reboiler 22-1-E-3. The combined condensate streams are then pumped by booster pump 30-1-P-2 to the 595 psig (42 bar) steam generators. These include the gas cooling units, 20-1-ME-3, and the Claus plant, Unit 23.

The recirculating boiler feedwater is continuously treated in a Benson-type water treating system for removal of dissolved salts as well as dissolved oxygen. This water treatment minimizes the makeup and blowdown requirements.

Raw water is treated in a semiautomatic, resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for the high-pressure boilers. Demineralized water is stored in Tank 30-TK-1. Demineralized water from the storage tank is transported to the deaerator through Pumps 30-P-1A&B. A small quantity of the makeup water is withdrawn from the discharge of 30-P-1A&B and transported to Unit 22. The balance of the demineralized water flows to the Deaerator, 50-DA-1.

Equipment Notes

All this equipment is commercially available.



30-ME-1 WATER DEMINERALIZATION UNIT CAPACITY: 166 GPM
30-TK-1 D.M.W. STORAGE TANK CAPACITY: 6,300 BBL EPOXY LINED
50-DA-1 DEAERATOR OPERATING: 82 PSIG @ 325°F DESIGN: GPM
50-1-T-1 H.P./I.P. TURBINE H.P. END BHP - SEE NOTE 2
30-1-V-2
50-1-T-1 H.P./I.P. TURBINE I.P. END BHP - SEE NOTE 2
11-1-T-1 AIR COMPRESSOR TURBINE
11-1-T-2 OXYGEN COMPRESSOR TURBINE
50-1-T-2 L.P. TURBINE BPH - SEE NOTE 2
30-1-V-1 BLOWDOWN FLASH DRUM 60" I.D. x 13'-0" T/T DESIGN: 75 PSIG @ 350°F CARBON STEEL
H.P. STEAM 5,699,606 LB/HR 2,669 PSIG 977°F
I.P. STEAM 5,415,086 LB/HR 525 PSIG 977°F
SAT. STM 595 PSIG
50 PSIG STEAM
100 PSIG STM
GLAND STEAM
SUPERCHARGED BOILER SUPERHEATER 50-1-B-1; E-2
GASIFIER W.H.B. 20-1-E-1A, B & C
SUPERCHARGED BOILER H.P. STM GENERATOR 50-1-B-1; E-1
SUPERCHARGED BOILER I.P. STM REHEATER 50-1-B-1; E-3
ECONOMIZER Nº 1 50-1-E-8
BFW HEATER Nº 2 50-1-E-7
ECONOMIZER Nº 2 50-1-E-9
BFW HEATER Nº 1 50-1-E-6
AIR RECUPERATOR Nº 2 50-1-E-12
SULFUR PLANT W.H.B. UNIT 23
I.P. STEAM GENERATOR 20-1-E-2
COLD CONDENSATE HEATER Nº 5 50-1-E-13
STEAM TRACING
SULFUR PIT UNIT 23
50 PSIG STM GENERATORS UNIT 23
100 PSIG STEAM GENERATORS 20-1-E-3
100 PSIG STEAM GENERATORS 21-1-E-1
REGENERATOR REBOILERS 22-1-E-3
SULFUR HEATER 23-EJ-1
30-1-P-2
30-1-P-5
30-P-1A&B
RAW WATER
RINSE WATER
TO ACID GAS REMOVAL REGENERATOR 22-1-V-4 DWG. EXTC-22-1
50 PSIG CONDENSATE 21,500 LB/HR 298°F
VENT
COLD CONDENSATE HEATER Nº 4 50-1-E-4
COLD CONDENSATE HEATER Nº 3 50-1-E-3
COLD CONDENSATE HEATER Nº 2 50-1-E-2
COLD CONDENSATE HEATER Nº 1 50-1-E-1
ECONOMIZER Nº 3 50-1-E-10
GAS COOLER 21-1-E-5
AIR RECUPERATOR Nº 1 50-1-E-11
30-1-P-4 174 GPM ΔP = 537 PSI
30-1-P-3A&B 871 GPM ΔP = 56 PSI
50-P-1A&B 12,797 GPM ΔP = 3456 PSI
TO SUPERCHARGED BOILER I.P. STEAM REHEATER 50-1-B-1; E-1
BLOWDOWN FM STEAM DRUMS
30-V-1
NOTES:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY
2. TOTAL FOR H.P., I.P., AND L.P. STEAM TURBINES IS 1,104,380 BHP
3. GLAND STEAM IS LEAKAGE FROM H.P. TURBINE.
FLUOR
PROCESS FLOW DIAGRAM
STEAM, B.F.W. AND CONDENSATE SYSTEM
TEXACO PROCESS - OXYGEN BLOWN - STEAG CYCLE
EPRI
PALO ALTO, CALIFORNIA
DRAWN BY JOSE SANTANA
CHECKED BY J. SOMOANO
SUPERVISOR J. SMITH
CRAWFORD
SCALE NONE
DRAWING NUMBER EXTS-30-1

COMBINED-CYCLE SYSTEM

Process Flow Diagram EXTS-50-1 depicts the combined-cycle system designed by STEAG for Case EXTS. The diagram shows the total power block flows and also shows various sections of the gasification plant depicted on other flow diagrams, including the steam system and parts of the oxidant supply system.

The power block is divided into two independent parallel trains each of 50 percent capacity. Each train contains two pressurized boilers, 50-1-B-1, one gas turbine, 50-1-GT-1, one gas turbine generator, 50-1-G-1, one steam turbine, 50-1-T-1&2, and one steam turbine generator, 50-1-G-2. Each train is designated a "Kombi-block." Refer to Appendix A for a detailed description of the pressurized boiler combined-cycle system, including detailed performance information for the power block components, i.e., pressurized boilers, steam turbines, and gas turbines.

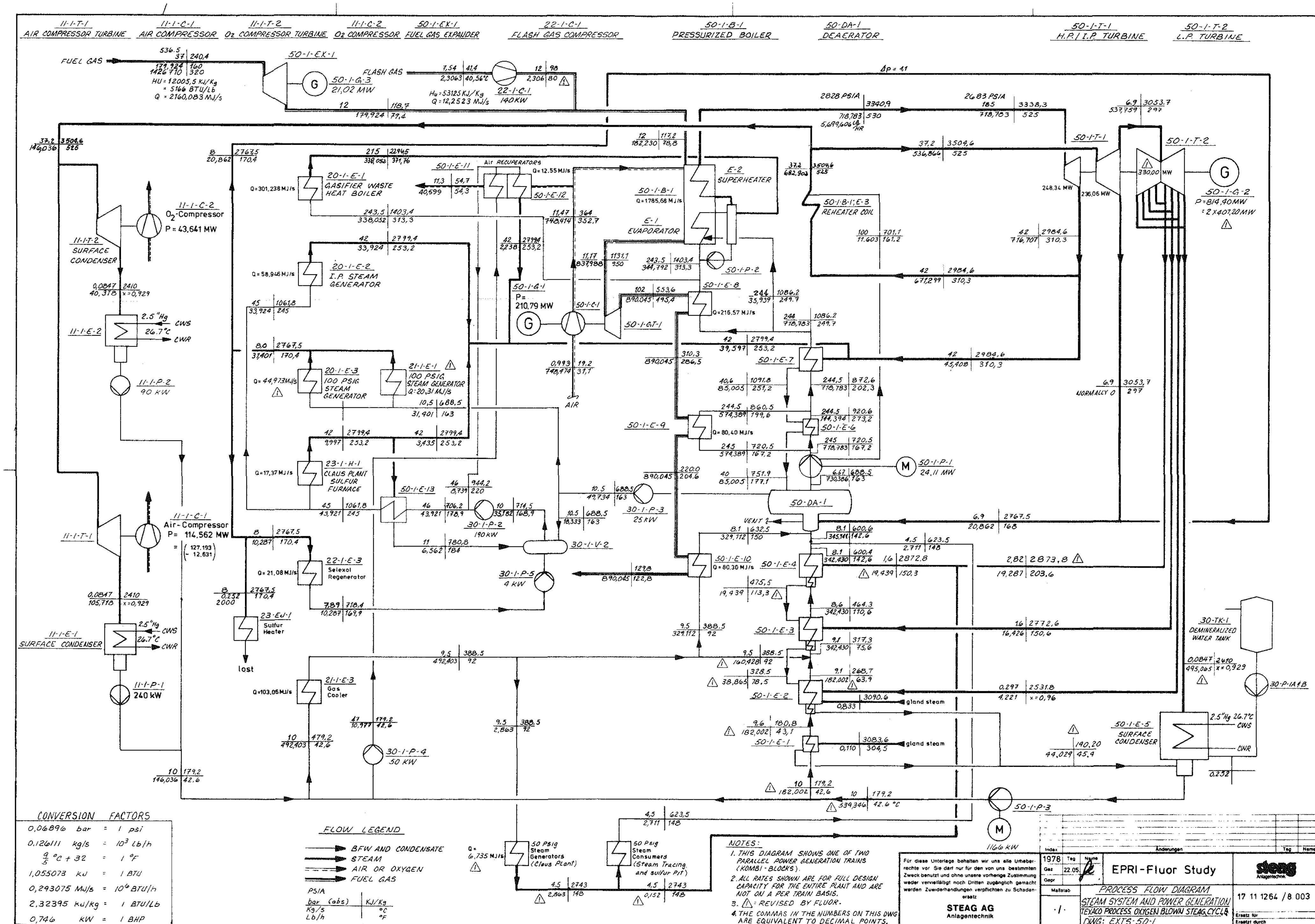
The following is a brief description of the process flow of the fuel gas received from the gasification area. The steam system is described separately in connection with Process Flow Diagram EXTS-30-1. Fuel gas enters the power block at 522 psig and 320°F and is expanded to the operating pressure of the pressurized boiler (160 psig) through fuel gas expander 50-1-EX-1. The fuel gas then flows to the pressurized boiler at 160 psig and 174°F for combustion. The flow of fuel to the pressurized boiler 50-1-B-1 is augmented by flash gas received from Unit 22.

Both the fuel gas and flash gas streams are combusted in the pressurized boiler 50-1-B-1 to produce superheated high-pressure steam. Flue gases exhaust from the pressurized boiler at 1742°F and 148 psig and enter gas turbine 50-1-GT-1 for expansion to atmospheric pressure. The hot turbine exit gases then flow through economizers 50-1-E-8, 50-1-E-9, and 50-1-E-10 to preheat boiler feedwater flowing to the pressurized boiler 50-1-B-1, and finally exhaust to the atmosphere through a tall stack.

Gas turbine 50-1-GT-1 drives an integrated air compressor 50-1-C-1, and also electric generator 50-1-G-1. Air from 50-1-C-1 flows to the pressurized boiler for combustion of fuel gas. A portion of this air flows to the oxidant supply system in Unit 11 through air recuperators 50-1-E-11 and 50-1-E-12.

Equipment Notes

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


11-1-T-1 AIR COMPRESSOR TURBINE
11-1-C-1 AIR COMPRESSOR
11-1-T-2 O2 COMPRESSOR TURBINE
11-1-C-2 O2 COMPRESSOR
50-1-EX-1 FUEL GAS EXPANDER
22-1-C-1 FLASH GAS COMPRESSOR
50-1-B-1 PRESSURIZED BOILER
50-DA-1 DEAERATOR
50-1-T-1 H.P./I.P. TURBINE
50-1-T-2 L.P. TURBINE
FUEL GAS
FLASH GAS
50-1-G-3 21,02 MW
22-1-C-1 140KW
20-1-E-1 GASIFIER WASTE HEAT BOILER
50-1-E-11
Air RECUPERATORS
50-1-E-12
E-2 SUPERHEATER
E-1 EVAPORATOR
50-1-B-1:E-3 REHEATER COIL
50-1-G-2 P=814,40MW =2x407,20MW
11-1-C-2 O2-Compressor P = 43,641 MW
11-1-T-2 SURFACE CONDENSER
11-1-E-2
11-1-P-2 90 KW
20-1-E-2 I.P. STEAM GENERATOR
50-1-G-1 P= 210,79 MW
50-1-C-1
50-1-GT-1
50-1-P-2
50-1-E-8
20-1-E-3 100 PSIG STEAM GENERATOR
21-1-E-1 100 PSIG STEAM GENERATOR
AIR
50-1-E-7
50-1-E-9
50-1-E-6
50-1-P-1 24,11 MW
23-1-H-1 CLAUS PLANT SULFUR FURNACE
50-1-E-13
30-1-P-2 190 KW
30-1-P-3 25 KW
30-1-V-2
50-DA-1
VENT
11-1-T-1
11-1-C-1 Air-Compressor P= 114,562 MW
22-1-E-3 Selexol Regenerator
30-1-P-5 4 KW
50-1-E-10
50-1-E-4
23-EJ-1 Sulfur Heater
lost
11-1-E-1 SURFACE CONDENSER
11-1-P-1 240 KW
21-1-E-3 Gas Cooler
30-1-P-4 50 KW
50-1-E-3
50-1-E-2
50-1-E-1
gland steam
30-TK-1 DEMINERALIZED WATER TANK
30-P-1A+B
50-1-E-5 SURFACE CONDENSER
50-1-P-3
1166 KW
CONVERSION FACTORS
0,06896 bar = 1 psi
0.126111 kg/s = 10³ Lb/h
9/5 °C + 32 = 1 °F
1,055073 kJ = 1 BTU
0,293075 MJ/s = 10⁶ BTU/h
2,32395 kJ/kg = 1 BTU/Lb
0,746 kW = 1 BHP
FLOW LEGEND
BFW AND CONDENSATE
STEAM
AIR OR OXYGEN
FUEL GAS
PSIA
bar (abs) | KJ/Kg
Kg/s | °C
Lb/h | °F
50 PSIG Steam Generators (Claus Plant)
50 PSIG Steam Consumers (Steam Tracing and sulfur PIT)
NOTES:
1. THIS DIAGRAM SHOWS ONE OF TWO PARALLEL POWER GENERATION TRAINS (KOMBI-BLOCKS).
2. ALL RATES SHOWN ARE FOR FULL DESIGN CAPACITY FOR THE ENTIRE PLANT AND ARE NOT ON A PER TRAIN BASIS.
3. REVISED BY FLUOR.
4. THE COMMAS IN THE NUMBERS ON THIS DWG ARE EQUIVALENT TO DECIMAL POINTS.
STEAG AG Anlagentechnik
1978
EPRI-Fluor Study
steag
PROCESS FLOW DIAGRAM
STEAM SYSTEM AND POWER GENERATION
TEXACO PROCESS OXYGEN BLOWN STEAG CYCLE
DWG: EXTS-50-1
17 11 1264 /8 003
FLUOR REV. 01 11-5-79

PROCESS DISCUSSION - CASE EXTS

The table below summarizes pertinent heat and material balance results.

Table 5-2

SUMMARY OF SYSTEM PERFORMANCE - CASE EXTS

GASIFICATION AND GAS CLEANING SYSTEM	
Coal Feed Rate, lbs/hr (m.f.)	798,333
Oxygen*/Coal Ratio, lbs/lb m.f.	0.858
Oxidant Temperature, °F	300
Steam/Coal Ratio, lbs/lb m.f.	0
Slurry Water/Coal Ratio, lbs/lb m.f.	0.503
Gasification Section Average Pressure, psig	600
Crude Gas Temperature, °F	2,300-2,600
Crude Gas HHV (dry basis), Btu/SCF**	281.1
Temperature of Fuel Gas to Power System, °F	320
POWER SYSTEM	
Gas Turbine Inlet Temperature, °F (at pressurized boiler exit)	1,742
Turbine Air Compressor Pressure Ratio	12:1
Gas Turbine Exhaust Temperature, °F	924
Steam Cycle Conditions, psig/°F/°F (at turbine inlet)	2669/977/977
Condenser Pressure, Inches Hg abs	2.5
Stack Temperature, °F	253
Gas Turbine Power#, MW	211
Steam Turbine Power#, MW	814
Fuel Gas Expansion Turbine Power #, MW	21
Power Consumed, MW	73
Net System Power, MW	973
OVERALL SYSTEM	
Process and Deaerator Makeup Water, gpm/1000 MW, Net	670
Cooling Tower Makeup Water, gpm/1000 MW, Net	11,312
Cooling Water Circulation Rate, gpm/MW	450
Cooling Tower Heat Rejection, % of Coal HHV	50.1
Air Cooler Heat Rejection, % of Coal HHV	3.1
Net Heat Rate, Btu/kWh	10,479
Overall System Efficiency (Coal → Power), % of Coal HHV	32.6

*Dry basis, 100 percent O_2
**Excluding the HHV of H_2S, COS and NH_3
#At generator terminals

GASIFIER MATERIAL BALANCE

The oxygen-blown Texaco gasifier material balance for full capacity operation is given in Table 5-3 for the oxygen-blown Texaco gasifier.

Most of the data presented in the above table was received from Texaco Development Corporation for an earlier study published as EPRI Report AF-642. That study projected economics and operating data for the 1980-1985 time frame. For the particular coal used, Texaco indicated that slurry concentrations in the range of 60 percent solids to possibly 70 percent solids could be achieved in the 1980-1985 time frame. For study purposes, EPRI selected a slurry concentration of 66.5 percent 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 possess different slurrying characteristics. The material balance, including oxygen consumption, is based on a Texaco extrapolation of the state of the art to a period three to five years hence.

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 salts 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.

The figures given in the Tables and Flow Sheets for ammonia should be regarded as tentative estimates only.

The gasifier temperatures are believed to be high enough to destroy all hydrocarbons 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)

Table 5-3

MATERIAL BALANCE - CASE EXTS

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		

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. This process is used in these cases because it accomplishes this objective and exhibits favorable economics when compared with other similar processes.

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 only.

PROCESS ENERGY BALANCES

Table 5-4 presents an overall process energy balance at 100 percent capacity operation. The boundary for each balance encompasses the entire plant, exclusive of the cooling tower heat balance. 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. The energy balance closes to less than one-half of one percent. The discrepancy results from approximations used for some process units and for calculating some heat loads.

Data from Table 5-4 is shown in MM Btu/hr and as percent of coal higher heating value in Table 5-5. Coal charged at 10,000 ton/day is equivalent to 10,196 MM Btu/hr HHV. This feed produces 3320 MM Btu/hr power equivalent or 32.6 percent of the coal HHV as net electric power. The heat rate based on net power produced is 10,479 Btu/kWh. Heat rejected at all steam turbine surface condensers is 4883 MM Btu/hr or 47.9 percent of the coal HHV. Heat rejected with the power block stack gases is 930 MM Btu/hr or 9.1 percent of the coal HHV.

Table 5-4

ENERGY BALANCE - CASE EXTS

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 (Gasification)		20	48			68
Air Compressor Air (Power Block)		40	97			137
Demineralized and Raw Water		3				3
Auxiliary Power Inputs					249	249
TOTAL	10,196	68	145	0	249	10,658
HEAT OUT						
Ash Cake		6				6
Gasifier Heat Losses				26		26
Gas Cooling		19	6			25
Sulfur Product	105	1				106
Oxidant Compressor Inter/After Cooling		498	33			531
Oxidant Compressor Surface Condensers			1,112			1,112
Gas Turbines					792	792
Sulfur Plant Effluent Gas		2	19			21
Steam Turbines					2,777	2,777
Power Block Losses*				11	74	85
Steam Turbine Condenser			3,771			3,771
Power Block Stack Loss		334	596			930
Steam Heat Losses			25			25
Motor Losses (Air Cooler Fans, etc.)					212	212
Selexol Overhead Condenser			24			24
Selexol Solvent Cooler		54				54
Process Water Cooling		95				95
Air Separation Plant Waste Gas		31	20			51
Waste Water Effluent		10				10
TOTAL	105	1,050	5,606	37	3,855	10,653

$$\frac{\text{Input - Output}}{\text{Input}} = 0.04\%$$

*Includes mechanical and electrical losses

Table 5-5

ENERGY BALANCE AS PERCENT COAL HHV - CASE EXTS

		MM Btu/hr	Percent
IN			
	Coal HHV	10,196	100.0
OUT			
	Net Power	3,320	32.56
	Sulfur Product, HHV and Sensible	106	1.04
	Selexol Sensible and Latent	78	0.77
	Oxidant Inter/After Cooling	531	5.21
	Ash Cake	6	0.06
	Power Block Stack Gases	930	9.12
	Rejected at Surface Condensers	4,883	47.89
	Other Sensible Losses, Net	89	0.87
	Other Latent Losses, Net	(75)	(0.74)
	Gasifier Heat Losses	26	0.26
	Motor Losses	212	2.08
	Power Block Losses	85	0.83
		10,191	99.95

ECONOMICS - CASE EXTS

Table 5-6 summarizes the economics of Case EXTS.

Table 5-6

SUMMARY OF ECONOMIC RESULTS - CASE EXTS

PRODUCTION AT DESIGN CAPACITY	
Net Power, MW*	973
Overall Plant Heat Rate, Btu/kWh	10,479
TOTAL CAPITAL REQUIREMENTS**	
Total Capital Requirement, $1,000	949,098
Total Capital Requirement, $/kW	975.43
AVERAGE COSTS OF SERVICES**	
First Year Cost, $1,000/year	261,722
First Year Cost, mills/kWh	43.87
Thirty Year Levelized Cost, $1,000/year	345,121
Thirty Year Levelized Cost, mills/kWh	57.84

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

TOTAL PLANT INVESTMENT

Table 5-7 gives a detailed breakdown of the Total Plant Investment required for Case EXTS. The accuracy of the plant investment estimate is judged to be ±25 percent. Since other capital charges and working capital are keyed to elements of plant investment, this accuracy is reflected in other capital figures as well. However, due to the similar nature of the cases in this evaluation, the estimates

Table 5-7

TOTAL PLANT INVESTMENT - CASE EXTS

Plant Section	Cost Breakdown Without Contingencies: Direct Field Material#	Direct Field Labor##	Engr. & Support Costs§	Total Sales Tax	Total Cost $1,000*	Cost $/kW**	Per-cent	Contingencies: Process $1,000*	Contingencies: Project $1,000*	Total Plant Investment $1,000*	Total Plant Investment $/kW**
Coal Handling	10,499	4,435	7,127	530	22,591	23.22	3.40	-	3,389	25,980	26.70
Oxidant Feed	62,257	23,383	30,614	1,939	118,193	121.47	17.78	-	17,729	135,922	139.70
Gasification and Ash Handling	14,288	3,456	6,622	721	25,087	25.78	3.77	3,460	3,763	32,310	33.21
Gas Cooling	61,480	29,010	45,885	3,187	139,562	143.44	21.00	32,524	20,934	193,020	198.38
Acid Gas Removal and Sulfur Recovery	15,025	5,230	8,940	768	29,963	30.79	4.51	668	4,494	35,125	36.09
Steam, Condensate and BFW	631	265	420	31	1,347	1.38	0.20	-	202	1,549	1.59
Combined Cycle	141,005	43,171	77,172	7,266	268,614	276.07	40.41	13,431	40,292	322,337	331.28
General Facilities	30,471	12,110	15,833	957	59,371	61.02	8.93	-	8,906	68,277	70.17
Subtotal	335,656	121,060	192,613	15,399	664,728	683.17	100.00	50,083	99,709	814,520	837.12

TOTAL PLANT INVESTMENT SUMMARY

	$1,000*	$/kW**
Process Plant Investment and General Facilities	664,728	683.17
Process Contingency	50,083	51.47
Project Contingency	99,709	102.48
Total Plant Investment	814,520	837.12

*Mid-1976 dollars

**Based on 100 percent plant design power output (973 MW)

#All materials and equipment that become a part of the plant facility

##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)

§Includes: a) Indirect field costs including all labor, supervision and expenses required to support field construction; b) Home office costs including all salaries and expenses required for engineering design and procurement; and c) Contractor's fee

for all cases should reflect about the same accuracy. When these estimates are used comparatively, the effect of individual accuracies should be minimal.

Two contingencies are included in the Total Plant Investment shown for each plant section. First is a 15 percent project contingency which is included 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 innovative technology in an effort to quantify the uncertainty in the design, performance, and cost of the commercial scale equipment. This covers the additional expenditures required to solve any problems associated with innovative technology, since historically such technology requires more cost than initially estimated. Although all plant technology is judged to be probably commercially obtainable, not all of it is fully proven by commercial operation. Accordingly, a process contingency is applied and included in the estimated plant investment for sections containing innovative technology.

TOTAL CAPITAL REQUIREMENT

Table 5-8 gives a breakdown of the Total Capital Requirement to place the plant into initial operation at 100 percent capacity. Starting with the total plant investment, capital charges are added for allowance for funds during construction, initial catalyst and chemicals charge, prepaid royalties, preproduction costs, the start-up inventory capital, and the land required for the plant. Specific items included in each of these capital charges are described under the economic criteria.

COST OF SERVICES

Table 5-9 gives a cost of services breakdown for Case EXTS. The costs are busbar power costs based on plant operation at a 70 percent capacity factor with a $1/MM Btu coal cost. Table 5-9 shows both first-year and 30-year levelized power costs. Since the power costs vary with the plant capacity factor, this additional relationship is shown in Figure 5-9 on both a first-year and 30-year levelized basis. The plant capacity factor presented in Figure 5-1 is the percent plant time on-line at baseload (100 percent) plant power output of 865 MW. This capacity factor is not equivalent to plant operation in a turndown mode producing less than baseload power output, as the design assumes a baseload plant.

Table 5-8

TOTAL CAPITAL REQUIREMENT - CASE EXTS

	$1000*	$/kW**
TOTAL PLANT INVESTMENT	814,520	837.12
CAPITAL CHARGES		
Prepaid Royalties	3,024	3.11
Preproduction Costs	20,625	21.20
Inventory Capital	7,481	7.69
Initial Catalyst and Chemicals Charge	714	0.73
Allowance for Funds During Construction	101,734	104.55
Land	1,000	1.03
Total Capital Charges	134,578	138.31
TOTAL CAPITAL REQUIREMENT	949,098	975.43

*Mid-1976 dollars

**Based on 100 percent plant design power output

The largest single component of the power cost is contained in the levelized fixed charges. These charges represent repayment from operating revenue of the financing originally used to supply the total capital required for construction and startup of the plant. These levelized fixed charges amount to 65 percent of the first year power costs at a 70 percent capacity factor. The charges shown are fixed annual expenses which are independent of the plant capacity factor.

Coal constitutes the major operating charge. This charge amounts to 24 percent of the first year power costs at a 70 percent capacity factor.

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

- "A" Operators 5
- "B" Operators 17
- Foremen 2
- Lab and Instrument Technicians 4

Table 5-9

BUSBAR POWER COST AT 70 PERCENT CAPACITY FACTOR - CASE EXTS

	O_2- Blown Texaco EXTS	
NET PRODUCTION*		
Net Power, MW	973	
By-product Ammonia, ST/D	-0-	
By-product Sulfur, ST/D	301	
TOTAL CAPITAL REQUIREMENT, $1,000	949,098	
	First-Year Cost	30-Year Levelized Cost
FIXED OPERATING COST, $1,000/YEAR		
Operating Labor	3,066	5,782
Maintenance Labor	7,599	14,331
Maintenance Materials	11,398	21,497
Administrative & Support Labor	3,199	6,034
Total Fixed O&M Costs	25,262	47,644
VARIABLE OPERATING COSTS (EXCLUDING COAL), $1,000/YEAR		
Raw Water	1,801	3,397
Catalysts & Chemicals	319	602
Ash Disposal	981	1,850
Total Variable O&M Costs	3,101	5,849
COAL COST, $1,000/YEAR	62,521	120,790
BY-PRODUCT CREDITS, $1,000/YEAR		
By-product Ammonia	-0-	-0-
By-product Sulfur	-0-	-0-
Total Byproduct Credits	-0-	-0-
TOTAL OPERATING COSTS, $1,000/YEAR	90,884	174,283
LEVELIZED FIXED CHARGES, $1,000/YEAR	170,838	170,838
TOTAL COST OF ELECTRICITY**		
$1,000/year	261,722	345,121
mills/kWh	43.87	57.84

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

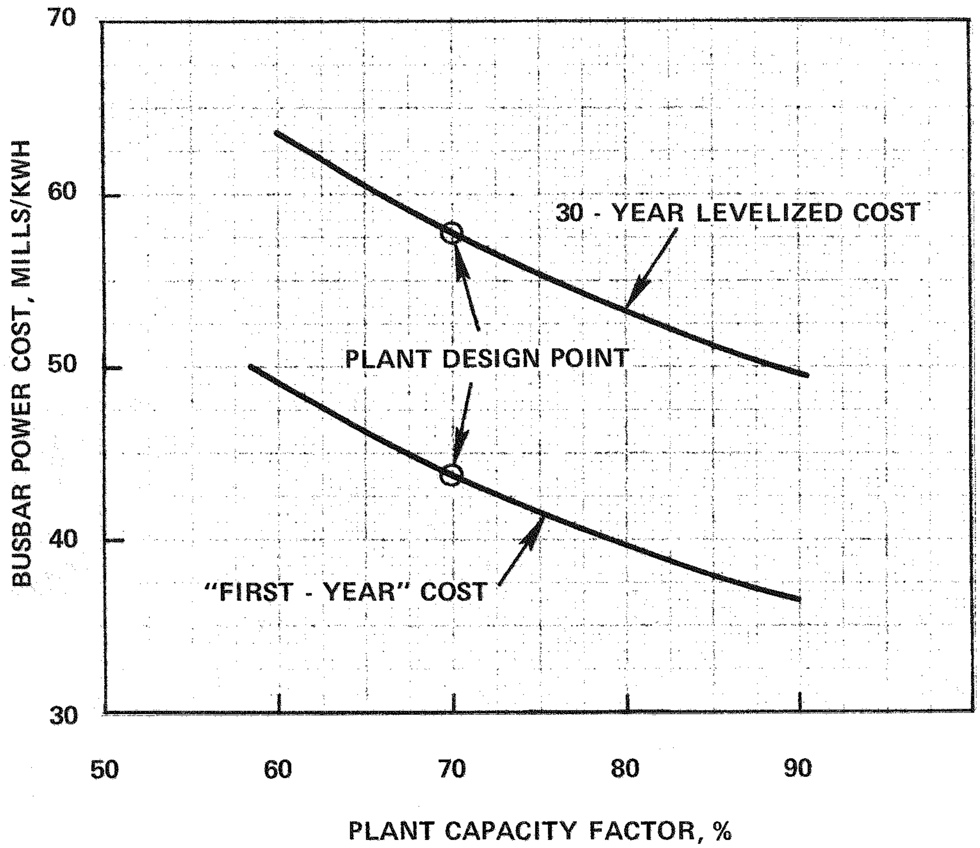


Figure 5-1. Power Cost Sensitivity Curve for Case EXTS

Maintenance labor and materials costs are calculated for each section as a percentage of the total plant investment for that section. These percentages are listed in the Economic Criteria section.

Catalyst and chemical costs are for chemicals consumed in the demineralizer, cooling tower and boiler feedwater treating systems, plus costs associated with making up solution losses in the acid gas removal, and tail gas treating units, as well as replacement of catalyst in the sulfur recovery unit.

Section 6

PLANT DESCRIPTION - CASE EATS

GENERAL

A grass roots plant for electric power generation in a STEAG pressurized boiler combined cycle integrated with single-stage, entrained-bed, air-blown gasifiers of the Texaco type is shown schematically on the Block Flow Diagram EATS-1-1. This plant consumes 10,000 ST/day of Illinois No. 6 coal.

The Case EATS plant differs from the Texaco-based plant, presented as Case EATC (Slurry Feed) in EPRI Report AF-753, primarily in that the STEAG pressurized boiler combined cycle is used for generation of electric power. Block Flow Diagram EATS-1-1 shows a power generation section consisting of two identical pressurized boiler combined-cycle units in parallel. This process arrangement is different from that of Case EATC which used six parallel gas turbines and heat recovery steam generators plus one steam turbogenerator. The remaining process units in Case EATS are relatively similar to their counterparts in Case EATC.

Table 6-1 summarizes major equipment sections in the plant and shows the numbers of operating and spare trains. The train count listed is the same as in Case EATC except as noted.

The following sections of this report contain descriptions of the major process units in the Case EATS plant. Emphasis is placed on describing the differences in process design between the Case EATS units and the units shown for Case EATC in EPRI Report AF-753.

Table 6-1

TRAINS OF EQUIPMENT IN MAIN PROCESSING UNITS - CASE EATS

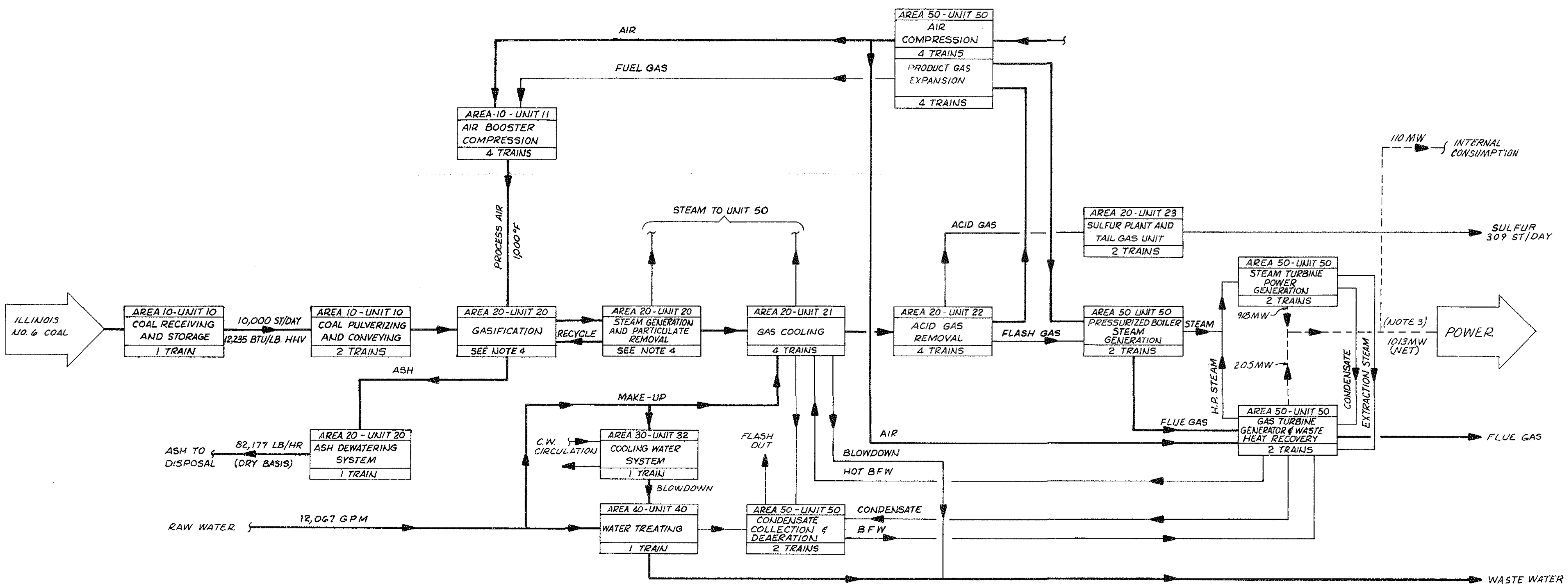
Unit No.	Name	Case EATS Operating	Case EATS Spare
10	Coal Handling	1	0
11	Oxidant Feed	4**	0
20	Wet Coal Grinding	2	0
20	Slurry Preparation	1	0
20	Gasification	*	0
20	Ash Handling	1	0
20	Particulate Scrubbing	*	0
21	Gas Cooling	4	0
22	Acid Gas Removal	4	0
23	Sulfur Recovery	2	1
23	Tail Gas Treating	2	1
30	Steam, BFW and Condensate System		
	o Condensate Collection and Deaeration	2#	0
	o Water Treating	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0
50	Power Generation	2##	0

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

**The oxidant feed system was increased from the three trains in Case EATC to four trains

#The train count increased from the one train in Case EATC to the two listed to match the two STEAG power generation trains

##These two STEAG power generation trains replace the six gas-turbine-heat-recovery-steam-generators of Case EATC



NOTES:

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

DATE	REVISION DESCRIPTION	DRAWN CHECK	APP

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FLUOR

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

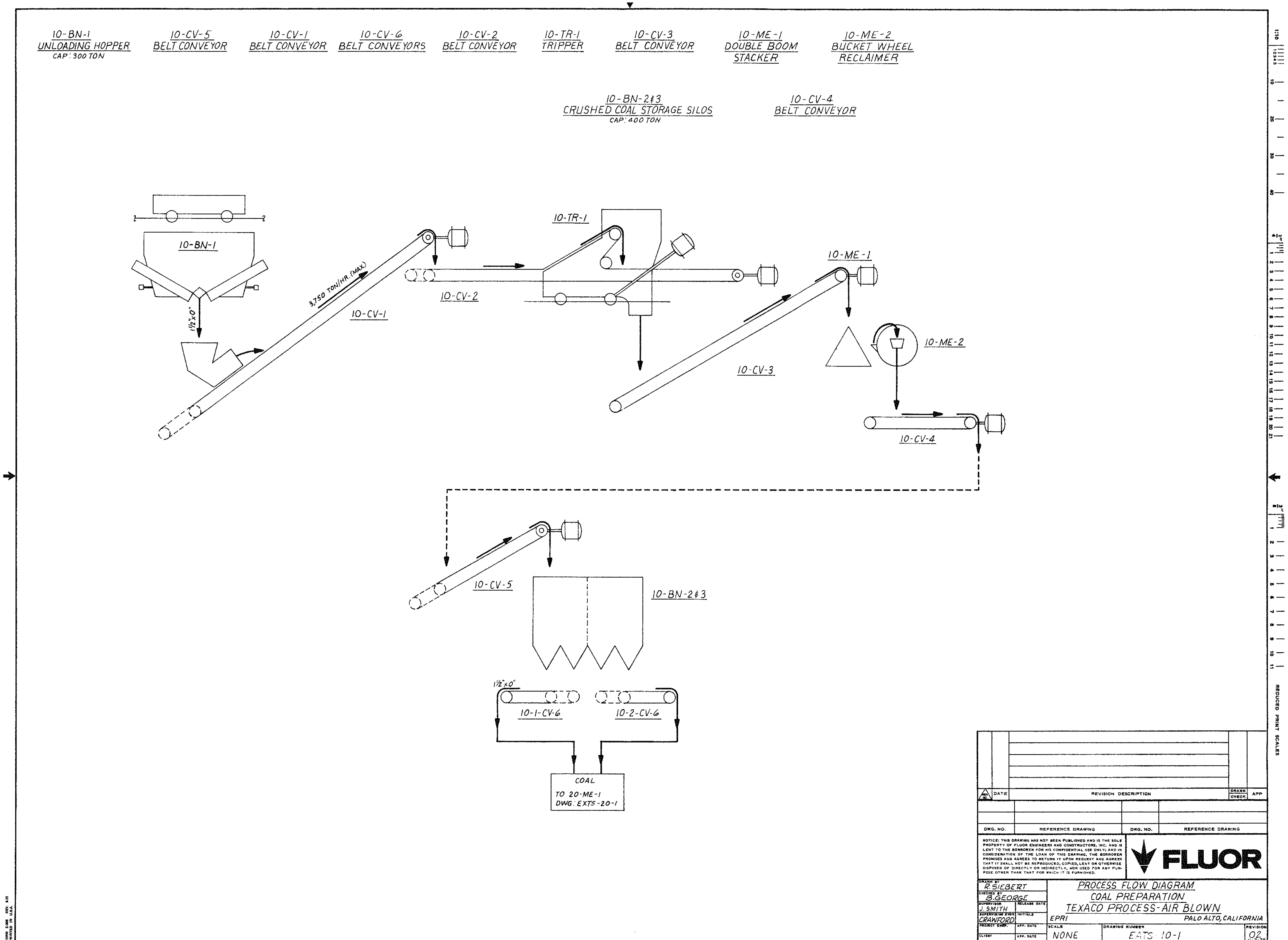
SCALE NONE — DRAWING NUMBER ATS-1-1 — REVISION 0 11-7-79

COAL PREPARATION

Process Flow Diagram EATS-10-1 depicts the process arrangement of equipment in this section for Case EATS. Coal preparation section process details are essentially the same as those described for Case EATC in EPRI Report AF-753.

Equipment Notes

All the equipment in this section is commercially available.



10-BN-1
UNLOADING HOPPER
CAP: 300 TON
10-CV-5
BELT CONVEYOR
10-CV-1
BELT CONVEYOR
10-CV-6
BELT CONVEYORS
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-BN-2&3
CRUSHED COAL STORAGE SILOS
CAP: 400 TON
10-CV-4
BELT CONVEYOR
10-BN-1
1½"x0"
3,750 TON/HR. (MAX)
10-CV-1
10-CV-2
10-TR-1
10-ME-1
10-CV-3
10-ME-2
10-CV-4
10-CV-5
10-BN-2&3
1½"x0"
10-1-CV-6
10-2-CV-6
COAL
TO 20-ME-1
DWG: EXTS-20-1
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PROCESS FLOW DIAGRAM
COAL PREPARATION
TEXACO PROCESS-AIR BLOWN
EPRI
PALO ALTO, CALIFORNIA
SCALE
NONE
DRAWING NUMBER
EATS 10-1
REVISION
02
11-7-79
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OXIDANT FEED

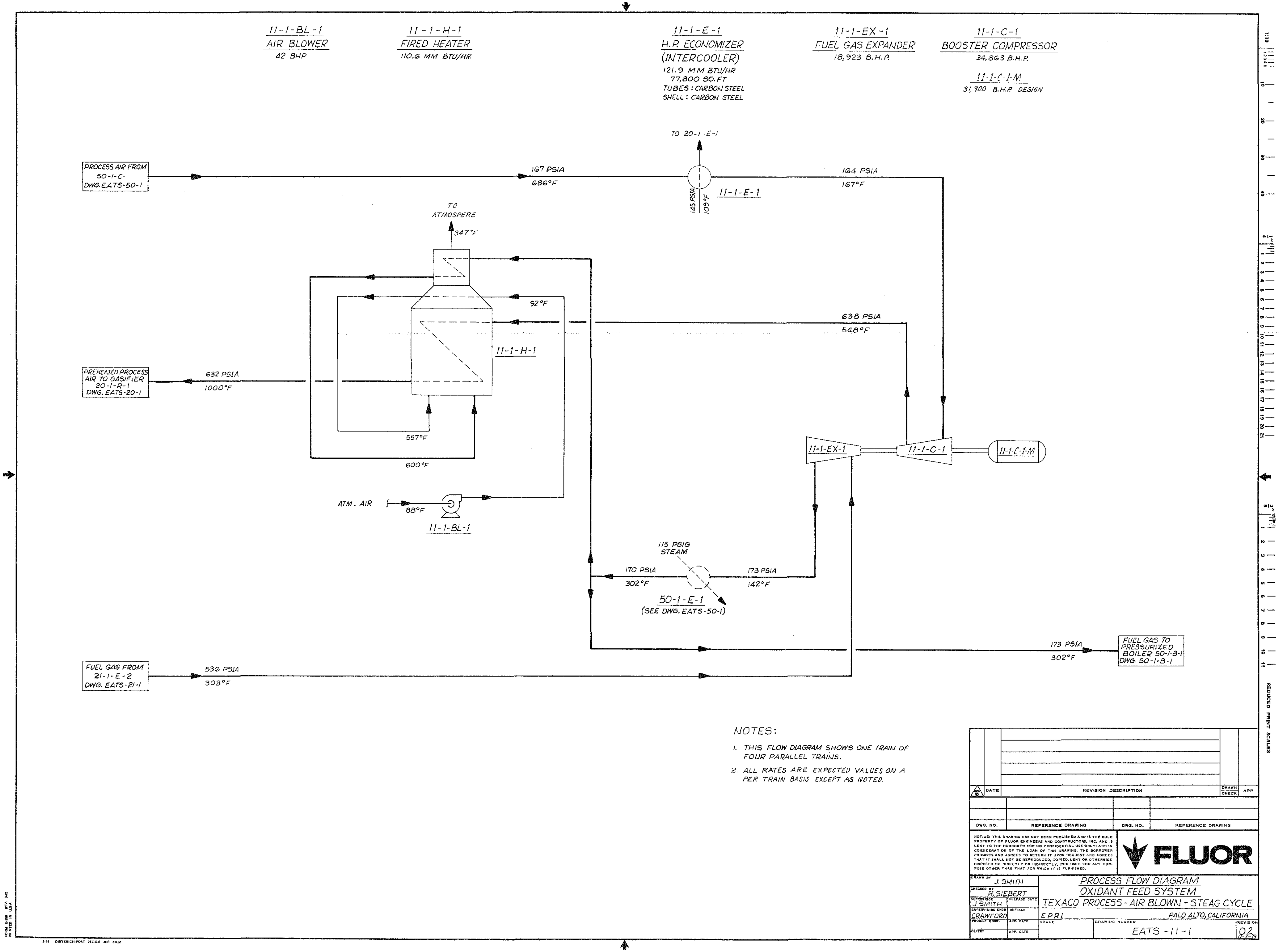
Process Flow Diagram EATS-11-1 shows the oxidant feed system for Case EATS. There are four parallel operating trains. Each train has one booster air compressor, the associated heat exchangers, and a fired heater to preheat the air to 1000°F before it flows to the gasifiers. No spare train is provided in this section.

Process air in Case EATS is extracted from the gas turbines at 686° and 167 psia. The air is cooled to 167° and compressed to 638 psia in the booster compressor, 11-1-C-1. The air is then preheated to 1000°F in the fired heater 11-1-H-1 and flows to the gasifiers.

The 34,863 hp required by each air booster compressor is supplied by a fuel gas expander, 11-1-EX-1, supplemented by a helper motor. The fuel gas is expanded from 536 psia, 303°F to approximately 173 psia, 142°F, and is then reheated to 302°F in 50-1-E-1 with 115 psig steam. A small part of the fuel gas stream is then used as fuel for fired heater 11-1-H-1. The remainder then flows to the power block as fuel for the supercharged boilers.

Equipment Notes

The compressors are within current technology, although the high discharge temperature may require prototype design features. The remainder of the equipment is commercially available.



11-1-BL-1
AIR BLOWER
42 BHP
11-1-H-1
FIRED HEATER
110.6 MM BTU/HR
11-1-E-1
H.P. ECONOMIZER
(INTERCOOLER)
121.9 MM BTU/HR
77,800 SQ. FT
TUBES: CARBON STEEL
SHELL: CARBON STEEL
11-1-EX-1
FUEL GAS EXPANDER
18,923 B.H.P.
11-1-C-1
BOOSTER COMPRESSOR
34,863 B.H.P.
11-1-C-1-M
31,900 B.H.P. DESIGN
TO 20-1-E-1
PROCESS AIR FROM 50-1-C- DWG. EATS-50-1
167 PSIA
686°F
145 PSIA
109°F
11-1-E-1
164 PSIA
167°F
TO ATMOSPERE
347°F
92°F
638 PSIA
548°F
11-1-H-1
PREHEATED PROCESS AIR TO GASIFIER 20-1-R-1 DWG. EATS-20-1
632 PSIA
1000°F
557°F
600°F
11-1-EX-1
11-1-C-1
11-1-C-1-M
ATM. AIR
88°F
11-1-BL-1
115 PSIG STEAM
170 PSIA
302°F
173 PSIA
142°F
50-1-E-1
(SEE DWG. EATS-50-1)
173 PSIA
302°F
FUEL GAS TO PRESSURIZED BOILER 50-1-B-1 DWG. 50-1-B-1
FUEL GAS FROM 21-1-E-2 DWG. EATS-21-1
536 PSIA
303°F
NOTES:
1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF FOUR PARALLEL TRAINS.
2. ALL RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS EXCEPT AS NOTED.
REV NO
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PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
TEXACO PROCESS - AIR BLOWN - STEAG CYCLE
EPRI
PALO ALTO, CALIFORNIA
SCALE
DRAWING NUMBER
EATS-11-1
REVISION
02
11-7-79
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GASIFICATION AND ASH HANDLING

Process Flow Diagram EATS-20-1 shows the gasification step for Case EATS. The details of gasification and ash handling are similar to Case EATC described in EPRI Report AF-753. The steam generation system, however, is different from Case EATC in that higher steam pressure levels are used.

Raw hot gas from the gasifier is first cooled in a gas cooling unit, 20-1-ME-3. The Gas Cooling Unit 20-1-ME-1 generates high-pressure (HP) steam at 2813 psig, saturated intermediate-pressure (IP) steam at 595 psig and saturated medium-pressure (MP) steam at 15 psig. The HP steam and IP steam both flow to steam turbines in the STEAG power generation unit for production of electric power. The MP steam is used both as stripping steam in the deaerator and for various process heating services. Preheated boiler feedwater streams are supplied from the STEAG power generation unit at temperatures approaching saturation for each steam pressure level.

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: EATS-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 shown in Flow Diagram EATS-21-1.

Equipment Notes

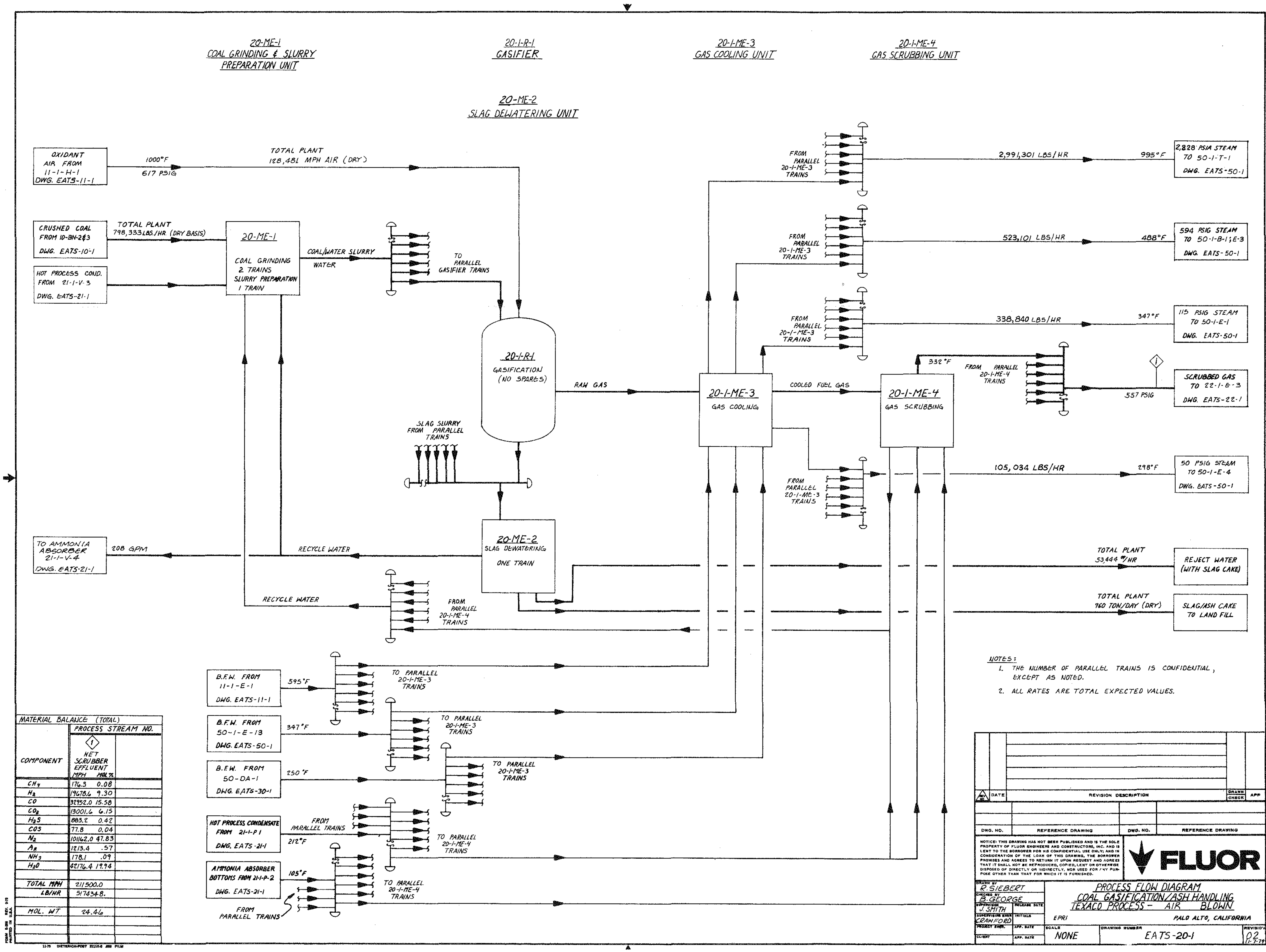
Since the 2813 psig superheated steam generation pressure is a feature of the supercharged boiler used in the STEAG cycle, a similarly high waste heat boiler steam generation and superheating pressure in the gasification area is required for integration with the steam system of the STEAG cycle. Therefore, development of a 2813 psig steam pressure waste heat boiler/steam superheater would be necessary for integration of the Texaco gasification process with the STEAG cycle. This would represent a significant extension of the present state of the art for waste heat boilers.

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 oxygen-blown coal gasifier is currently undergoing test runs in Germany. It should be noted that the air-blown Texaco process is at an earlier stage of development relative to the O_2-blown case.

The slag dewatering unit is commercially proven.

The gas scrubbing unit equipment is commercially available.



MATERIAL BALANCE (TOTAL)

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

GAS COOLING

The process details for each gas cooling train, shown in Process Flow Diagram EATS-21-1, are similar to those described for Case EATC in EPRI Report AF-753. The rate of generation of 50 psig steam in 21-1-E-3A has been changed, and the flow rate of cold condensate through 21-1-E-3B has also been revised in order to accommodate the integration with the STEAG cycle.

Equipment Notes

All equipment is commercially available.

21-1-V-1
K.O. DRUM
152" I.D. x 10'-0" T-T
CARBON STEEL
DESIGN: 600 PSIG@380°F

21-1-E-2
FUEL GAS PREHEATER #1
64.8MM BTU/HR
16,500 SQ. FT.
SHELL: CARBON STEEL
TUBES: 304 SS.

21-1-V-2
K.O. DRUM
114" I.D. x 9'-6" T-T
CARBON STEEL
DESIGN: 600 PSIG@380°F

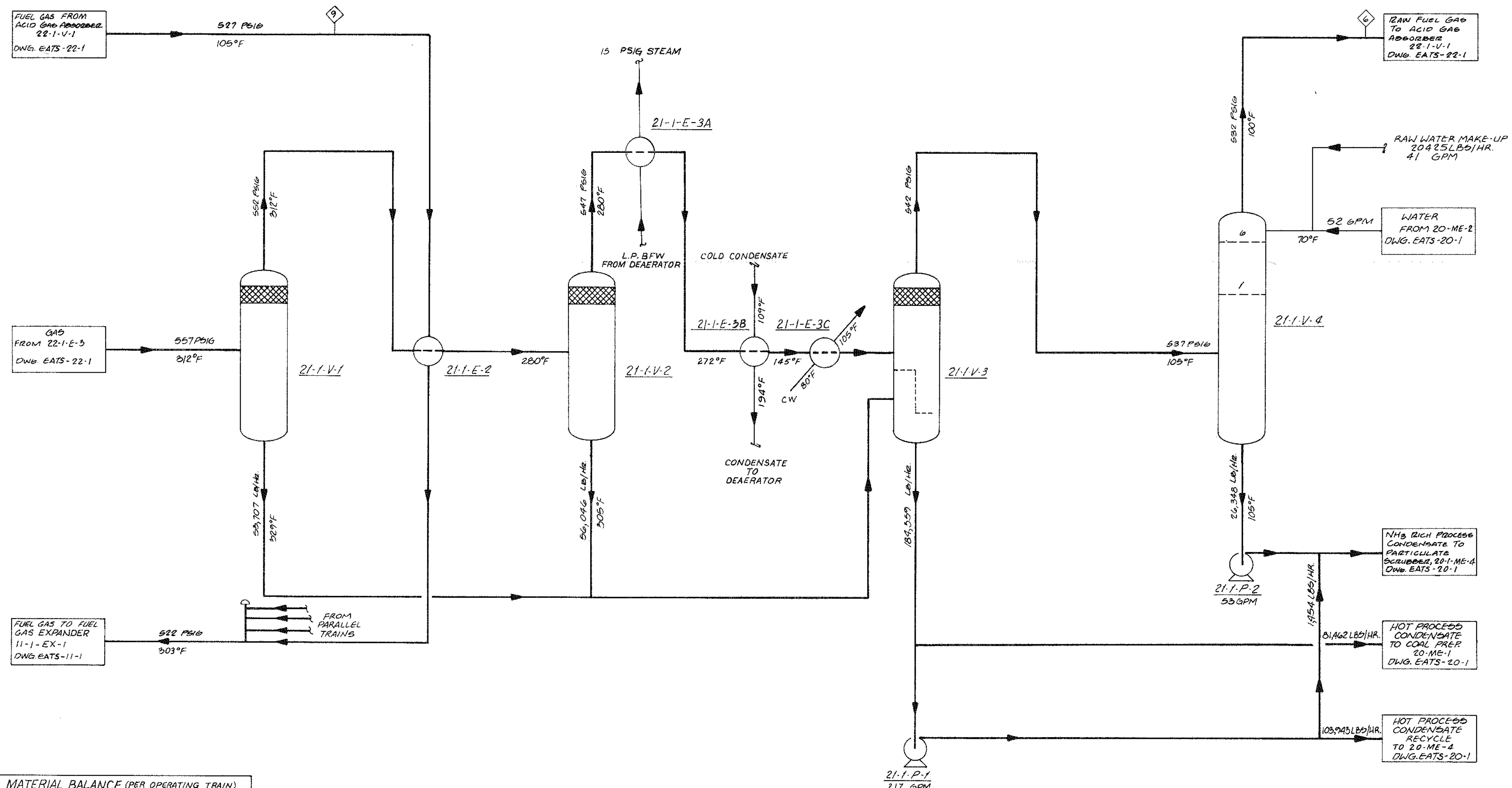
21-1-E-3A
L.P. STEAM GENERATOR
29.55 MM BTU/HR
10,347 SQ. FT.
SHELL: CARBON STL.
TUBES: 304 SS.

21-1-E-3B
COLD CONDENSATE HEATER
85.28 MMBTU/HR.
11,100 SQ. FT.
SHELL: CARBON STEEL
TUBES: 304 SS

21-1-V-3
K.O. DRUM
114" I.D. x 10'-6" T/T
CARBON STEEL
DESIGN: 600 PSIG @ 380°F

21-1-E-3C
TRIM COOLER
22.5 MMBTU/HR.
10764 SQ. FT.
SHELL: CARBON STEEL
TUBES: 304 S.S.

21-1-V-4
AMMONIA ABSORBER
120" I.D x 10'-0" T-T
CARBON STEEL
DESIGN: 600 PSIG@160°F



NOTES:

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

MATERIAL BALANCE (PER OPERATING TRAIN)

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

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

11-75 DIETERICH-POST 2111X-8 .003 FILM

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

SCALE NONE
DRAWING NUMBER EATS-21-1
REVISION 02 11-7-79

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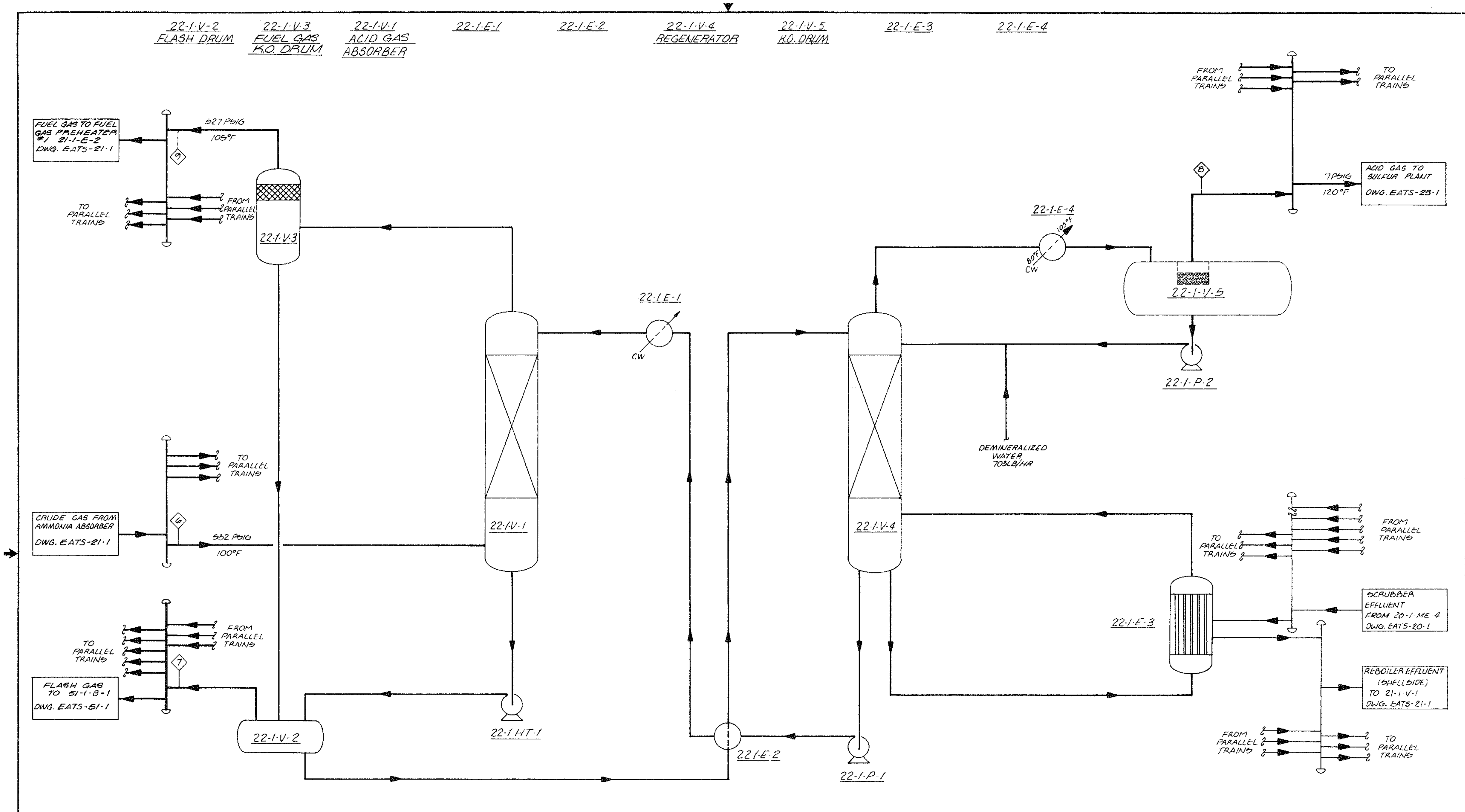
ACID GAS REMOVAL

Process Flow Diagram EATS-22-1 depicts one of the four parallel acid gas removal trains. This unit is similar to its counterpart designed for Case EATC and described in EPRI Report AF-753. 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.



COMPONENT	⟨6⟩ CRUDE GAS MPH	MOL%	⟨7⟩ FLASH GAS MPH	MOL%	⟨8⟩ ACID GAS MPH	MOL%	⟨9⟩ FUEL GAS MPH	MOL%
MATERIAL BALANCE — PROCESS STREAM NUMBER								
CH_4	44.1	0.10	0.3	.17	0.4	0.05	43.4	0.10
H_2	4,919.7	11.61	11.9	7.47	3.6	0.41	4,904.2	11.86
CO	8,238.0	19.45	34.1	21.43	22.2	2.51	8,181.7	19.79
CO_2	3,250.4	7.67	29.4	18.45	537.0	60.75	2,684.0	6.49
H_2S	220.8	0.52	1.9	1.20	210.7	23.83	8.2	0.02
COS	19.5	0.0[illegible]	.3	0.13	7.3	.82	11.9	0.03
N_2	25,290.5	59.71	80.0	50.24	34.0	3.84	25,176.5	60.88
AR	303.4	0.72	1.0	0.60	0.4	0.05	302.0	0.73
NH_3	0.4	0.00	0.0	0.00	.3	0.03	0.1	0.00
H_2O	71.2	0.17	.5	0.31	68.2	7.71	41.5	0.10
TOTAL MPH	42,358.0		159.5		884.1		41,355.5	
TOTAL LB/HR	1,115,017		4,650		34,095		1,076,987	
MOL WT	26.32		29.16		38.56		26.04	

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NOTES:

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

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

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

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

EPRI PALO ALTO, CALIFORNIA

SCALE NONE
DRAWING NUMBER EATS-22-1
REVISION 01 11-7-79

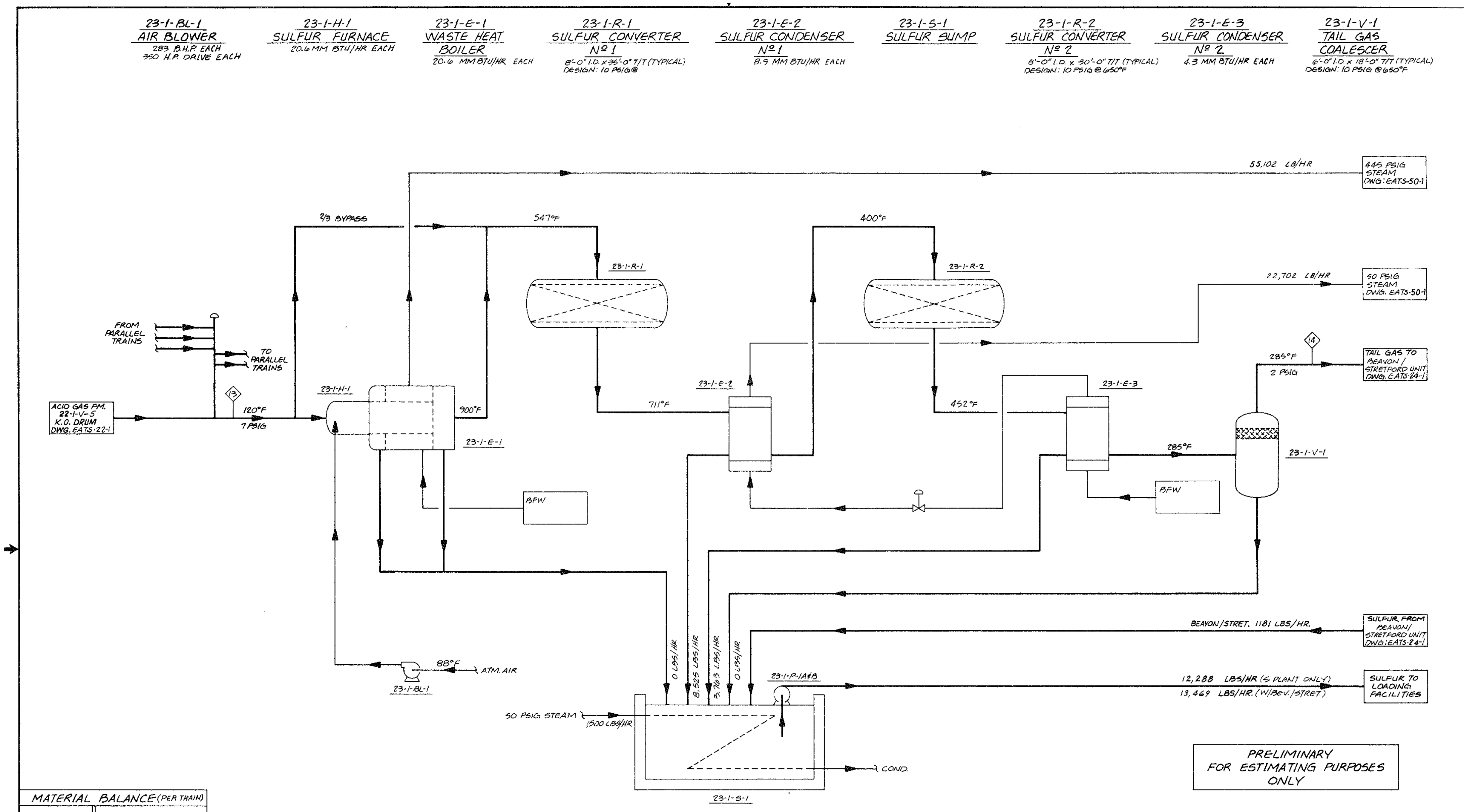
SULFUR RECOVERY AND TAIL GAS TREATING

Process Flow Diagram EATS-23-1 and EATS-24-1 depict the process arrangement for these units. Refer to EPRI Report AF-642 (January 1978, pages 74 through 87) for a detailed process description.

There are two 50 percent parallel operating sulfur recovery trains each followed by a tail gas treating unit. Sulfur recovery per train is 1545 short tons per day. There is a third (spare) train because of the important environmental requirements these units fulfill.

Equipment Notes

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



MATERIAL BALANCE (PER TRAIN)

STREAM Nº / COMPONENT	13 ACID GAS FM. 1½ SELEXOL UNIT MPH	MOL%	14 TAIL GAS TO T.G. TREATING UNIT MPH	MOL%
H_2	7.2	0.41	4.86	0.19
CO	44.3	2.53	29.56	1.16
CO_2	1074.0	61.31	1101.8	43.23
H_2S	405.1	23.12	31.0	1.22
COS	14.6	0.84	2.5	0.10
N_2	69.3	3.96	837.9	32.87
NH_3	0	0	0	0
CH_4	0.9	.05	0	0
H_2O VAPOR	136.4	7.79	540.3	21.19
SO_2	0	0	.86	.03
S_2	0.	0	0	0
S_6	0	0	0.05	0
S_8	0	0	0.23	.01
TOTAL VAPOR MPH	1751.8		2549.1	
H_2O LIQ.	0		0	
TOTAL LBS/HR	67623.8		83870.9	
MOL. WT. VAPOR	38.60		32.90	

FORM E-306 REV. 9-72 PRINTED IN U.S.A.

1. THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE-50% PARALLEL TRAINS. (TWO OPERATING AND ONE SPARE)
2. TWO SULFUR SUMPS WILL SERVICE THE THREE TRAINS.
3. ALL RATES ARE TYPICAL VALUES ON A TOTAL PLANT BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.
4. CONFIGURATION AND SIZE OF EQUIPMENT CORRESPOND TO A DESIGN SULFUR RECOVERY RATE OF 147.5 ST/D (91.8% RECOVERY)

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SUPERVISOR J. SMITH
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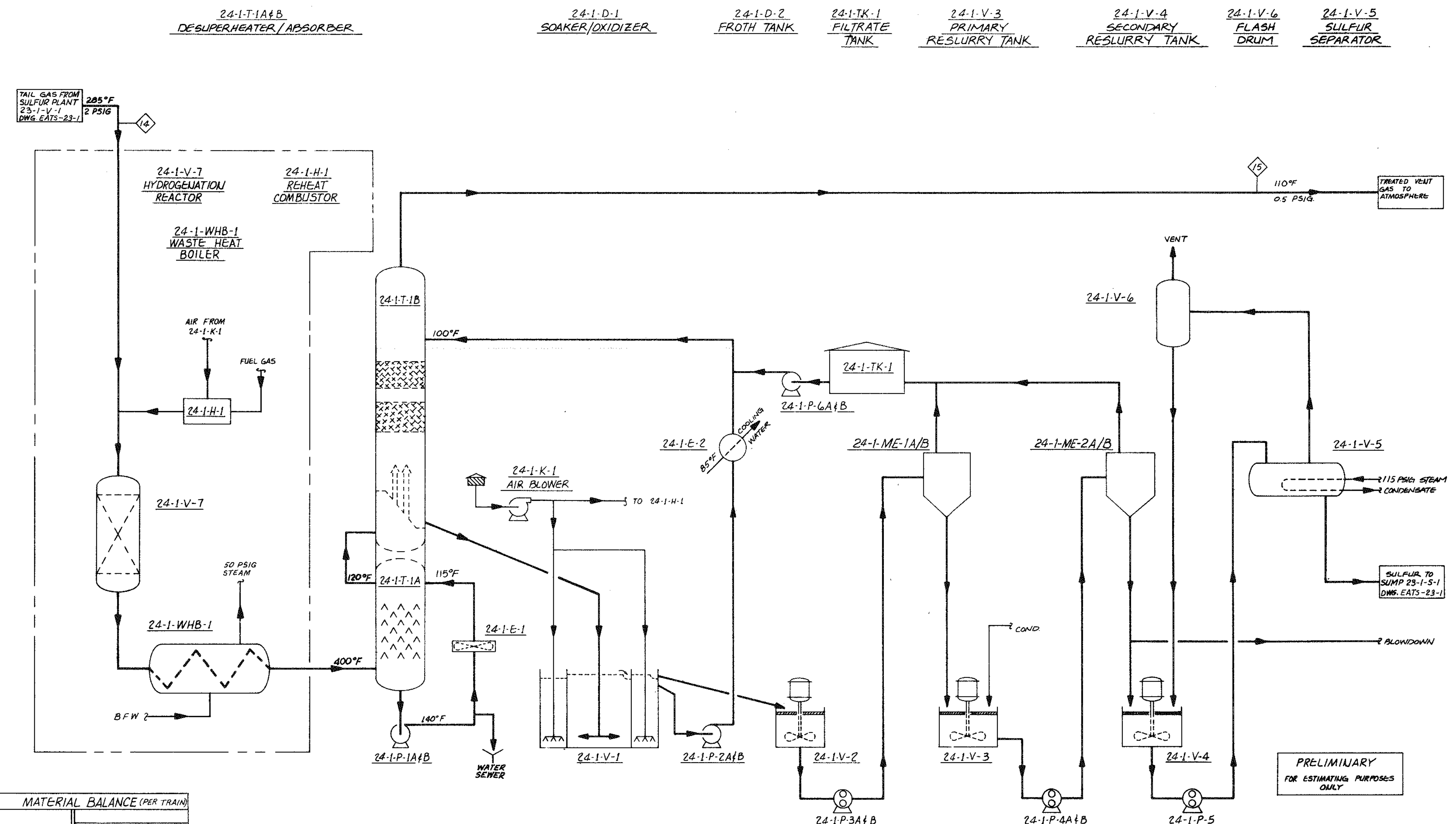
APP. DATE

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

SCALE NONE
DRAWING NUMBER EATS-23-1
REVISION 01 11-7-79



MATERIAL BALANCE (PER TRAIN)

STREAM No. / COMPONENT	14 TAIL GAS TO T.G. TREATING UNIT MPH	MOL %	15 ** VENT GAS FM. T.G. TREATING UNIT MPH	MOL %
H_2	4.86	0.19	4.86	0.23
CO	29.56	1.16	29.56	1.40
CO_2	1101.8	43.23	1104.3	52.27
H_2S	31.0	1.22	0.39*	100PPM
COS	2.5	0.10	0	0
N_2	837.9	32.87	837.9	39.67
NH_3	0	0	0	0
CH_4	0	0	0	0
H_2O VAPOR	540.3	.86	270.4	6.40
SO_2	0.86	0.03	–	0
S_2	0	21.19	–	0
S_6	.95	0	–	0
S_8	.23	.01	–	0
TOTAL VAPOR MPH	2549.1		2112.4	
H_2O LIQ.	0			
TOTAL LBS/HR	83870.9		77784.7	
MOL. WT. VAPOR	32.90		36.82	

* STATED AS H_2S BUT CONSISTS OF TOTAL SULFUR.

** EXCLUDING INCINERATION AND REDUCING GASES.

NOTES:

1. THIS PROCESS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF THREE 50% PARALLEL TRAINS (TWO OPERATING AND ONE SPARE).
2. ALL RATES ARE TYPICAL VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION, EXCEPT AS NOTED.
3. CONFIGURATION AND SIZE OF EQUIPMENT CORRESPOND TO DESIGN BASIS OF 14.17 ST/D PER TRAIN SULFUR RECOVERY.

DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

SCALE: NONE
DRAWING NUMBER: EATS-24-1
REVISION: 01 11-8-79

PROCESS CONDENSATE TREATING

Most of the sour process condensate generated is used for the preparation of the coal slurry feed to the gasifiers. A small stream of ammonia (as ammonium salts) contaminated effluent is produced in the gasification area (Flow Diagram: EATS-20-1). This effluent is small and treated in the effluent water treating unit (Unit 40).

A unit for the recovery of by-product ammonia is therefore not provided. The system for treatment of process condensate is similar to that provided for Case EXTC (Slurry Feed) as described in EPRI Report AF-642 (January 1978, page 327).

Equipment Notes

All equipment is commercially available.

STEAM, BOILER FEEDWATER, AND CONDENSATE

Process Flow Diagram EATS-30-1 schematically represents the steam, boiler feedwater, and condensate systems. This system is significantly different from that presented in EPRI Report AF-753 and is accordingly described in detail below. The description is typical for each of two parallel trains.

Process plant steam generation is closely integrated with the STEAG power generation unit. The integrated system operates at four steam pressure levels:

Turbine Inlet	- 2813 psig (195 bar), 995°F (535°C)
Intermediate Pressure (IP)	- 595 psig (42 bar)
Medium Pressure (MP)	- 115 psig (9 bar)
Low Pressure (LP)	- 50 psig (4.5 bar)

Boiler feedwater is pumped to the STEAG pressurized boiler 50-1-B-1 from the Deaerator 50-DA-1 by High-Pressure Boiler Feedwater Pumps 50-1-P-1A&B. A small slipstream of high-pressure boiler feedwater is withdrawn from the discharge of the pumps for steam desuperheating service. The remainder of the boiler feedwater stream is then preheated to 645°F before entering the boiler. This heating occurs in a series of boiler feedwater heaters, 50-1-E-6, 50-1-E-7, 50-1-E-8, and 50-1-E-9. The boiler feedwater heater, 50-1-E-7 is heated by a slipstream of steam from the 595 psig steam system. The heaters 50-1-E-8 and 50-1-E-9 are economizers which are heated by hot flue gas exiting from the Pressurized Boiler 50-1-B-1. The preheated boiler feedwater at 645°F is then divided into two streams. One stream flows to the evaporator/superheater coils of the STEAG pressurized boiler 50-1-B-1, E-1 and E-2 for generation of superheated 2813 psig steam from the firing of fuel gas. The other stream flows to the Gas Cooling Unit 20-1-ME-3 in the gasification area for the generation of superheated 2813 psig steam from gasifier effluent heat. The combined superheated HP steam streams then flow to the HP end of Turbine 50-1-T-1 for the generation of electric power. The HP steam enters 51-1-T-1 at 2813 psig and 995°F.

The HP end of the turbine, 50-1-T-1, exhausts steam at 595 psig (42 bar). This turbine exhaust steam then combines with saturated 595 psig steam produced in the process area. The saturated 595 psig steam from the process area is generated in the Claus plant waste heat boilers and the gas cooling units 20-1-ME-3. A large stream of 595 psig steam is diverted for boiler feedwater heating in 50-1-E-7 and 50-1-E-13. The remaining 595 psig steam is reheated to 977°F in the reheat coils of the STEAG Pressurized Boiler 50-1-B-1; E-3. The reheated intermediate-pressure (IP) steam then flows to the IP end of turbine 50-1-T-1.

The IP end of turbine 50-1-T-1 exhausts at 85 psig (6.9 bar). A portion flows to the Deaerator 50-DA-1. The remainder of this steam flows to the low-pressure end of turbine 50-1-T-2 for generation of electric power. The LP end of 50-1-T-2 is a condensing turbine exhausting at 2.5 inches Hg abs.

The 115 psig steam header is supplied by steam generation in the gas cooling units, 20-1-ME-3. This 115 psig steam is used in the fuel gas preheater 50-1-E-1, and in the sulfur separator heater, Unit 24. The remainder flows to the Deaerator 50-DA-1 as stripping steam.

The 50 psig steam header is supplied by steam generation from 50 psig Steam Generator 20-1-ME-3, and from the 50 psig Steam Generator in the Claus plant (Unit 23). The 50 psig steam is mainly used for boiler feedwater heating in 50-1-E-4. A small amount of 50 psig steam is used for steam tracing and also in the sulfur pit located in Unit 23.

The 15 psig steam header is supplied by steam generation from 15 psig Steam Generator 21-1-E-3A. A portion of the 15 psig steam is used for boiler feedwater heating in 50-1-E-2. The remainder flows to the low-pressure end of turbine 50-1-T-2 for generation of electric power.

Condensate from steam users throughout the plant is returned to the Deaerator, 50-DA-1. This includes vacuum steam condensate (2.5 inches Hg abs.) from the surface condenser coupled to condensing steam turbine 50-1-T-2. Condensate pumps located at the discharge of the surface condenser pump a part of the vacuum condensate through Condensate Heaters 50-1-E-1, 50-1-E-2, 50-1-E-3, and 50-1-E-4 and subsequently back to the deaerator. The remainder of the vacuum condensate is pumped as feedwater to various exchangers and steam generators, including Air Cooler 11-1-E-1, 15 psig Steam Generator 21-1-E-3A, HP steam generation in 20-1-ME-3 and Cold Condensate Heater 21-1-E-3B.

The Deaerator, 50-DA-1, in the STEAG cycle operates at 82 psig. Stripping steam is supplied from the 115 psig steam header and supplemented by extraction from the LP end of turbine 50-1-T-2. Boiler feedwater from the deaerator is pumped by

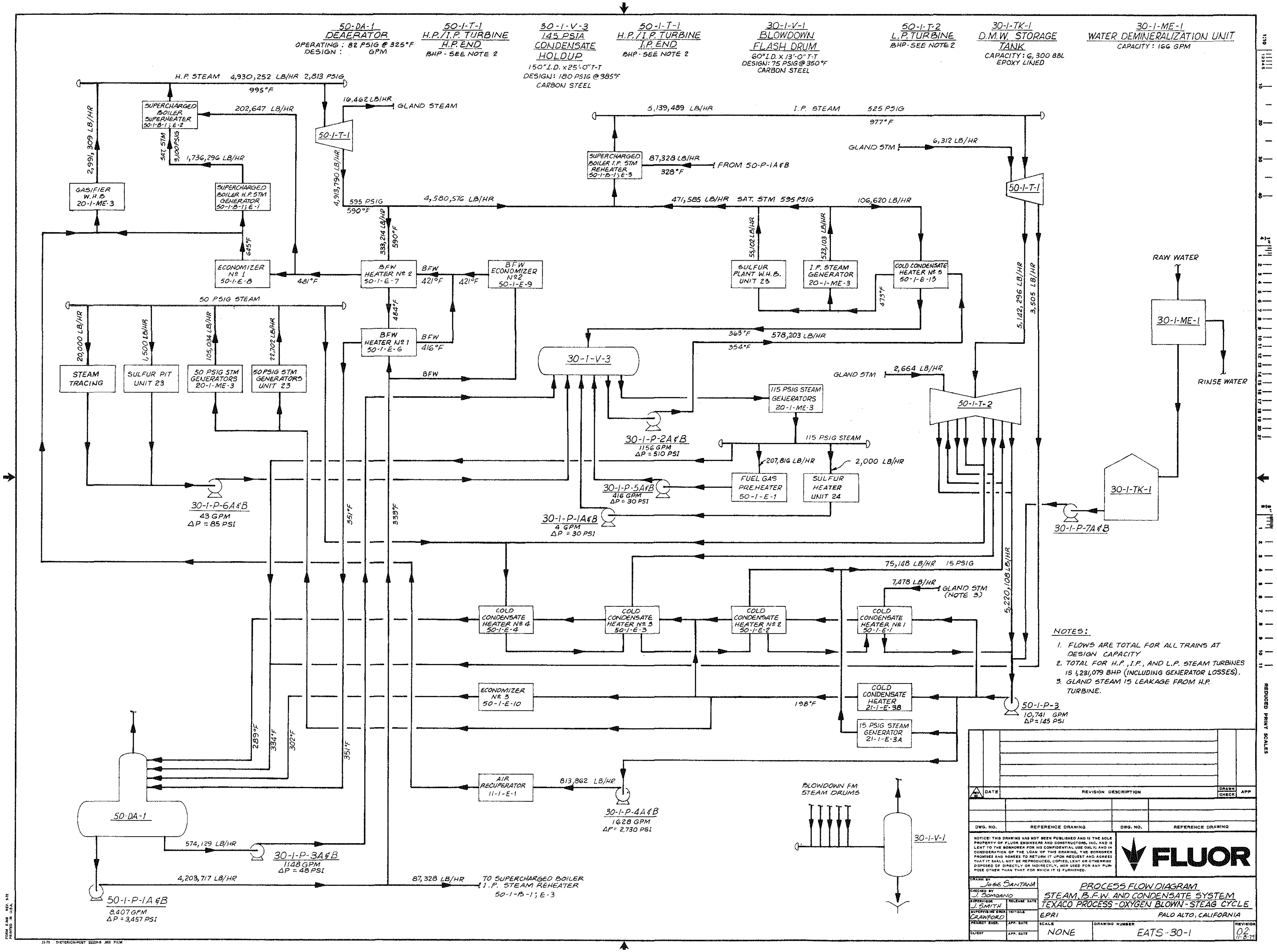
High-Pressure Boiler Feedwater Pumps (50-1-P-1A&B) to generation of HP steam as described above. Additional boiler feedwater from the Deaerator 50-DA-1 is pumped by IP boiler feedwater pumps (30-1-P-2A&B) to the gas cooling units, 20-1-ME-3, and Claus plant, Unit 23, for generation of 595 psig steam.

The recirculating boiler feedwater is continuously treated in a Benson-type water treating system for removal of dissolved salts as well as dissolved oxygen. This water treatment minimizes the makeup and blowdown requirements.

Raw water is treated in a semiautomatic, resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for the high-pressure boilers. Demineralized water is stored in Tank 30-TK-1. Demineralized water from the storage tank is transported to the deaerator through Pumps 30-P-7A&B.

Equipment Notes

All this equipment is commercially available.


50-DA-1 DEAERATOR
OPERATING: 82 PSIG @ 325°F
DESIGN: GPM
50-1-T-1 H.P./I.P. TURBINE H.P. END
BHP - SEE NOTE 2
30-1-V-3 145 PSIA CONDENSATE HOLDUP
150"I.D. x 25'-0"T-T
DESIGN: 180 PSIG @ 385°F
CARBON STEEL
50-1-T-1 H.P./I.P. TURBINE I.P. END
BHP - SEE NOTE 2
30-1-V-1 BLOWDOWN FLASH DRUM
60"I.D. x 13'-0"T-T
DESIGN: 75 PSIG @ 350°F
CARBON STEEL
50-1-T-2 L.P. TURBINE
BHP - SEE NOTE 2
30-1-TK-1 D.M.W. STORAGE TANK
CAPACITY: 6,300 BBL
EPOXY LINED
30-1-ME-1 WATER DEMINERALIZATION UNIT
CAPACITY: 166 GPM
H.P. STEAM 4,930,252 LB/HR 2,813 PSIG
995°F
SUPERCHARGED BOILER SUPERHEATER 50-1-B-1; E-2
202,647 LB/HR
16,462 LB/HR
GLAND STEAM
5,139,489 LB/HR
I.P. STEAM 525 PSIG
977°F
SAT. STM
3,100 PSIG
1,736,296 LB/HR
2,991,309 LB/HR
4,913,790 LB/HR
GASIFIER W.H.B. 20-1-ME-3
SUPERCHARGED BOILER H.P. STM GENERATOR 50-1-B-1; E-1
SUPERCHARGED BOILER I.P. STM REHEATER 50-1-B-1; E-3
87,328 LB/HR
328°F
FROM 50-P-1A & B
GLAND STM
6,312 LB/HR
595 PSIG
590°F
4,580,576 LB/HR
471,585 LB/HR SAT. STM 595 PSIG
106,620 LB/HR
645°F
333,214 LB/HR
590°F
56,102 LB/HR
523,103 LB/HR
ECONOMIZER Nº 1 50-1-E-8
481°F
BFW HEATER Nº 2 50-1-E-7
BFW
421°F
421°F
BFW ECONOMIZER Nº 2 50-1-E-9
SULFUR PLANT W.H.B. UNIT 23
I.P. STEAM GENERATOR 20-1-ME-3
COLD CONDENSATE HEATER Nº 5 50-1-E-13
473°F
5,142,296 LB/HR
3,505 LB/HR
RAW WATER
50 PSIG STEAM
20,000 LB/HR
1,500 LB/HR
105,034 LB/HR
22,202 LB/HR
464°F
BFW HEATER Nº 1 50-1-E-6
BFW
416°F
363°F
578,203 LB/HR
354°F
30-1-ME-1
RINSE WATER
STEAM TRACING
SULFUR PIT UNIT 23
50 PSIG STM GENERATORS 20-1-ME-3
50 PSIG STM GENERATORS UNIT 23
BFW
30-1-V-3
GLAND STM
2,664 LB/HR
115 PSIG STEAM GENERATORS 20-1-ME-3
50-1-T-2
30-1-P-2A & B
1156 GPM
ΔP = 510 PSI
115 PSIG STEAM
207,816 LB/HR
2,000 LB/HR
FUEL GAS PREHEATER 50-1-E-1
SULFUR HEATER UNIT 24
30-1-P-5A & B
416 GPM
ΔP = 30 PSI
30-1-TK-1
30-1-P-6A & B
43 GPM
ΔP = 85 PSI
30-1-P-1A & B
4 GPM
ΔP = 30 PSI
30-1-P-7A & B
351°F
339°F
75,148 LB/HR 15 PSIG
5,220,108 LB/HR
7,478 LB/HR
GLAND STM (NOTE 3)
COLD CONDENSATE HEATER Nº 4 50-1-E-4
COLD CONDENSATE HEATER Nº 3 50-1-E-3
COLD CONDENSATE HEATER Nº 2 50-1-E-2
COLD CONDENSATE HEATER Nº 1 50-1-E-1
NOTES:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY
2. TOTAL FOR H.P., I.P., AND L.P. STEAM TURBINES IS 1,231,079 BHP (INCLUDING GENERATOR LOSSES).
3. GLAND STEAM IS LEAKAGE FROM H.P. TURBINE.
ECONOMIZER Nº 3 50-1-E-10
198°F
COLD CONDENSATE HEATER 21-1-E-3B
50-1-P-3
10,741 GPM
ΔP = 145 PSI
15 PSIG STEAM GENERATOR 21-1-E-3A
289°F
334°F
302°F
351°F
AIR RECUPERATOR 11-1-E-1
813,862 LB/HR
30-1-P-4A & B
1628 GPM
ΔP = 2,730 PSI
BLOWDOWN FM STEAM DRUMS
30-1-V-1
50-DA-1
574,129 LB/HR
30-1-P-3A & B
1148 GPM
ΔP = 48 PSI
4,203,717 LB/HR
87,328 LB/HR
TO SUPERCHARGED BOILER I.P. STEAM REHEATER 50-1-B-1; E-3
50-1-P-1A & B
8,407 GPM
ΔP = 3,457 PSI
DATE
REVISION DESCRIPTION
DWG. NO.
REFERENCE DRAWING
FLUOR
DRAWN BY Jose Santana
CHECKED BY J. Somoano
SUPERVISOR J. SMITH
CRAWFORD
PROCESS FLOW DIAGRAM
STEAM, B.F.W. AND CONDENSATE SYSTEM
TEXACO PROCESS - OXYGEN BLOWN - STEAG CYCLE
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EATS-30-1
REVISION 02
REDUCED PRINT SCALES

COMBINED-CYCLE SYSTEM

Process Flow Diagram EATS-50-1 depicts the combined-cycle system designed by STEAG for Case EXTS. The diagram shows the total power block flows and also shows various sections of the gasification plant depicted on other flow diagrams, including the steam system and parts of the oxidant supply system.

The power block is divided into two independent parallel trains each of 50 percent capacity. Each train contains two pressurized boilers, 50-1-B-1, one gas turbine, 50-1-GT-1, one steam turbine, 50-1-T-1&2, and one steam turbine generator, 50-1-G-2. Each train is designated a "Kombi-block." Refer to Appendix A for a detailed description of the pressurized boiler combined-cycle system, including detailed performance information for the power block components, i.e., pressurized boilers, steam turbines, and gas turbines.

The following is a brief description of the process flow of fuel gas received from the gasification area. The steam system is described separately in connection with Process Flow Diagram EATS-30-1. Fuel gas enters the power block at 158 psig and 142°F from fuel gas expander 11-1-EX-1 in Unit 11. The fuel gas is then preheated to 302°F in 50-1-E-1 using 115 psig steam. A portion of the preheated fuel gas flows as fuel to fired heater 11-1-H-1 in Unit 11. The remainder is combined with a small flow of flash gas from Unit 22 and flows as fuel to the pressurized boiler 50-1-B-1.

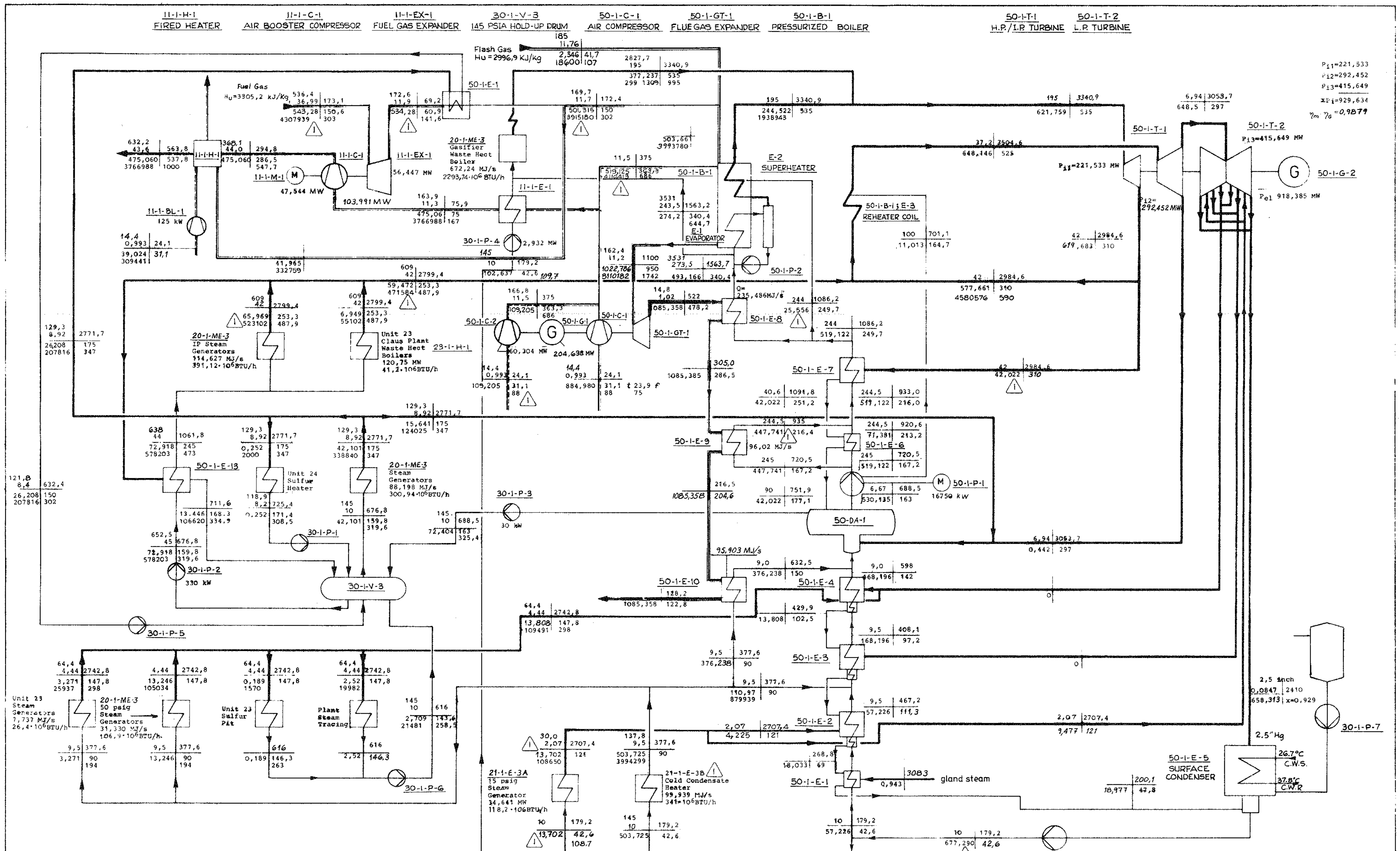
Both the fuel gas and flash gas streams are combusted in the pressurized boiler 50-1-B-1 to produce superheated high-pressure steam. Flue gases exhaust from the pressurized boiler at 1742°F and 148 psig and enter gas turbine 50-1-GT-1 for expansion to atmospheric pressure. The hot turbine exit gases then flow through economizers 50-1-E-8, 50-1-E-9, and 50-1-E-10 to preheat boiler feedwater flowing to the pressurized boiler 50-1-B-1, and finally exhaust to the atmosphere through a tall stack.

Gas turbine 50-1-GT-1 drives air compressors 50-1-C-1 and 50-1-C-2 which are direct coupled to the turbine, as well as electric generator 50-1-G-1. Air from the two compressors operating in parallel is combined and flows to the pressurized boiler for combustion of fuel gas. A portion of this air flows to the oxidant supply system in Unit 11.

Process cooling loads are integrated where possible into the power block steam and condensate system.

Equipment Notes

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



CONVERSION FACTORS

0,06896	bar	=	1	psi
0,126111	kg/s	=	10^3	lb/h
	$\frac{9}{5}$ °C+32	=	1	°F
1,055073	kJ	=	1	BTU
0,293075	MJ/s	=	10^6	BTU/h
2,32395	kJ/kg	=	1	BTU/lb
0,746	kW	=	1	BHP

FLOW LEGEND

- BFW & CONDENSATE
- STEAM
- AIR OR OXYGEN
- FUEL GAS

PSIA	
bar (abs)	kJ/kg
kg/s	°C
lb/h	°F

NOTES:

1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. ⚠1 = REVISED BY FLUOR.
3. THE COMMAS IN THE NUMBERS ON THIS DRWG ARE EQUIVLENT TO DECIMAL POINTS.

STEAG AG
Anlagentechnik

1978

EPRI-Fluor Study

steag

PROCESS FLOW DIAGRAM
STEAM SYSTEM & POWER GENERATION
TEXACO PROCESS-AIR BLOWN-STEAG CYCLE
DWG. EATS-50-1

17 11 1264 /8 006

FLUOR REV. 01 11-5-79 CONTRACT # 448334

PROCESS DISCUSSION - CASE EATS

The table below summarizes pertinent heat and material balance results.

Table 6-2

SUMMARY OF SYSTEM PERFORMANCE - CASE EATS

GASIFICATION AND GAS CLEANING SYSTEM	
Coal Feed Rate, lbs/hr (m.f.)	798,333
Oxygen*/Coal Ratio, lbs/lb m.f.	1.081
Oxidant Temperature, °F	1,000
Steam/Coal Ratio, lbs/lb m.f. (moisture in air)	0.0086
Slurry Water/Coal Ratio, lbs/lb m.f.	0.522
Gasification Section Average Pressure, psig	600
Crude Gas Temperature, °F	2,300-2,600
Crude Gas HHV (dry basis), Btu/SCF**	102.5
Temperature of Fuel Gas to Power System, °F	142
POWER SYSTEM	
Gas Turbine Inlet Temperature, °F (at pressurized boiler exit)	1,742
Turbine Air Compression Pressure Ratio	11.6:1
Turbine Exhaust Temperature, °F	893
Steam Cycle Conditions, psig/°F/°F (at turbine inlet)	2,814/995/977
Condenser Pressure, Inches Hg abs	2.5
Stack Temperature, °F	253
Gas Turbine Power#, MW	205
Steam Turbine Power#, MW	918
Power Consumed, MW	110
Net System Power, MW	1,013
OVERALL SYSTEM	
Process and Deaerator Makeup Water, gpm/1000 MW, Net	324
Cooling Tower Makeup Water, gpm/1000 MW, Net	11,588
Cooling Water Circulation Rate, gpm/MW	469
Cooling Tower Heat Rejection, % of Coal HHV	53.3
Air Cooler Heat Rejection, % of Coal HHV	0.0
Net Heat Rate, Btu/kWh	10,065
Overall System Efficiency (Coal → Power), % of Coal HHV	33.9

*Dry basis, 100 percent O_2
**Excluding the HHV of H_2S, COS and NH_3
#At generator terminals

GASIFIER MATERIAL BALANCE

The air-blown Texaco gasifier material balance for full capacity operation is given in Table 6-3 for the air-blown Texaco gasifier.

Most of the data presented in the above table was received from Texaco Development Corporation for an earlier study published as EPRI Report AF-753. That study projected economics and operating data for the 1980-1985 time frame. For the particular coal used, Texaco indicated that slurry concentrations in the range of 60 percent solids to possibly 70 percent solids could be achieved in the 1980-1785 time frame. For study purposes, EPRI selected a slurry concentration of 65.7 percent solids for Case EATS. 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 material balance, including oxygen consumption, is based on a Texaco extrapolation of the state of the art to a period three to five years hence.

The required number of gasifiers for this air-blown case is not stated in this report because Texaco considers the information proprietary and subject to refinement. The basis used has been described in EPRI Report AF-753.

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 salts 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.

The figures given in the Tables and Flow Sheets for ammonia should be regarded as tentative estimates only.

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

Table 6-3

MATERIAL BALANCE - CASE EATS

FEEDS

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

EFFLUENTS

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

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. This process is used in these cases because it accomplishes this objective and it exhibits favorable economics when compared with other similar processes.

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 only.

PROCESS ENERGY BALANCES

Table 6-4 presents an overall process energy balance at 100 percent capacity operation. The boundary for each balance encompasses the entire plant, exclusive of the cooling tower heat balance. 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. The energy balance closes to less than one-half of one percent. The discrepancy results from approximations used for some process units and for calculating some heat loads.

Data from Table 6-4 is shown in MM Btu/hr and as percent of coal higher heating value in Table 6-5. Coal charged at 10,000 ton/day is equivalent to 10,196 MM Btu/hr HHV. This feed produces 3458 MM Btu/hr power equivalent or 33.9 percent of the coal HHV as net electric power. The heat rate based on net power produced is 10,065 Btu/kWh. Heat rejected at all steam turbine surface condensers is 5012 MM Btu/hr or 49.2 percent of the coal HHV. Heat rejected with the power block stack gases is 849 MM Btu/hr or 8.3 percent of the coal HHV.

Table 6-4

ENERGY BALANCE - CASE EATS

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 Compressors Suction Air		54	131			185
Fired Heater Combustion Air		2	5			7
Demineralized and Raw Water		1				1
Auxiliary Power Inputs					375	375
TOTAL	10,196	62	136	0	375	10,769
HEAT OUT						
Ash Cake		6				6
Gasifier Heat Losses				112		112
Gas Cooling		68				68
Sulfur Product	105	1				106
Fired Heater Flue Gas		48	35			83
Gas Turbines					699	699
Sulfur Plant Effluent Gas		2	18			20
Steam Turbines					3,134	3,134
Power Block Losses				11	88	99
Turbogenerator Condensers			5,012			5,012
Power Block Stack Losses		415	434			849
Motor Losses (Air Cooler Fans, etc.)					133	133
Steam Heat Losses		21				21
Selexol Overhead Condenser			68			68
Selexol Solvent Cooler		176				176
Process Water Cooling		121				121
Waste Water Effluent		28				28
TOTAL	105	886	5,567	123	4,054	10,735

$$\frac{\text{Input - Output}}{\text{Input}} = 0.32\%$$

Table 6-5

ENERGY BALANCE AS PERCENT COAL HHV - CASE EATS

	Case EATS	
	MM Btu/hr	Percent
IN		
Coal HHV	10,196	100.0
OUT		
Net Power	3,458	33.92
Sulfur Product, HHV	106	1.04
Selexol Sensible and Latent	244	2.39
Ash Cake	6	0.06
Power Block Stack Losses	849	8.33
Fired Heater Flue Gases	83	0.81
Rejected at Condensers	5,012	49.16
Other Sensible Losses, Net	178	1.75
Other Latent Losses, Net	(118)	(1.16)
Gasifier Heat Losses	112	1.10
Power Block Losses	99	0.97
Motor Losses	133	1.30
	10,162	99.67

ECONOMICS - CASE EATS

Table 6-6 summarizes the economics of Case EATS.

Table 6-6

SUMMARY OF ECONOMIC RESULTS - CASE EATS

PRODUCTION AT DESIGN CAPACITY	
Net Power, MW*	1,013
Overall Plant Heat Rate, Btu/kWh	10,065
TOTAL CAPITAL REQUIREMENTS**	
Total Capital Requirement, $1,000	1,194,422
Total Capital Requirement, $/kW	1,179
AVERAGE COSTS OF SERVICES**	
First Year Cost, $1,000/year	315,508
First Year Cost, mills/kWh	50.79
Thirty Year Levelized Cost, $1,000/year	407,438
Thirty Year Levelized Cost, mills/kWh	65.59

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

TOTAL PLANT INVESTMENT

Table 6-7 gives a detailed breakdown of the Total Plant Investment required for Case EATS. The accuracy of the plant investment estimate is judged to be ±25 percent. Since other capital charges and working capital are keyed to elements of plant investment, this accuracy is reflected in other capital figures as well. However, due to the similar nature of the cases in this evaluation, the

Table 6-7

TOTAL PLANT INVESTMENT - CASE EATS

Field Plant Section	Cost Breakdown Without Contingencies: Direct Material#	Direct Field Labor##	Engr. & Support Costs§	Total Sales Tax	Total Cost $1,000*	Cost $/kW**	Per-cent	Contingencies: Process $1,000*	Project $1,000*	Total Plant Investment: $1,000*	$/kW**
Coal Handling	10,714	4,524	7,262	540	23,040	22.74	2.85	-0-	3,456	26,496	26.16
Oxidant Feed	22,053	8,875	14,504	1,123	46,555	45.96	5.77	-0-	6,983	53,538	52.85
Gasification and Ash Handling	34,874	8,729	16,535	1,763	61,901	61.11	7.67	8,988	9,285	80,174	79.15
Gas Cooling	133,385	62,937	99,550	6,913	302,785	298.90	37.51	75,057	45,418	423,260	417.83
Acid Gas Removal and Sulfur Recovery	21,573	8,051	13,342	1,085	44,051	43.48	5.45	935	6,608	51,594	50.93
Steam, Condensate and B&W	1,108	469	749	56	2,382	2.35	.30	-0-	357	2,739	2.70
Combined Cycle	140,184	42,920	76,725	7,224	267,053	263.63	33.09	13,353	40,058	320,464	316.35
General Facilities	30,439	12,129	15,854	956	59,378	58.62	7.36	0	8,907	68,285	67.41
Subtotal	394,330	148,634	244,521	19,660	807,145	796.79	100.00	98,333	121,072	1,026,550	1,013.38

TOTAL PLANT INVESTMENT SUMMARY

	$1,000*	$/kW**
Process Plant Investment and General Facilities	807,145	796.79
Process Contingency	98,333	97.07
Project Contingency	121,072	119.52
Total Plant Investment	1,026,550	1,013.38

*Mid-1976 Dollars.
**Based on 100 Percent Plant Design Power Output (973 MW).
#All materials and equipment that become a part of the plant facility.
##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits).
§Includes: a) Indirect field costs including all labor, supervision and expenses required to support field construction; b) Home office costs including all salaries and expenses required for engineering design and procurement; and c) Contractor's fee.

estimates for all cases should reflect about the same accuracy. When these estimates are used comparatively, the effect of individual accuracies should be minimal.

Two contingencies are included in the Total Plant Investment shown for each plant section. First is a 15 percent project contingency which is included 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 innovative technology in an effort to quantify the uncertainty in the design, performance, and cost of the commercial scale equipment. This covers the additional expenditures required to solve any problems associated with innovative technology, since historically such technology requires more cost than initially estimated. Although all plant technology is judged to be probably commercially obtainable, not all of it is fully proven by commercial operation. Accordingly, a process contingency is applied and included in the estimated plant investment for sections containing innovative technology.

TOTAL CAPITAL REQUIREMENT

Table 6-8 gives a breakdown of the Total Capital Requirement to place the plant into initial operation at 100 percent capacity. Starting with the total plant investment, capital charges are added for allowance for funds during construction, initial catalyst and chemicals charge, prepaid royalties, preproduction costs, the startup inventory capital, and the land required for the plant. Specific items included in each of these capital charges are described under the economic criteria.

COST OF SERVICES

Table 6-9 gives a cost of services breakdown for Case EATS. The costs are busbar power costs based on plant operation at a 70 percent capacity factor with a $1/MM Btu coal cost. Table 6-9 shows both first-year and 30-year levelized power costs. Since the power costs vary with the plant capacity factor, this additional relationship is shown in Figure 6-1 on both a first-year and 30-year levelized basis. The plant capacity factor presented in Figure 6-1 is the percent plant time on-line at baseload (100 percent) plant power output of 1013 MW. This capacity factor is not equivalent to plant operation in a turndown mode producing less than baseload power output, as the design assumes a baseload plant.

Table 6-8

TOTAL CAPITAL REQUIREMENT - CASE EATS

	$1000*	$/kW**
TOTAL PLANT INVESTMENT	1,026,550	1,013.38
CAPITAL CHARGES		
Prepaid Royalties	3,739	3.69
Preproduction Costs	25,668	25.34
Inventory Capital	7,487	7.39
Initial Catalyst and Chemicals Charge	1,762	1.74
Allowance for Funds During Construction	128,216	126.57
Land	1,000	0.99
Total Capital Charges	167,872	165.72
TOTAL CAPITAL REQUIREMENT	1,194,422	1,179.10

*Mid-1976 dollars

**Based on 100 percent plant design power output

The largest single component of the power cost is contained in the levelized fixed charges. These charges represent repayment from operating revenue of the financing originally used to supply the total capital required for construction and startup of the plant. These levelized fixed charges amount to 68 percent of the first year power costs at a 70 percent capacity factor. The charges shown are fixed annual expenses which are independent of the plant capacity factor.

Coal constitutes the major operating charge. This charge amounts to 20 percent of the first-year power costs at a 70 percent capacity factor.

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

- "A" Operators 5
- "B" Operators 20
- Foremen 2
- Lab and Instrument Technicians 4

Table 6-9

BUSBAR POWER COST AT 70 PERCENT CAPACITY FACTOR - CASE EATS

	Air-Blown Texaco EATS	
NET PRODUCTION*		
Net Power, MW	1,013	
By-product Ammonia, ST/D	-0-	
By-product Sulfur, ST/D	309	
TOTAL CAPITAL REQUIREMENT, $1,000	1,194,422	
	First-Year Cost	30-Year Levelized Cost
FIXED OPERATING COST, $1,000/YEAR		
Operating Labor	3,395	6,403
Maintenance Labor	10,885	20,529
Maintenance Materials	16,328	30,795
Administrative & Support Labor	4,284	8,079
Total Fixed O&M Costs	34,892	65,806
VARIABLE OPERATING COSTS (EXCLUDING COAL), $1,000/YEAR		
Raw Water	1,753	3,306
Catalysts & Chemicals	365	689
Ash Disposal	981	1,850
Total Variable O&M Costs	3,099	5,845
COAL COST, $1,000/YEAR	62,521	120,790
BY-PRODUCT CREDITS, $1,000/YEAR		
By-product Ammonia	-0-	-0-
By-product Sulfur	-0-	-0-
Total By-product Credits	-0-	-0-
TOTAL OPERATING COSTS, $1,000/YEAR	100,512	192,441
LEVELIZED FIXED CHARGES, $1,000/YEAR	214,996	214,996
TOTAL COST OF ELECTRICITY**		
$1,000/year	315,508	407,437
mills/kWh	50.79	65.59

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

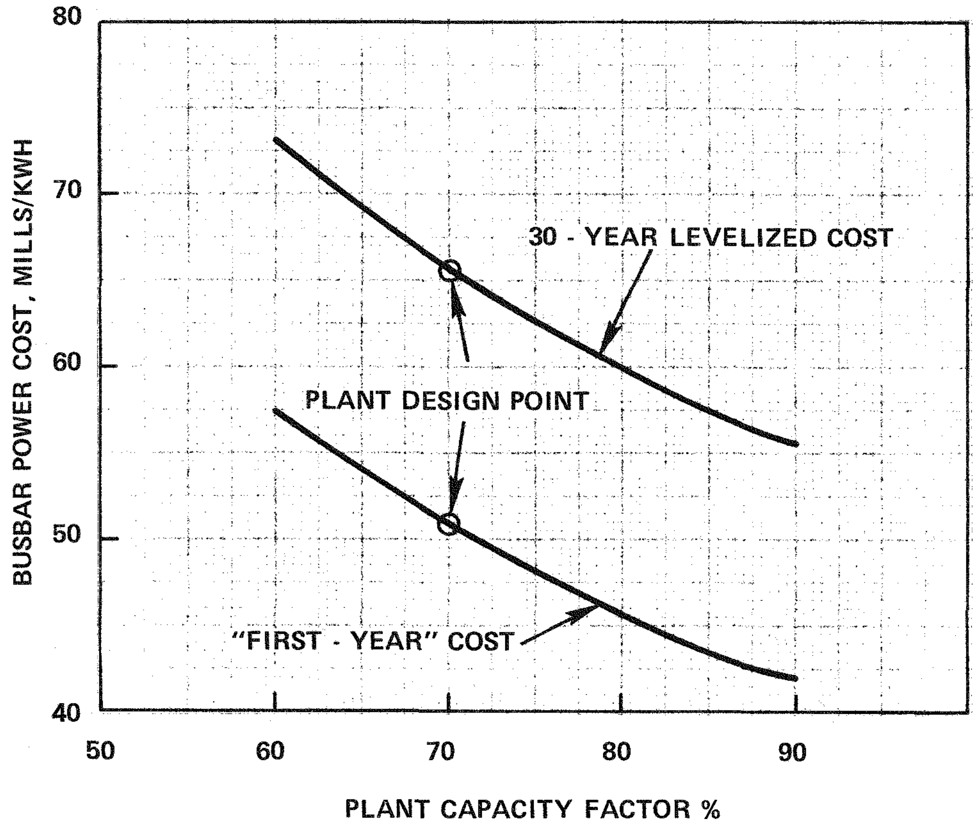


Figure 6-1. Power Cost Sensivity Curve for Case EATS

Maintenance labor and materials costs are calculated for each section as a percentage of the total plant investment for that section. These percentages are listed in the economic criteria section.

Catalyst and chemical costs are for chemicals consumed in the demineralized, cooling tower and boiler feedwater treating, plus costs associated with making up Selexol solvent losses in the acid gas removal, and tail gas treating units, and replacement of catalyst in the sulfur recovery unit.

Section 7

PLANT DESCRIPTION - CASE EXTC78

GENERAL

A grass roots plant for electric power generation based on single-stage, entrained-bed, oxygen-blown gasifiers of the Texaco type, integrated with combined-cycle generating equipment, is shown schematically on the Block Flow Diagram EXTC78-1-1. This plant consumes 10,000 ST/day of Illinois No. 6 coal.

The Case EXTC78 plant is similar to the Texaco-based plant presented as Case EXTC (Slurry Feed) in EPRI Report AF-642, but uses combustion turbines which are commercially available in 1978 in the power generation section. The process arrangement and number of parallel operating trains are shown on Block Flow Diagram EXTC78-1-1.

Table 7-1 summarizes major equipment sections in the plant and shows the number of operating and spare sections.

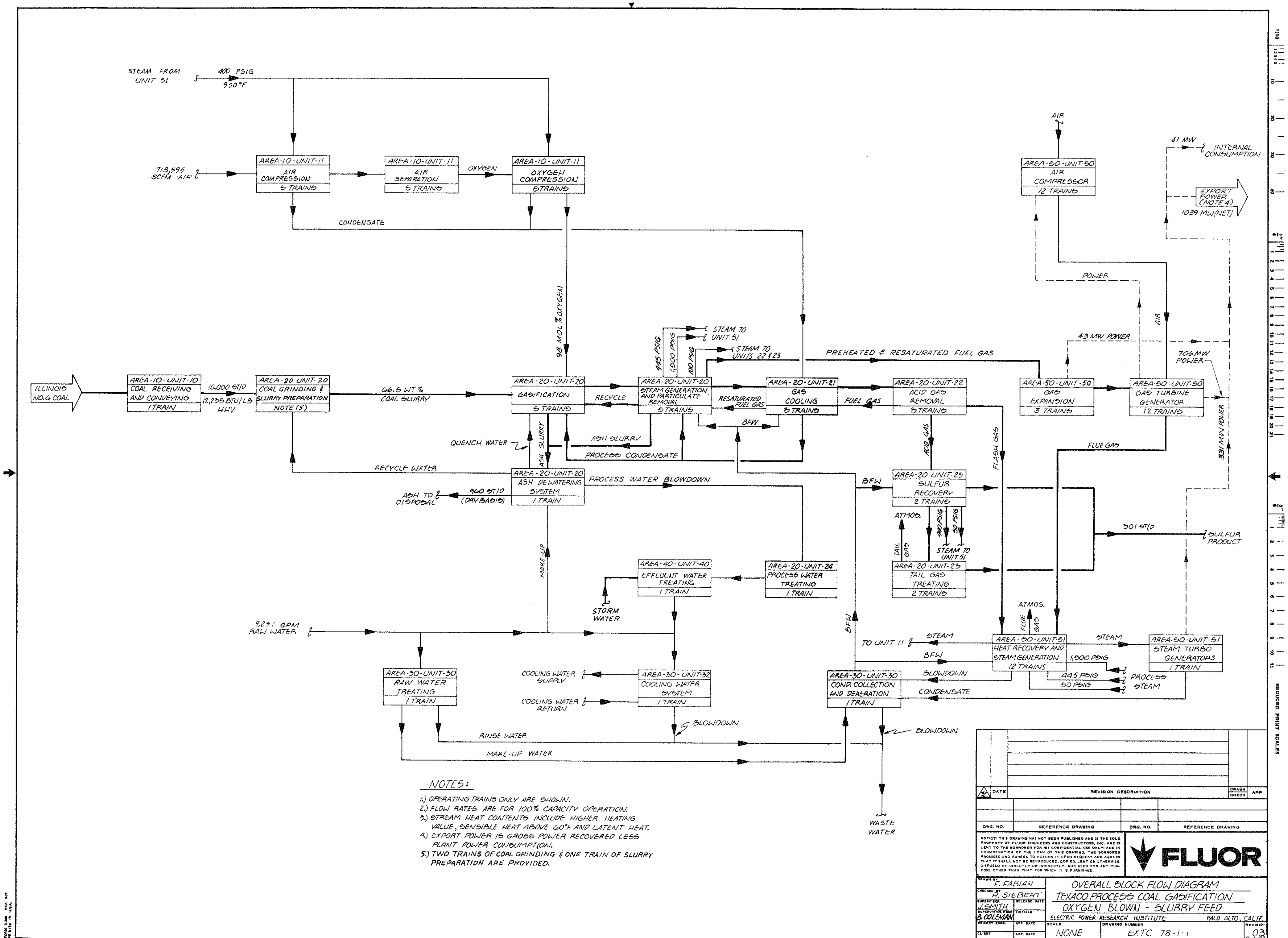
The following sections of this report contain descriptions of the major process units in the Case EXTC78 plant. Emphasis is placed on describing the differences in process design between the Case EXTC78 units and similar units shown for Case EXTC (Slurry Feed) in EPRI Report AF-642.

Table 7-1

TRAINS OF EQUIPMENT IN MAIN PROCESSING UNITS - CASE EXTC78

Unit No.	Name	Case EXTC78 Operating	Spare
10	Coal Receiving and Conveying	1	0
11	Oxidant Feed	5	0
20	Wet Coal Grinding	2	0
20	Slurry Preparation	1	0
20	Gasification	5	1
20	Ash Handling	1	0
20	Particulate Scrubbing	5	1
21	Gas Cooling	3	0
22	Acid Gas Removal	3	0
23	• Sulfur Recovery	2	1
	• Tail Gas Treating	2	1
24	Process Water Treating	1	0
30	Steam, BFW and Condensate System		
	• Condensate Collection and Deaeration	1	0
	• Raw Water Treating	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0
50	Gas Turbine/Generator	12*	0
51	Heat Recovery Steam Generator	12*	0
51	Steam Turbine/Generator	1	0

*Each of the 12 gas turbine/generator units consist of two gas turbines coupled to a single heat recovery steam generator



STEAM FROM UNIT 51
400 PSIG 900°F
713,595 SCFM AIR
AREA-10-UNIT-11 AIR COMPRESSION 5 TRAINS
AREA-10-UNIT-11 AIR SEPARATION 5 TRAINS
OXYGEN
AREA-10-UNIT-11 OXYGEN COMPRESSION 5 TRAINS
CONDENSATE
98 MOL % OXYGEN
ILLINOIS NO.6 COAL
AREA-10-UNIT-10 COAL RECEIVING AND CONVEYING 1 TRAIN
10,000 ST/D
12,255 BTU/LB HHV
AREA-20-UNIT-20 COAL GRINDING & SLURRY PREPARATION NOTE (5)
66.5 WT % COAL SLURRY
AREA-20-UNIT-20 GASIFICATION 5 TRAINS
RECYCLE
445 PSIG
1,500 PSIG
100 PSIG
STEAM TO UNIT 51
STEAM TO UNITS 22 & 23
AREA-20-UNIT-20 STEAM GENERATION AND PARTICULATE REMOVAL 5 TRAINS
RESATURATED FUEL GAS
BFW
PREHEATED & RESATURATED FUEL GAS
AREA-20-UNIT-21 GAS COOLING 5 TRAINS
FUEL GAS
AREA-20-UNIT-22 ACID GAS REMOVAL 5 TRAINS
ACID GAS
FLASH GAS
AREA-50-UNIT-50 GAS EXPANSION 3 TRAINS
AREA-50-UNIT-50 GAS TURBINE GENERATOR 12 TRAINS
AIR
AREA-50-UNIT-50 AIR COMPRESSOR 12 TRAINS
POWER
43 MW POWER
706 MW POWER
331 MW POWER
41 MW
INTERNAL CONSUMPTION
EXPORT POWER (NOTE 4)
1039 MW (NET)
FLUE GAS
QUENCH WATER
ASH SLURRY
PROCESS CONDENSATE
RECYCLE WATER
AREA-20-UNIT-20 ASH DEWATERING SYSTEM 1 TRAIN
ASH TO DISPOSAL
960 ST/D (DRY BASIS)
PROCESS WATER BLOWDOWN
BFW
AREA-20-UNIT-23 SULFUR RECOVERY 2 TRAINS
ATMOS.
TAIL GAS
STEAM TO UNIT 51
501 ST/D
SULFUR PRODUCT
AREA-20-UNIT-23 TAIL GAS TREATING 2 TRAINS
MAKE-UP
AREA-40-UNIT-40 EFFLUENT WATER TREATING 1 TRAIN
AREA-20-UNIT-24 PROCESS WATER TREATING 1 TRAIN
STORM WATER
9,251 GPM RAW WATER
BFW
ATMOS.
FLUE GAS
TO UNIT 11
STEAM
AREA-50-UNIT-51 HEAT RECOVERY AND STEAM GENERATION 12 TRAINS
1,500 PSIG
445 PSIG
50 PSIG
PROCESS STEAM
AREA-50-UNIT-51 STEAM TURBO GENERATORS 1 TRAIN
AREA-30-UNIT-30 RAW WATER TREATING 1 TRAIN
COOLING WATER SUPPLY
COOLING WATER RETURN
AREA-30-UNIT-32 COOLING WATER SYSTEM 1 TRAIN
BLOWDOWN
AREA-30-UNIT-30 COND. COLLECTION AND DEAERATION 1 TRAIN
BLOWDOWN
CONDENSATE
RINSE WATER
MAKE-UP WATER
WASTE WATER
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.
REVISION DESCRIPTION
REFERENCE DRAWING
FLUOR
F. FABIAN
R. SIEBERT
J. SMITH
B. COLEMAN
OVERALL BLOCK FLOW DIAGRAM
TEXACO PROCESS COAL GASIFICATION
OXYGEN BLOWN - SLURRY FEED
ELECTRIC POWER RESEARCH INSTITUTE
PALO ALTO, CALIF.
NONE
EXTC 78-1-1
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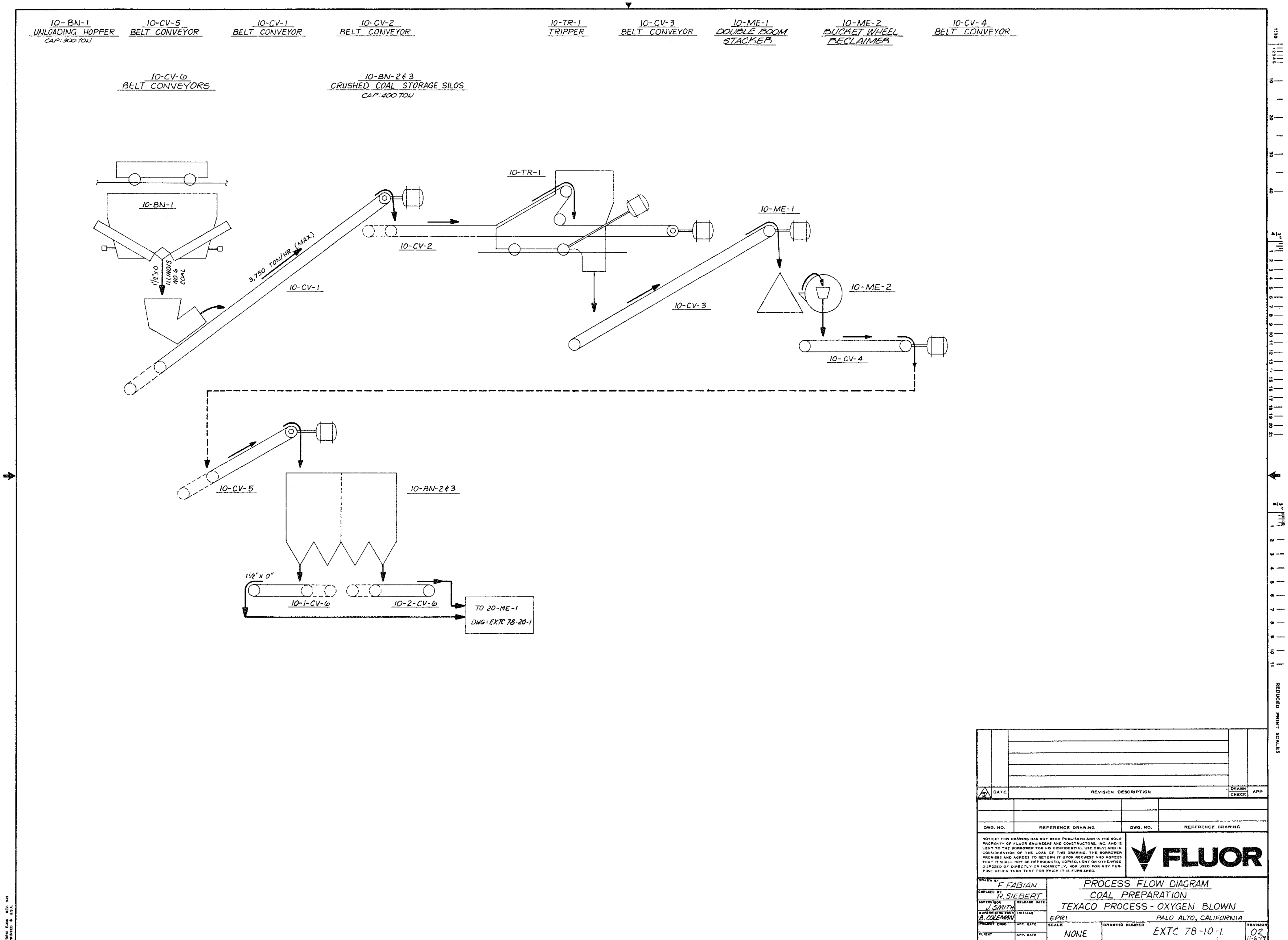
COAL PREPARATION

Process Flow Diagram EXTC78-10-1 depicts the process arrangement of equipment in this section for Case EXTC78. Coal preparation section process details are essentially the same as those described for Case EXTC (Slurry Feed) in EPRI Report AF-642 (January 1978, page 304).

Equipment Notes

All the equipment in this section 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
1½"x0 ILLINOIS NO. 6 COAL
3,750 TON/HR (MAX.)
1½" x 0"
10-1-CV-6
10-2-CV-6
TO 20-ME-1
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TEXACO PROCESS - OXYGEN BLOWN
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EXTC 78-10-1
REVISION 02 11-6-79
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OXIDANT FEED

Process Flow Diagram EXTC78-11-1 shows the gasifier oxidant feed system for Case EXTC78. The process details are similar to those presented for Case EXTC (Slurry Feed) in EPRI Report AF-642 (January 1978, page 307), with the exception that Fuel Gas Expander 50-1-EX-1 is not coupled to Air Compressor 11-1-C-1, but instead drives an electric generator. Thus, the total compressor power requirement will be supplied by Air Compressor Turbine 11-1-T-1, resulting in a simpler drive train. Motive steam inlet temperature for the condensing driver turbines 11-1-T-1 and 11-1-T-2 has been reduced to 900°F to be compatible with both the Case EXTC78 HRSG design and the requirements of current, commercially available machines.

Equipment Notes

All equipment is commercially available.

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-V-2
K.O. DRUM
162" I.D. x 13'-0" T/T
DESIGN: 50 PSIG @ 175°F
CARBON STEEL

11-1-C-1
AIR COMPRESSOR
34,100 BHP

11-1-T-1
AIR COMP TURBINE
34,100 BHP

11-1-E-1
SURFACE CONDENSER
196 MM BTU/HR
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-E-10
SURFACE CONDENSER
67 MM BTU/HR
TUBES: CARBON STEEL
SHELL: CARBON STEEL

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-T-2
OXYGEN COMP TURBINE
11,700 BHP

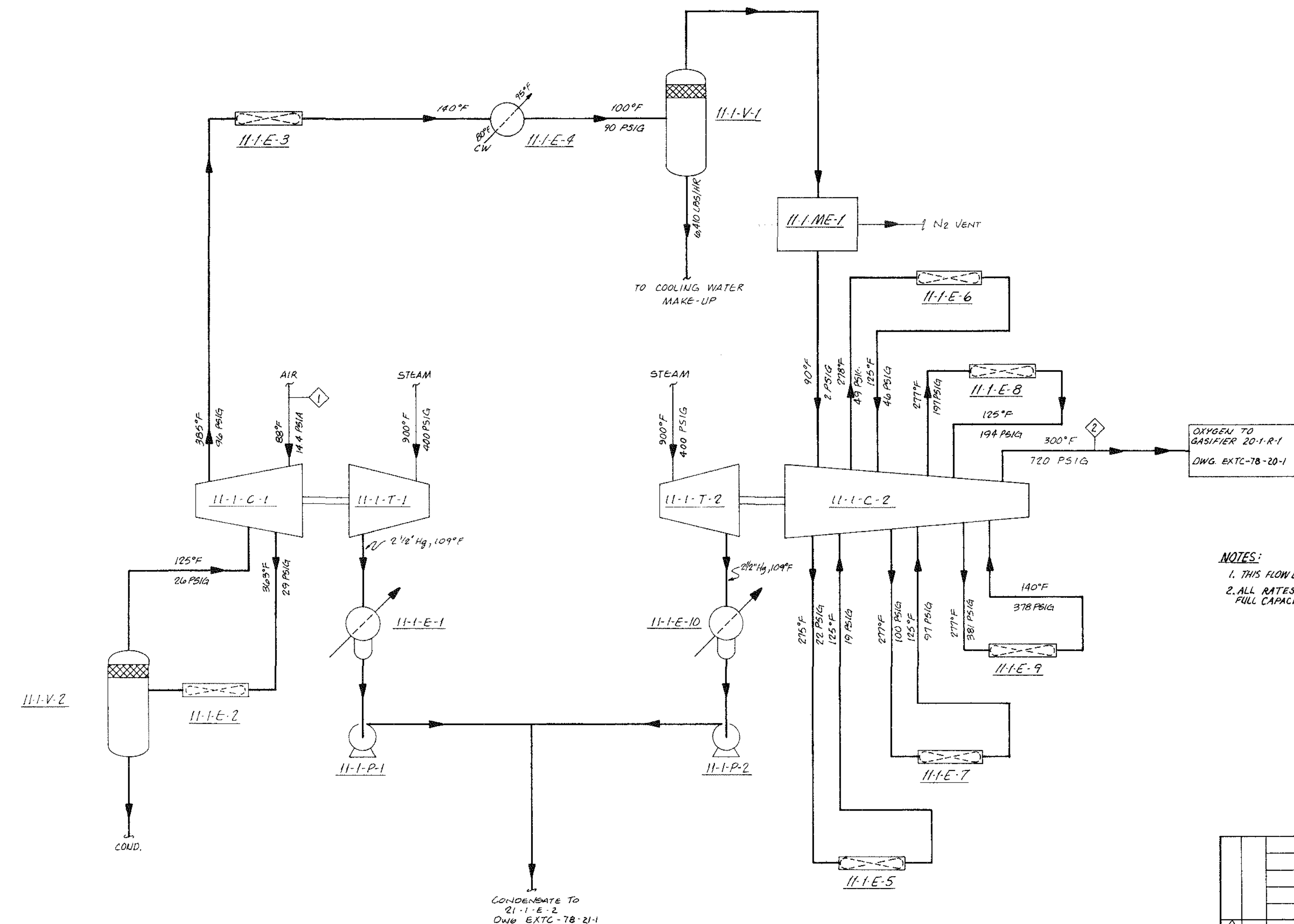
11-1-E-5
1ST STAGE O_2 COOLER
4.7 MM BTU/HR
3,489 SQ. FT. BARE
TUBES: 316 SS

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



NOTES:

1. THIS FLOW DIAGRAM SHOWS ONE TRAIN OF A TOTAL OF FIVE PARALLEL TRAINS.
2. ALL RATES 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	4279.5	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			

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APP. DATE:
CLIENT:
APP. DATE:

PROCESS FLOW DIAGRAM
OXIDANT FEED SYSTEM
TEXACO PROCESS-OXYGEN BLOWN

EPRI — PALO ALTO, CALIFORNIA

SCALE: NONE
DRAWING NUMBER: EXTC 78-11-1
REVISION: 03 11-6-79

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GASIFICATION AND ASH HANDLING

Process Flow Diagram EXTC78-20-1 shows the gasification step for Case EXTC78. The details of gasification and ash handling are similar to Case EXTC (Slurry Feed) described in EPRI Report AF-642 (January 1978, pages 312 and 313). Raw hot gas flows from the gasifier to Gas Cooling Unit 20-1-ME-3 where high-pressure (HP) steam is generated at 1500 psig, saturated intermediate-pressure (IP) steam at 445 psig and saturated medium-pressure (MP) steam at 100 psig. Clean fuel gas which has been saturated with heated water from Unit 21, is exchanged against the raw hot gas in 20-1-ME-3 before being sent to Fuel Gas Expander 50-EX-1 at a temperature of 1000°F. Details of the resaturator are discussed in the section for Gas Cooling, Unit 21.

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: EXTC78-21-1) 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 shown in Flow Diagram EXTC78-21-1.

Equipment Notes

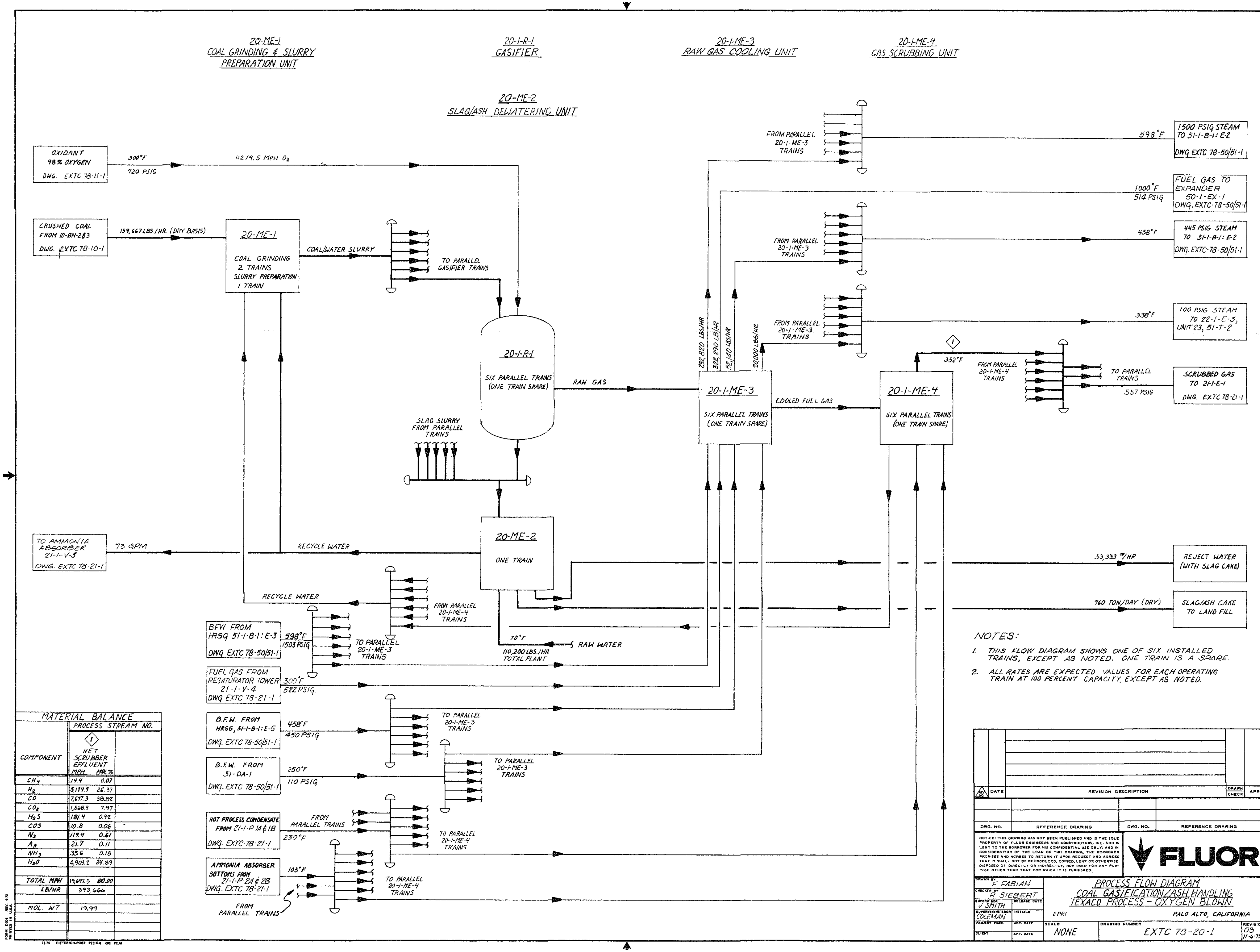
The Texaco gasifier is commercially proven for the gasification of liquid hydrocarbons. Coal gasification is still in the pilot plant stage.

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 currently undergoing test runs in Germany.

The slag dewatering unit is commercially proven.

The gas scrubbing unit equipment is commercially available.

The steam generation pressure levels are within the range of current technology.



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NO. 1 NET SCRUBBER EFFLUENT MPH	MOL %	
CH_4	14.4	0.07	
H_2	5,194.9	26.37	
CO	7,647.3	38.82	
CO_2	1,568.9	7.97	
H_2S	181.4	0.92	
COS	10.8	0.06	
N_2	119.4	0.61	
A_R	21.7	0.11	
NH_3	35.6	0.18	
H_2O	4,903.2	24.89	
TOTAL MPH	19,697.5	100.00	
LB/HR	393,666		
MOL. WT.	19.99		

GAS COOLING

Process Flow Diagram EXTC78-21-1 shows one of three parallel trains in the gas cooling section. No spare train is provided. The process details for each train are significantly different from those described for Case EXTC (Slurry Feed) in EPRI Report AF-642 (January 1978, page 317) and are accordingly described in more detail below.

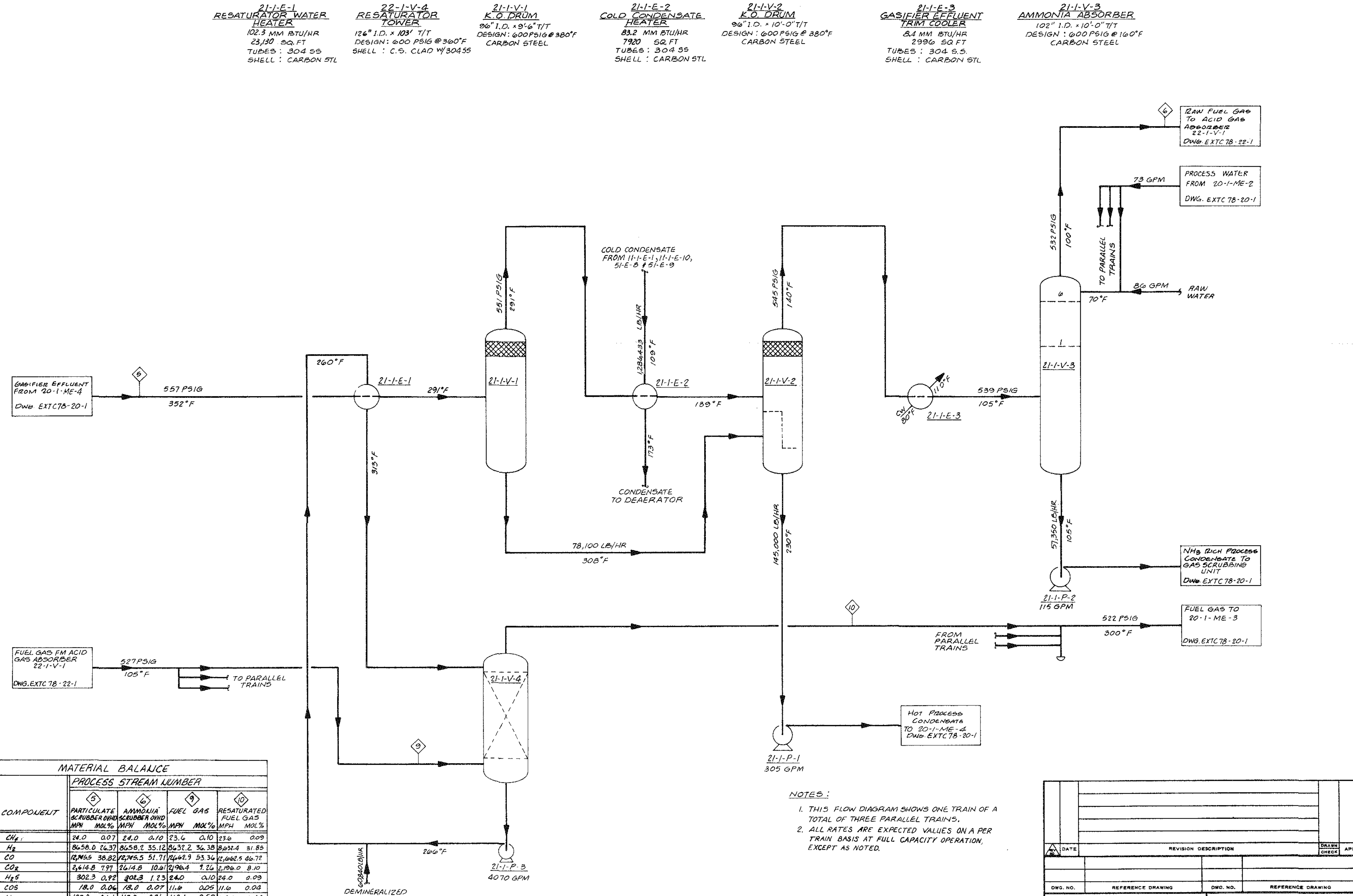
Clean gasifier effluent from the particulate scrubbing section is cooled to 105°F in a series of exchangers, 21-1-E-1, 21-1-E-2, and 21-1-E-3. Exchanger 21-1-E-1 heats resaturator tower bottoms water. Exchanger 21-1-E-2 heats vacuum condensate water recovered from the various surface condensers in the plant. Exchanger 21-1-E-3 then cools the gas to 105°F with cooling water. Exchangers 21-1-E-1 and 21-1-E-2 are each followed by a knockout drum to recover condensed process water from the effluent gas stream. These condensed water streams are combined and pumped to the gasification unit (Flow Diagram: EXTC78-20-1). Effluent gases from the last exchanger in the train, 21-1-E-3, flow to an ammonia absorber, 21-1-V-3. Ammonia is removed by contacting the gas countercurrently with the water on the trays of the absorber. 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 Gas Scrubbing Unit, 20-1-ME-4.

The primary difference between the gas cooling units of Case EXTC and Case EXTC78 is the addition of Resaturator Tower, 21-1-V-4, and its associated equipment. Fuel gas from Acid Gas Absorber, 22-1-V-1, at approximately 105°F is contacted countercurrently with a hot demineralized water stream in the packing of Resaturator Tower, 21-1-V-4. The water saturated effluent fuel gas stream at approximately 300°F then flows to the Gas Cooling Unit, 20-1-ME-3.

Bottoms from the Resaturator Tower are pumped through the recirculation pump 21-1-P-3, combined with demineralized water makeup and heated by exchange with hot gasifier effluent in the Resaturator Water Heater, 21-1-E-1. The heater effluent at approximately 313°F is then recycled to the Resaturator Tower for contact with the clean fuel gas stream.

Equipment Notes

All equipment is commercially available.



MATERIAL BALANCE

COMPONENT	PROCESS STREAM NUMBER							
	5 PARTICULATE SCRUBBER OVHD		6 AMMONIA SCRUBBER OVHD		9 FUEL GAS		10 RESATURATED FUEL GAS	
	MPH	MOL%	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	24.0	0.07	24.0	0.10	23.6	0.10	23.6	0.09
H_2	8658.0	26.37	8658.2	35.12	8632.2	36.38	8,632.4	31.85
CO	12,745.5	38.82	12,745.5	51.71	12,662.9	53.36	12,662.5	46.72
CO_2	2,614.8	7.97	2614.8	10.61	2196.4	9.26	2,196.0	8.10
H_2S	302.3	0.92	302.3	1.23	24.0	0.10	24.0	0.09
COS	18.0	0.06	18.0	0.07	11.6	0.05	11.6	0.04
N_2	199.0	0.61	119.0	0.81	118.1	0.50	118.1	0.44
AR	36.2	0.11	36.2	0.14	36.1	0.15	36.1	0.13
NH_3	59.3	0.18	—	—	—	—	—	—
H_2O VAPOR	8,172.0	24.89	51.70	0.21	22.5	0.10	3,398.7	12.54
TOTAL VAPOR MPH	32,829.1	100.00	24,649.5	100.00	23,727.4	100.00	27,103.0	100.00
H_2O LIQ	—		—		—		—	
TOTAL LB/HR	656,107.5		508,934		475,512.6		536,410	
MOL WT VAPOR	19.99		20.66		20.04		19.79	

NOTES :

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

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PROCESS FLOW DIAGRAM
GAS COOLING
TEXACO PROCESS-OXYGEN BLOWN
EPRI PALO ALTO, CALIFORNIA

SCALE NONE
DRAWING NUMBER EXTC78-21-1
REVISION 03 11-6-79

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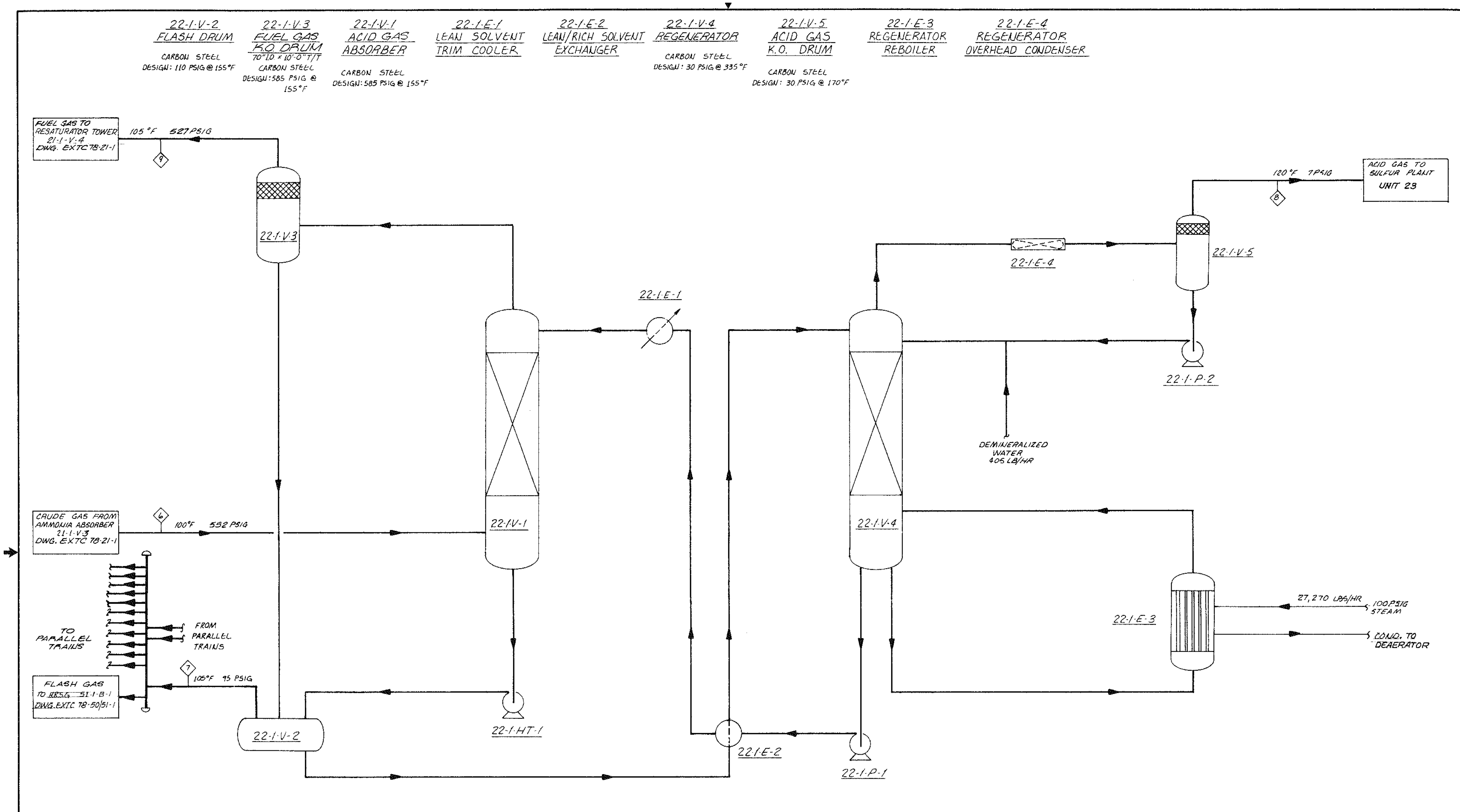
ACID GAS REMOVAL

Process Flow Diagram EXTC78-22-1 depicts one of the three parallel acid gas removal trains. This unit is similar to its counterpart designed for Case EXTC (Slurry Feed) and described in EPRI Report AF-642 (January 1978, page 321). 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.



MATERIAL BALANCE								
	PROCESS STREAM NUMBER							
COMPONENT	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	8658.0	35.12	24.1	12.61	1.7	0.25	8632.2	36.38
CO	12,745.5	51.71	71.9	37.62	10.7	1.59	12,662.9	53.37
CO_2	2614.8	10.61	81.2	42.49	337.2	49.98	2196.4	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	0.81	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	51.70	0.21	1.2	0.63	51.7	7.66	22.5	0.90
TOTAL MPH	24,649.5	100.00	191.1	100.00	674.7	100.00	23,727.4	100.00
TOTAL LB/HR	508,934		6096		26539		475,512.6	
MOL WT	20.66		31.90		39.33		20.04	

NOTES:

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

2. ALL RATES ARE EXPECTED VALUES ON A PER TRAIN BASIS AT FULL CAPACITY OPERATION EXCEPT AS NOTED.

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

EPRI PALO ALTO, CALIFORNIA

SCALE NONE
DRAWING NUMBER EXTC 78-22-1
REVISION 02 11-6-79

SULFUR RECOVERY AND TAIL GAS TREATING

The processes used in these units are the same as for Case MASW. Refer to Case MASW and Process Flow Diagrams MASW-23-1, MASW-23-2 and MASW-23-3. The detailed process descriptions of these units are contained in EPRI Report AF-642 (January 1978, pages 74 through 84).

Equipment Notes

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

PROCESS CONDENSATE TREATING

Most of the sour process condensate generated in the plant is used for the preparation of the coal slurry feed to the gasifiers. A small stream of ammonia (as ammonium salts) contaminated effluent is produced in the gasification area (Flow Diagram: EXTC78-20-1). This effluent is treated in Effluent Water Treating Unit 40.

A unit for the recovery of by-product ammonia is not provided. The system for treatment of process condensate is similar to that provided for Case EXTC (Slurry Feed) as described in EPRI Report AF-642 (January 1978, page 327).

Equipment Notes

All equipment is commercially available.

STEAM, BOILER FEEDWATER, AND CONDENSATE

Process Flow Diagram EXTC78-30-1 schematically represents the steam, boiler feedwater, and condensate systems. This system is significantly different from that presented for Case EXTC (Slurry Feed) in EPRI Report AF-642 and is accordingly described in detail below. One train is provided.

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

High Pressure	- 1,450 psig 900°F
Intermediate Pressure	- 400 psig
Medium Pressure	- 100 psig
Low Pressure	- 50 psig

All high-pressure (HP) steam generation is carried out only in the gasification area gas cooling units, 20-1-ME-3. All the saturated HP steam is superheated to 900°F in the HRSG superheaters (51-B-1:E-2) then used to drive the single back pressure-type power generation turbine, 51-T-1. The HP end of Turbine 51-T-1 takes steam at 1,450 psig, 900°F and exhausts at the intermediate-steam pressure of 445 psig.

Saturated intermediate-pressure (IP) steam at 445 psig is also obtained in the IP steam generators located in the sulfur plant, gasification unit (20-1-ME-3) and the HRSGs (51-B-1:E-4). The saturated IP steam together with the exhaust steam from 51-T-1 is superheated to 900°F in the HRSG's reheaters (51-B-1:E-1). The superheated IP steam at 400 psig, 900°F is then used in the IP end of condensing turbine 51-T-1 and in condensing turbine drivers, 11-1-T-1 and 11-1-T-2.

Saturated medium-pressure (MP) steam at 100 psig is produced in the MP steam generators (20-1-ME-3) in the gasification unit. The primary users of MP steam are the Regenerator Reboilers, 22-1-E-3, located in the acid gas removal units. A small quantity of the saturated MP steam is also used in the sulfur plant (Unit 23). The remainder is combined with 50 psig LP steam and used in condensing turbine 51-T-2.

Saturated 50 psig steam is produced in the HRSG's generators (51-B-1:E-6) and in the steam generators in the sulfur plant. The 50 psig steam is used primarily in condensing turbine 51-T-2. A small amount of the 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 use in a 1500 psig boiler. The demineralized water is stored in Tank 30-TK-1. Demineralized water from the storage tank flows to the deaerator through Pumps 30-P-4A&B. A sidestreams are pumped to Units 21 and 22 as makeup to the fuel gas resaturator and Selexol systems.

The vacuum condensate from turbines 11-1-T-1, 11-1-T-2, 51-T-1 and 51-T-2 is combined and used to recover heat from the crude gasifier effluent in 21-1-E-2. The resulting heated stream then flows to the deaerator. The condensate streams produced by the 100 psig and 50 psig steam users also flow to the deaerator.

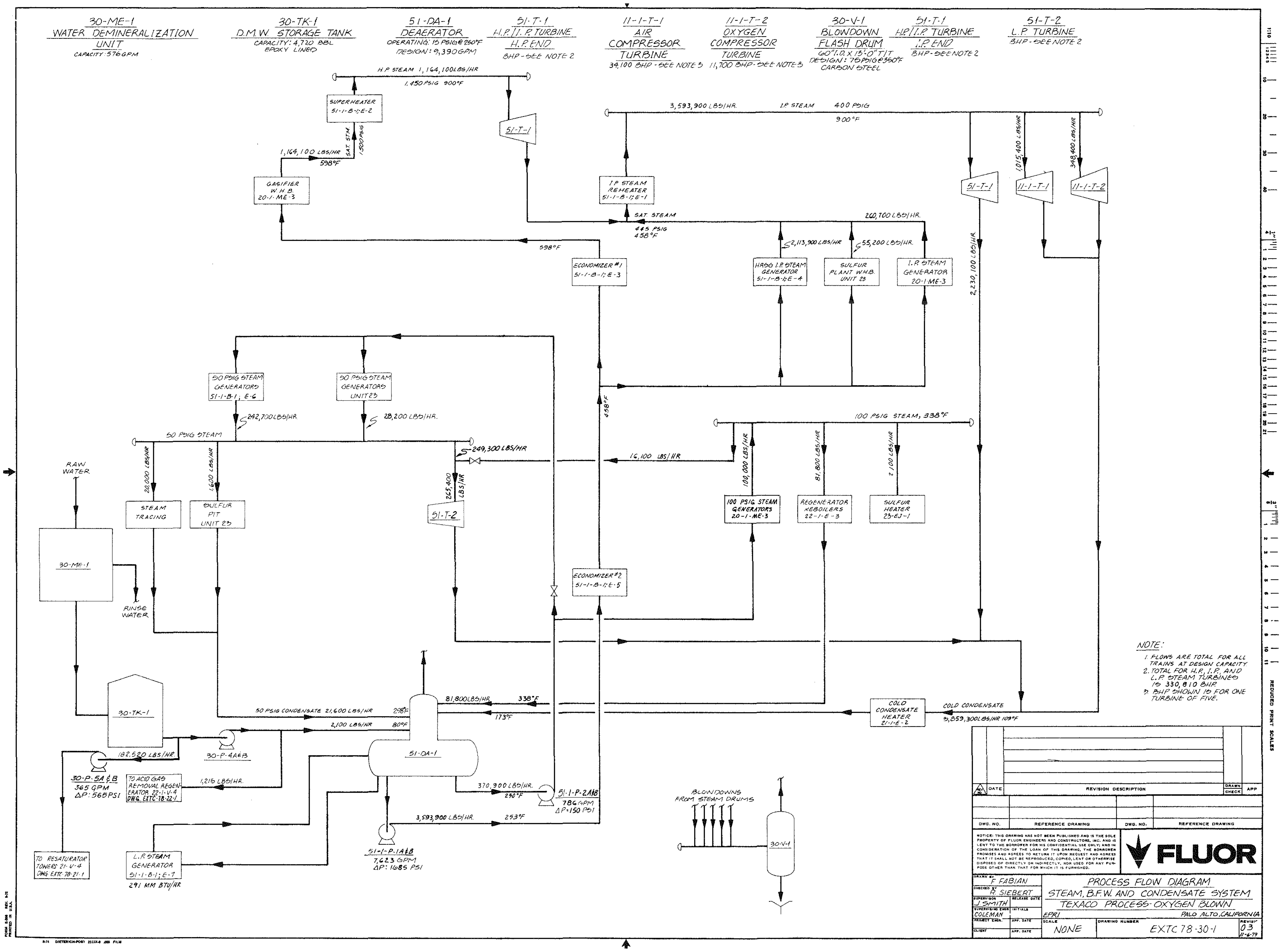
The deaerator is a tray-type unit which provides 10 minute storage. The deaerating steam (15 psig) is generated in the LP steam evaporators 51-B-1:E-7 of the HRSG. The deaerator operates at slightly lower than 15 psig to permit the flow of deaerating steam into the vessel.

Boiler feedwater (BFW) from the deaerator is pumped through the high-pressure boiler feedwater pump (51-P-1) to the gasification unit (20-1-ME-3) for generation of HP and IP steam and to the sulfur plant waste heat boilers and the IP evaporator 51-B-1:E-4 of the HRSG. Both the 50 psig and the 100 psig steam generators are supplied with boiler feedwater by the low pressure BFW pump 51-P-2.

BFW is first heated to the IP steam saturation temperature (458°F) in economizer 51-B-1:E-5. Part of the BFW is withdrawn downstream of 51-B-1:E-5 and supplied to the IP steam generators located in the Unit 23 sulfur plant, gasification unit (20-1-ME-3) and the HRSG (51-B-1:E-4). The balance of the BFW is heated to the HP steam saturation temperature (598°F) in the economizer 51-B-1:E-3 of the HRSG prior to flowing to the HP steam generators in the gasification units.

Equipment Notes

All this equipment is commercially available.


30-ME-1
WATER DEMINERALIZATION UNIT
CAPACITY: 576 GPM
30-TK-1
D.M.W. STORAGE TANK
CAPACITY: 4,720 BBL
EPOXY LINED
51-DA-1
DEAERATOR
OPERATING: 15 PSIG@250°F
DESIGN: 9,390 GPM
51-T-1
H.P./I.P. TURBINE
H.P. END
BHP - SEE NOTE 2
11-1-T-1
AIR COMPRESSOR TURBINE
39,100 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
H.P./I.P. TURBINE
I.P. END
BHP - SEE NOTE 2
51-T-2
L.P. TURBINE
BHP - SEE NOTE 2
H.P. STEAM 1,164,100 LBS/HR
1,450 PSIG 900°F
SUPERHEATER
51-1-B-1; E-2
SAT. STM. 1,500 PSIG
1,164,100 LBS/HR
598°F
GASIFIER W.H.B.
20-1-ME-3
51-T-1
3,593,900 LBS/HR
I.P. STEAM 400 PSIG
900°F
I.P. STEAM REHEATER
51-1-B-1; E-1
SAT. STEAM
445 PSIG
458°F
200,700 LBS/HR
598°F
ECONOMIZER #1
51-1-B-1; E-3
2,113,900 LBS/HR
55,200 LBS/HR
HRSG I.P. STEAM GENERATOR
51-1-B-1; E-4
SULFUR PLANT W.H.B.
UNIT 23
I.P. STEAM GENERATOR
20-1-ME-3
1,015,400 LBS/HR
348,400 LBS/HR
51-T-1
11-1-T-1
11-1-T-2
2,230,100 LBS/HR
50 PSIG STEAM GENERATORS
51-1-B-1; E-6
50 PSIG STEAM GENERATORS
UNIT 23
242,700 LBS/HR
28,200 LBS/HR
50 PSIG STEAM
249,300 LBS/HR
458°F
16,100 LBS/HR
100 PSIG STEAM, 338°F
RAW WATER
20,000 LBS/HR
1,600 LBS/HR
265,400 LBS/HR
100,000 LBS/HR
81,800 LBS/HR
2,100 LBS/HR
STEAM TRACING
SULFUR PIT
UNIT 23
51-T-2
100 PSIG STEAM GENERATORS
20-1-ME-3
REGENERATOR REBOILERS
22-1-E-3
SULFUR HEATER
23-EJ-1
30-ME-1
RINSE WATER
ECONOMIZER #2
51-1-B-1; E-5
NOTE:
1. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
2. TOTAL FOR H.P., I.P. AND L.P. STEAM TURBINES IS 330,810 BHP
3. BHP SHOWN IS FOR ONE TURBINE OF FIVE.
30-TK-1
50 PSIG CONDENSATE 21,600 LBS/HR
298°F
2,100 LBS/HR
80°F
81,800 LBS/HR
338°F
173°F
COLD CONDENSATE HEATER
21-1-E-2
COLD CONDENSATE
3,859,300 LBS/HR 109°F
182,520 LBS/HR
30-P-4A&B
30-P-5A & B
365 GPM
ΔP: 565 PSI
TO ACID GAS REMOVAL REGENERATOR 22-1-V-4
DWG. EXTC-78-22-1
1,215 LBS/HR
51-DA-1
370,900 LBS/HR
250°F
51-1-P-2A&B
786 GPM
ΔP=150 PSI
3,593,900 LBS/HR
253°F
BLOWDOWNS FROM STEAM DRUMS
30-V-1
TO RESATURATOR TOWERS 21-V-4
DWG. EXTC-78-21-1
L.P. STEAM GENERATOR
51-1-B-1; E-7
291 MM BTU/HR
51-1-P-1A&B
7,623 GPM
ΔP: 1685 PSI
DATE
REVISION DESCRIPTION
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REFERENCE DRAWING
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DRAWN BY F. FABIAN
CHECKED R. SIEBERT
J. SMITH
COLEMAN
PROCESS FLOW DIAGRAM
STEAM, B.F.W. AND CONDENSATE SYSTEM
TEXACO PROCESS - OXYGEN BLOWN
EPRI
PALO ALTO, CALIFORNIA
SCALE NONE
DRAWING NUMBER EXTC 78-30-1
REVISION 03

COMBINED-CYCLE SYSTEM

Process Flow Diagram EXTC78-50/51-1 depicts the combined-cycle system for Case EXTC78. This system is significantly different from the combined cycle presented for Case EXTC (Slurry Feed) in EPRI Report AF-642. The Case EXTC (Slurry Feed) combined cycle employed advanced gas turbines fired at 2400°F. These advanced, 2400°F gas turbines are expected to be state-of-the-art machines for delivery after 1985. In contrast, the Case EXTC78 combined cycle is based upon 1978 state-of-the-art gas turbines with a 1980°F combustor exit temperature.

The combined-cycle system for Case EXTC78 consists of twelve parallel trains of gas turbines. Each train contains two gas turbines, 50-1-GT-1A&B, coupled to a single generator, 50-1-G-1, and exhausting into a single HRSG, 51-1-B-1. Steam produced in the HRSGs of the twelve gas turbine trains flows to a single steam turbine train. This steam turbine train consists of one HP/IP steam turbine (51-T-1) driving generator 51-G-1. A fuel gas expander (50-EX-1) and generator unit (50-G-2) are also provided. Detailed performance information on the power block components, i.e., gas turbines, HRSGs and the steam turbine is provided in Appendix B. The combined-cycle system process flowplan for Case EXTC78 is discussed below.

Resaturated fuel gas from the gasification units (20-1-ME-3) at 514 psig, 1000°F is expanded to 228 psig in expander 50-EX-1 to generate power in generator unit 50-1-G-2. The fuel gas is then combusted in the conventional 1980°F gas turbine 50-1-GT-1. The exhaust gas from 50-1-GT-1 then combines with flash gas from Unit 22 and enters the HRSG 51-1-B-1 at a temperature of 948°F.

The HRSG generates saturated steam at three pressure levels; intermediate-pressure (IP) at 445 psig, low-pressure (LP) at 50 psig and low-pressure at 15 psig. The 15 psig steam generated in the HRSG is used as deaerating steam in Deaerator 51-DA-1.

The arrangement of heat recovery sections of the HRSG in the direction of gas flow is as follows:

Reheater and Superheater	51-B-1:E-1 and 51-B-1:E-2
Economizer One	51-B-1:E-3
IP Evaporator	51-B-1:E-4
Economizer Two	51-B-1:E-5
LP (50 psig) Evaporator	51-B-1:E-6
Deaerator LP (15 psig) Evaporator	51-B-1:E-7

High-pressure saturated steam from the gasification unit is heated to 900°F in the superheater 51-1-B-1:E-2 and flows to the HP end of the back pressure turbine 51-T-1.

Saturated steam produced in the IP evaporator is combined with both intermediate-pressure steam from the process generators and cold reheat steam from the high-pressure steam turbine, 51-T-1, and is reheated to 900°F in the 51-1-B-1:E-1. This reheated IP steam is then used in condensing driver turbines 11-1-T-1 and 11-1-T-2 and in the IP end of condensing power turbines 51-T-1.

The 50 psig saturated steam produced in LP evaporator 51-1-B-1:E-6 combines with excess 100 psig steam from the process units and is used in condensing turbine 51-T-2 to drive generator unit 51-G-2.

LP evaporators 51-1-B-1:E-7 located in the topmost section of the HRSG's supply deaerating steam to the tray type deaerator, 51-DA-1, at a pressure of 15 psig. One common deaerator is provided for the multiple HRSGs and the process steam generators.

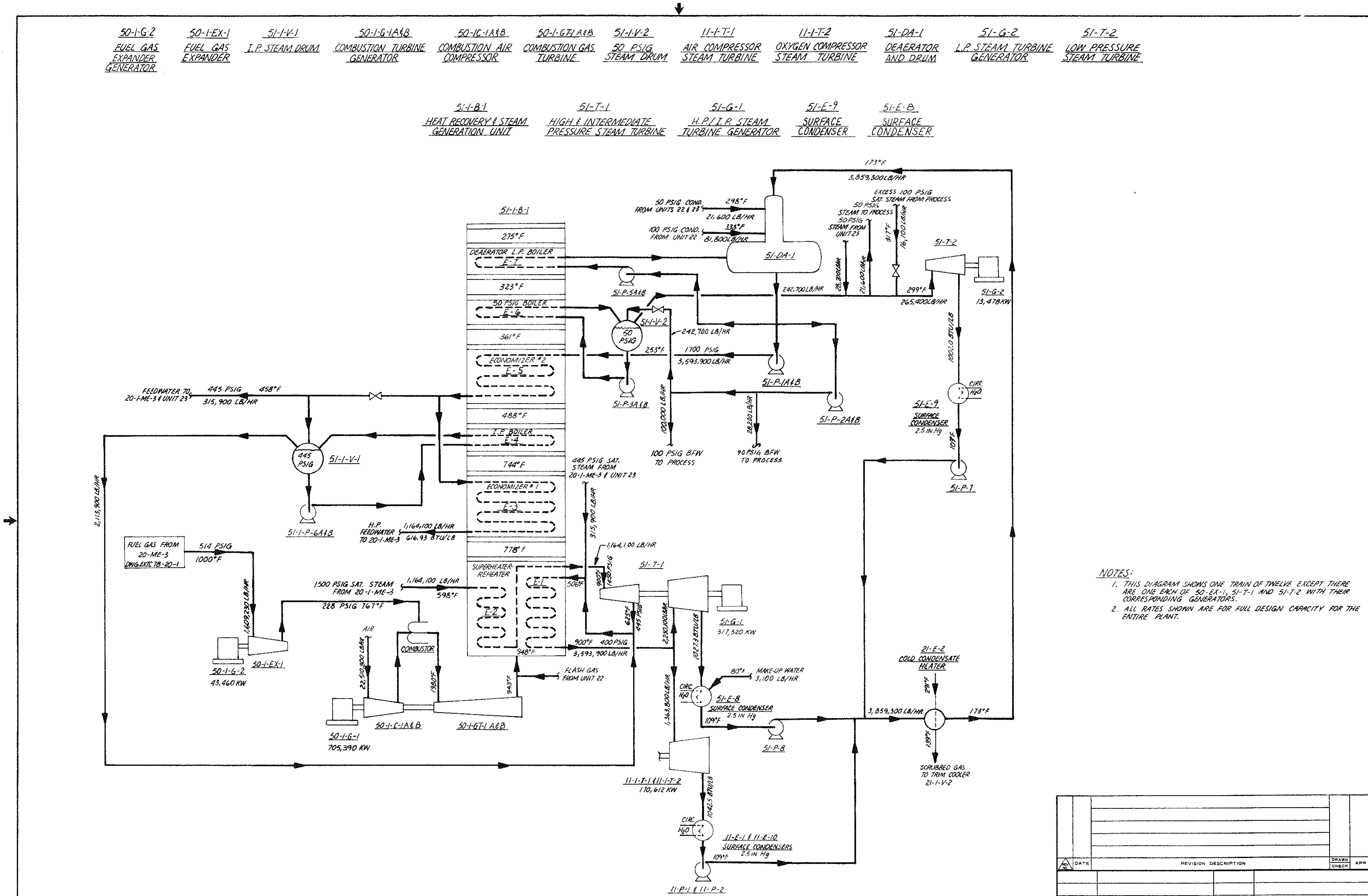
The boiler feedwater (BFW) from the deaerator is first preheated to the IP steam saturation temperature (458°F) in Economizer Two 51-1-B-1:E-5. A portion of the BFW flows to IP evaporator 51-1-B-1:E-4 and to the process IP steam generators. The balance of the BFW is further preheated to the HP steam saturation temperature (598°F) in Economizer One 51-1-B-1:E-3 and then flows to the gasification unit waste heat boilers for the generation of HP steam.

Each HRSG is provided with its own LP and IP steam drums and corresponding BFW circulation pumps.

The HRSG exhaust (stack) gas temperature of 275°F, established in conjunction with the low-pressure (15 psig) evaporator section, allows the gas side surface of the LP evaporator to operate safely above the dew point for acidic SO_2-bearing mist in the exhaust gas.

Equipment Notes

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



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CHECKED BY L. ACEVEDO
SUPERVISOR J. SMITH
SUPERVISING ENGR B. COLEMAN

PROCESS FLOW DIAGRAM
STEAM & POWER GENERATION
TEXACO PROCESS-OXYGEN BLOWN

EPRI — PALO ALTO, CALIF.

SCALE: NONE — DRAWING NUMBER: EXTC 78-50/51-1 — REVISION: 03 11-6-79

PROCESS DISCUSSION - CASE EXTC78

The table below summarizes pertinent heat and material balance results.

Table 7-2

SUMMARY OF SYSTEM PERFORMANCE - CASE EXTC78

GASIFICATION AND GAS CLEANING SYSTEM	
Coal Feed Rate, lbs/hr (m.f.)	798,333
Oxygen*/Coal Ratio, lbs/lb m.f.	0.858
Oxidant Temperature, °F	300
Slurry Water/Coal Ratio, lbs/lb m.f.	0.503
Gasification Section Average Pressure, psig	600
Crude Gas Temperature, °F	2,300-2,600
Crude Gas HHV (dry basis), Btu/SCF**	281.1
Temperature of Fuel Gas to Gas Turbine, °F	767
POWER SYSTEM	
Gas Turbine Combustor Exit Temperature, °F	1,980
Pressure Ratio	13.35:1
Turbine Exhaust Temperature, °F	943
Steam Conditions, psig/°F/°F	1,450/900/900
Condenser Pressure, Inches Hg abs	2.5
Stack Temperature, °F	275
Gas Turbine Power#, MW	706
Steam Turbine Power#, MW	331
Fuel Gas Expander Power#, MW	43
Power Consumed, MW	41
Net System Power, MW	1,039
OVERALL SYSTEM	
Process and Deaerator Makeup Water, gpm/1000 MW##	804
Cooling Tower Makeup Water, gpm/1000 MW	8,100
Cooling Water Circulation Rate, gpm/MW	374
Cooling Tower Heat Rejection, % of Coal HHV	38.1
Air Cooler Heat Rejection, % of Coal HHV	5.2
Net Heat Rate, Btu/kWh	9,813
Overall System Efficiency (Coal → Power), % of Coal HHV	34.76

*Dry basis, 100 percent O_2
**Excluding the HHV of H_2S, COS and NH_3
#At Generator terminals
##Includes make-up water for resaturator tower

GASIFIER MATERIAL BALANCE

Gasifier material balances for full capacity operation are given in Table 7-3 for the oxygen-blown Texaco gasifier.

Most of the data presented in the above table was received from Texaco Development Corporation for an earlier study published as EPRI report AF-642. That study projected economics and operating data for the 1980-1985 time frame. For the particular coal used, Texaco indicated that slurry concentrations in the range of 60 percent solids to possibly 70 percent solids could be achieved in the 1980-1985 time frame. For study purposes, EPRI selected a slurry concentration of 66.5 percent 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 possess different slurrying characteristics. The material balance, including oxygen consumption, is based on a Texaco extrapolation of the state of the art to a period three to five years hence.

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 salts 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.

The figures given in the tables and flow sheets for ammonia should be regarded as tentative estimates only.

The gasifier temperatures are believed to be high enough to destroy all hydrocarbon except methane.

Table 7-3

MATERIAL BALANCE - CASE EXTC78

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		

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 rejection 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. This process is used in these cases because it accomplishes this objective and exhibits favorable economics when compared with other similar processes.

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 only.

PROCESS ENERGY BALANCES

Table 7-4 presents an overall process energy balance at 100 percent capacity operation. The boundary for each balance encompasses the entire plant, exclusive of the cooling tower heat balance. 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. The energy balance closes to less than one-half of one percent. The discrepancy results from approximations used for some process units and for calculating some heat loads.

Data from Table 7-4 is shown in MM Btu/hr and as percent of coal higher heating value in Table 7-5. Coal charged at 10,000 ton/day is equivalent to 10,196 MM Btu/hr HHV. This feed produces 3544 MM Btu/hr power equivalent or 34.8 percent of the coal HHV as net electric power. The heat rate based on net power produced is 9813 Btu/kWh. Heat rejected at all steam turbine surface condensers is 3699 MM Btu/hr or 36.3 percent of the coal HHV. Heat rejected with the HRSG stack gases is 2404 MM Btu/hr or 23.6 percent of the coal HHV.

Table 7-4

ENERGY BALANCE - CASE EXTC78

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		153	379			532
Demineralized and Raw Water		4				4
Auxiliary Power Inputs					140	140
TOTAL	10,196	184	432	0	140	10,952
HEAT OUT						
Ash Cake		6				6
Gasifier Heat Losses				26		26
Gas Cooling		19	6			25
Sulfur Product	105	1				106
Oxidant Compressor Inter/After Cooling		535	33			568
Oxidant Compressor Surface Condensers			1,317			1,317
Gas Turbines and Expanders					2,555	2,555
Sulfur Plant Effluent Gas		2	19			21
Steam Turbines					1,129	1,129
Power Block Losses*				26	56	82
Steam Turbine Condenser			2,382			2,382
HRSG Stack Gas		1,332	1,072			2,404
Steam Heat Losses			25			25
Motor Losses (Air Cooler Fans, etc.)					89	89
Selexol Overhead Condenser			24			24
Selexol Solvent Cooler		54				54
Process Water Cooling		95				95
Air Separation Plant Waste Gas		31	20			51
Waste Water Effluent		13				13
TOTAL	105	2,088	4,898	52	3,829	10,972

$$\frac{\text{Output} - \text{Input}}{\text{Input}} = 0.18\%$$

*Includes mechanical and electrical losses

Table 7-5

ENERGY BALANCE AS PERCENT COAL HHV - CASE EXTC78

	MM Btu/hr	Percent
IN		
Coal HHV	10,196	100.0
OUT		
Net Power	3,544	34.76
Sulfur Product, HHV and Sensible	106	1.04
Selexol Sensible and Latent	78	0.77
Oxidant Inter/After Cooling	568	5.57
Ash Cake Sensible	6	0.06
HRSG Stack Gases	2,404	23.57
Rejected at Condensers	3,699	36.28
Other Sensible Losses, Net	28	0.28
Other Latent Losses, Net	(362)	(3.55)
Gasifier Heat Losses	26	0.25
Motor Losses	89	0.87
Power Block Losses	82	0.80
	10,268	100.70

ECONOMICS - CASE EXTC78

Table 7-6 summarizes the economics of Case EXTC78.

Table 7-6

SUMMARY OF ECONOMIC RESULTS - CASE EXTC78

PRODUCTION AT DESIGN CAPACITY	
Net Power, MW*	1,039
Overall Plant Heat Rate, Btu/kWh	9,813
TOTAL CAPITAL REQUIREMENT**	
Total Capital Requirement, $1,000	898,425
Total Capital Requirement, $/kW	864.70
AVERAGE COSTS OF SERVICES**	
First Year Cost, $1,000/year	247,629
First Year Cost, mills/kWh	38.87
Thirty Year Levelized Cost, $1,000/year	326,626
Thirty Year Levelized Cost, mills/kWh	51.28

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

TOTAL PLANT INVESTMENT

Table 7-7 gives a detailed breakdown of the Total Plant Investment required for Case EXTC78. The accuracy of the plant investment estimate is judged to be ±25 percent. Since other capital charges are keyed to elements of plant investment, this accuracy is reflected in the computed Total Capital Requirement and Cost of Services.

Table 7-7

TOTAL PLANT INVESTMENT - CASE EXTC78

Plant Section	Cost Breakdown Without Contingencies: Direct Field Material#	Direct Field Labor##	Engr. & Support Costs§	Total Sales Tax	Total Cost $1,000*	Cost $/kW**	Per-cent	Contingencies: Process $1,000*	Project $1,000*	Total Plant Investment: $1,000*	$/kW**
Coal Handling	10,499	4,435	7,127	530	22,591	21.74	3.41	-	3,389	25,980	25.00
Oxidant Feed	62,828	23,613	30,991	1,968	119,400	114.92	18.00	-	17,910	137,310	132.16
Gasification and Ash Handling	14,288	3,456	6,622	721	25,087	24.14	3.78	3,460	3,763	32,310	31.10
Gas Cooling	26,735	12,615	19,952	1,386	60,688	58.41	9.15	3,730	9,103	73,521	70.76
Acid Gas Removal and Sulfur Recovery	14,711	5,120	8,754	752	29,337	28.24	4.42	668	4,400	34,405	33.11
Steam, Condensate and B&W	966	400	611	44	2,021	1.95	0.31	-	303	2,324	2.24
Combined Cycle	179,640	54,998	98,318	9,257	342,213	329.37	51.59	-	51,332	393,545	378.77
General Facilities	31,510	12,603	16,812	1,041	61,966	59.64	9.34	-	9,295	71,261	68.59
Subtotal	341,177	117,240	189,187	15,699	663,303	638.41	100.00	7,858	99,495	770,656	741.73

TOTAL PLANT INVESTMENT SUMMARY

	$1,000*	$/kW**
Process Plant Investment and General Facilities	663,303	638.41
Process Contingency	7,858	7.56
Project Contingency	99,495	95.76
Total Plant Investment	770,656	741.73

*Mid-1976 dollars
**Based on 100 percent plant design power output (1039 MW)
#All materials and equipment that become a part of the plant facility
##Labor cost for installing direct field materials (exclusive of payroll burdens and craft benefits)
§Includes: a) Indirect field costs including all labor, supervision and expenses required to support field construction; b) Home office costs including all salaries and expenses required for engineering design and procurement; and c) Contractor's fee

However, due to the similar nature of the cases in this evaluation, the estimates for all cases should reflect about the same accuracy. When these estimates are used comparatively, the effect of individual accuracies should be minimal.

Two contingencies are included in the Total Plant Investment shown for each plant section. First is a 15 percent project contingency which is intended to cover additional equipment and material 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 innovative technology in an effort to quantify the uncertainty in the design, performance, and cost of the commercial scale equipment. This contingency covers the additional expenditures required to solve any problems associated with innovative technology, since historically such technology requires more cost than initially estimated. Although all plant technology is judged to be currently commercially available, not all of it is fully proven by commercial operation. Accordingly, a process contingency is applied and included in the estimated plant investment for sections containing innovative technology.

TOTAL CAPITAL REQUIREMENT

Table 7-8 gives a breakdown of the Total Capital Requirement to place the plant into initial operation. Starting with the total plant investment, capital charges are added for allowance for funds during construction, initial catalyst and chemicals charge, prepaid royalties, preproduction costs, the startup inventory capital, and the cost of the land required for the plant. Specific items included in each of these capital charges are described under the economic criteria.

COST OF SERVICES

Table 7-9 gives a cost of services breakdown for Case EXTC78. The costs are busbar power costs based on plant operation at a 70 percent capacity factor with a $1/MM Btu coal cost. Table 7-9 shows both first-year and 30-year levelized power costs. Since the power costs vary with the plant capacity factor, this additional relationship is shown in Figure 7-1 on both a first-year and 30-year levelized basis. The plant capacity factor presented in Figure 7-1 is the percent plant time on-line at baseload (100 percent) plant power output of 1038.8 MW. This capacity factor is not equivalent to plant operation in a turndown mode producing less than baseload power output, as the design assumes a baseload plant.

Table 7-8

TOTAL CAPITAL REQUIREMENT - CASE EXTC78

	$1000*	$/KW**
TOTAL PLANT INVESTMENT	770,656	741.73
CAPITAL CHARGES		
Prepaid Royalties	3,007	2.89
Preproduction Costs	19,317	18.59
Inventory Capital	7,476	7.20
Initial Catalyst and Chemicals Charge	714	0.69
Allowance for Funds During Construction	96,255	92.64
Land	1,000	0.96
Total Capital Charges	127,769	122.97
TOTAL CAPITAL REQUIREMENT	898,425	864.70

*Mid-1976 dollars

**Based on 100 percent plant design power output

The largest single component of the cost of power is contained in the levelized fixed charges. These charges represent repayment from operating revenue of the financing originally used to supply the total capital required for construction and startup of the plant. These levelized fixed charges amount to 65 percent of the "first-year" power costs at a 70 percent capacity factor. The charges shown are fixed annual expenses which are independent of the plant capacity factor.

Coal constitutes the major operating charge. This charge amounts to 25 percent of the "first-year" power costs at a 70 percent capacity factor.

Operating labor requirements are a function of the number of units and trains. These labor requirements we shown below on a per-shift basis.

• "A" Operators	5
• "B" Operators	17
• Foremen	2
• Lab and Instrument Technicians	4

Table 7-9

BUSBAR POWER COST AT 70 PERCENT CAPACITY FACTOR - CASE EXTC78

	O_2- Blown Texaco EXTC78	
NET PRODUCTION*		
Net Power, MW	1,039	
By-product Ammonia, ST/D	-0-	
By-product Sulfur, ST/D	301	
TOTAL CAPITAL REQUIREMENT, $1,000/YEAR	898,425	
	First-Year Cost	**30-Year Levelized Cost**
FIXED OPERATING COST, $1,000/YEAR		
Operating Labor	3,066	5,783
Maintenance Labor	5,995	11,307
Maintenance Materials	8,992	16,959
Administrative & Support Labor	2,718	5,127
Total Fixed O&M Costs	20,771	39,176
VARIABLE OPERATING COSTS (EXCLUDING COAL), $1,000/YEAR		
Raw Water	1,361	2,568
Catalysts & Chemicals	279	526
Other Consumables (if any)	-0-	-0-
Ash Disposal	981	1,850
Variable Maintenance (if any)	-0-	-0-
Total Variable O&M Costs	2,621	4,944
COAL COST, $1,000/YEAR	62,521	120,790
BY-PRODUCT CREDITS, $1,000/YEAR		
By-product Ammonia	-0-	-0-
By-product Sulfur	-0-	-0-
Total Byproduct Credits	-0-	-0-
TOTAL OPERATING COSTS, $1,000/YEAR	85,913	164,910
LEVELIZED FIXED CHARGES, $1,000/YEAR	161,716	161,716
TOTAL COST OF ELECTRICITY**		
$1,000/year	247,629	326,626
mills/kWh	38.87	51.28

*At 100 percent plant design power output

**Mid-1976 dollars and 70 percent plant capacity factor, $1/MM Btu coal

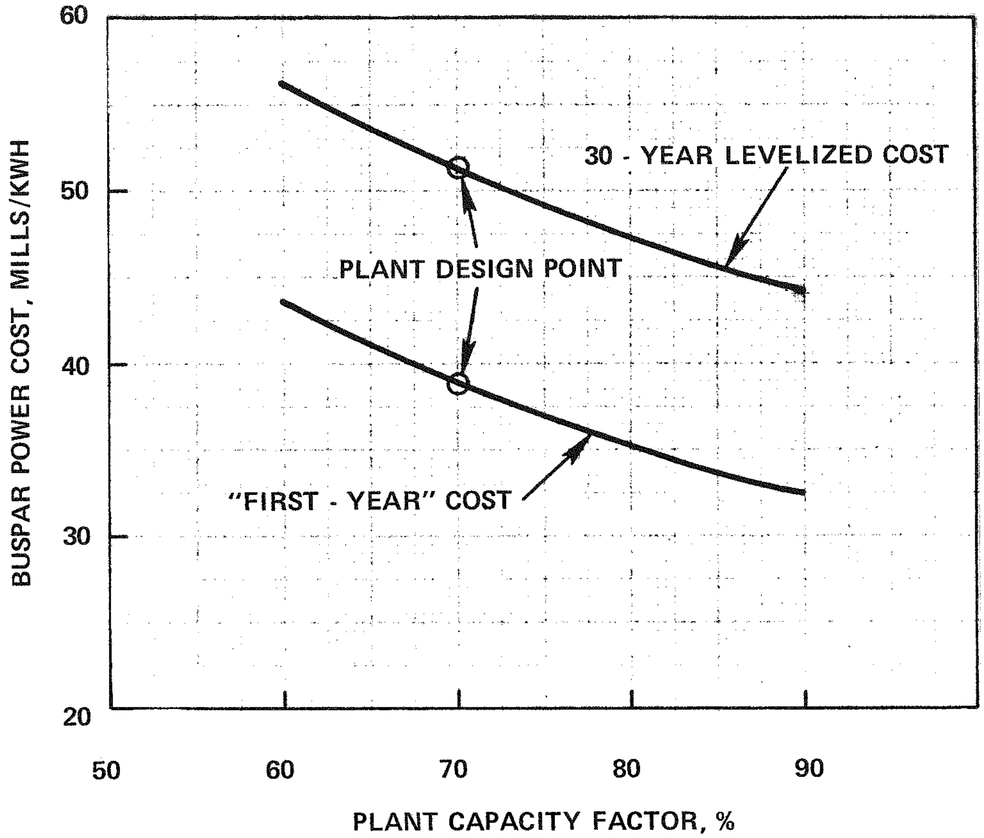

Figure 7-1. Power Cost Sensitivity Curve for Case EXTC78

Maintenance labor and materials costs are calculated for each section as a percentage of the total plant investment for that section. These percentages are listed in the Economic Criteria section.

Catalyst and chemical costs are for chemicals consumed in the demineralized, cooling tower and boiler feedwater treating, plus costs associated with making up Selexol solvent losses in the acid gas removal, and tail gas treating units, and replacement of catalyst in the sulfur recovery unit.

Appendix A

PRESSURIZED BOILER COMBINED-CYCLE SYSTEM DETAILS

GENERAL

The designs of the pressurized boiler combined-cycle systems for Cases MASW, EXTS, and EATS were supplied by STEAG, a subsidiary of RUHRKOHLE AG (West Germany). The STEAG designs are based on interface conditions for heat integration between the fuel processing area and combined-cycle area as provided by Fluor.

In the STEAG combined-cycle design fuel gas from the gasification plant is combusted at 150-200 psig in a specially designed pressurized boiler. Steam is generated in the boiler at pressures as high as 3100 psig to drive a steam turbogenerator which produces the major portion of the power from the cycle.

The flue gas from the pressurized boiler is expanded through a gas turbine in the STEAG combined-cycle design. Thus, the pressurized boiler acts as both a combustor for the gas turbine and a steam generator. The gas turbine is coupled to both an electric generator and to an air compressor which provides combustion air to the pressurized boiler. Both the gas turbine and the air compressor are axial flow machines within a single frame. The hot exhaust gases from the gas turbine are then used to preheat boiler feedwater in an exhaust heat recovery device.

The STEAG pressurized boiler combined cycle is fundamentally different from typical U.S. combined-cycle systems. In the U.S. system, the prime cycle is commonly provided by a combustion gas turbine which drives an electric generator. The bottoming cycle operates by using the waste heat in the gas turbine exhaust gas to generate steam, which is then used to drive a steam turbogenerator. The United Technologies design presented in Appendix B of this report is an example of a typical U.S. combined-cycle system.

The alternate combined-cycle configurations described above are diagrammed schematically in Figure S-1 in the Summary. Some potentially advantageous aspects of the STEAG pressurized boiler combined-cycle designs used in this study are discussed below.

First, the pressurized boiler exhaust gas at less than 1800°F enters the gas turbine. This low temperature allows standard turbine blading materials to be used. Thus, development of high-temperature gas turbines is not required for the STEAG pressurized boiler combined cycle.

Second, the pressurized boiler design may offer greater potential for successfully burning a wide range low HHV gases or heavy liquid fuels than do the U.S. "can"-type gas turbine combustors. Both low HHV gases and heavy liquid fuels have been burned in the STEAG pressurized boiler system at Lünen, West Germany. Low HHV fuel gases generated in air-blown gasifiers are particularly suitable for the pressurized boiler. The large internal volume of the boiler relative to a combustion can and special burner design are capable of supporting stable combustion with fuel gases derived from air blown gasifiers as has been proven in plant operations at Lünen.

Third, the pressurized boiler combined cycle is the only combined cycle that has been integrated with coal gasification in commercial scale operations. STEAG has operated a 170 MW plant at Lünen, West Germany since 1973, based on air-blown Lurgi moving-bed coal gasifiers and the pressurized boiler combined cycle. Heavy liquids have also been successfully utilized as an alternate fuel source. This plant has accumulated over 8600 hours of operation. The STEAG combined-cycle plant with the pressurized boilers has reportedly operated reliably and demonstrated good operational behavior in integrated operation with the coal gasification plant. In general, the pressurized boiler combined-cycle design utilizes proven equipment for the power block. This equipment is considered to be currently commercially obtainable. STEAG has used standard equipment models for the cases presented in this report, although in some of the cases, slight modifications of past equipment designs may be necessary. However, STEAG has performed design studies of the power cycle equipment required for integration of the STEAG cycle with Lurgi gasifiers, so that the equipment designs needed for this integration have been defined.

Fourth, the STEAG cycle equipment is inherently compact. The pressurized boiler arrangement requires less plot space than some other types of power generating cycles.

A summary of the calculated power output for the STEAG power block equipment and of the heat rejected to the station cooling tower is presented in Table A-1.

The basic differences between the STEAG pressurized boiler combined cycle and typical U.S. combined cycles result in much larger percentages of the total power output being generated by the steam turbines. This heavy emphasis on steam-cycle-derived power results in larger heat rejection to the cooling tower from the steam turbine surface condensers.

TECHNICAL INPUT DATA

Steam Conditions

The high-pressure steam conditions used for the STEAG cycle are listed below:

Pressurized Boiler Outlet:	2814 psig 986°F superheat
Turbine Throttle:	2669 psig (2813 psig in Case EATS) 977°F superheat (995°F in Case EATS) 977°F reheat
Condenser:	2.5" Hg abs.

The high-pressure steam conditions are typical for STEAG commercial power plants used in West Germany. However, the pressure level is higher than that used in the pressurized boilers in the 170 MW STEAG GCC demonstration plant at Lünen, West Germany. Steam turbine throttle conditions for the Lünen plant were approximately 1800 psig and 968°F. Although the pressurized boiler used at Lünen did not include a reheat coil for intermediate-pressure steam, a design which would include this feature plus the higher steam generation pressure shown above are considered to be within the capabilities of current technology. To compensate for the foregoing differences, the STEAG combined-cycle designs in this study bear a small process contingency of five percent.

Pressurized Boiler

Balcke-Dürr designed and constructed the pressurized boilers used in the STEAG plant at Lünen, West Germany. This study is based on the Balcke-Dürr type design.

Gas Turbine

The design uses a 1978 state-of-the-art Kraftwerk Union (KWU) V94 standard single-shaft turbine with one extra stage to obtain a higher pressure ratio. The gas turbine air compression ratios are listed in Table A-4. The design data shown in Table A-4 is considered to be well within current gas turbine technology.

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. Heat integration between the process units and the power block is considered wherever possible for the maximum utilization of energy.

POWER BLOCK

Pressurized Boiler (50-1-B-1)

The Balcke-Dürr pressurized boiler is illustrated in Figure A-1. Pertinent performance data is listed in Table A-2. Two boilers are used for each gas turbine, each providing 50 percent of the design capacity. Two pressurized boilers are arranged on opposite sides of each gas turbine and are connected to it through coaxial air and flue gas nozzles. Each boiler is contained in a vertical cylindrical pressure vessel.

The pressurized boilers are fired from multiple gas burners at the top of the vessel. Oil burners are provided at the top of the vessel for startup. The combustion gases flow vertically downward through the boiler and exit to the gas turbine from a horizontal nozzle near the bottom of the vessel.

The arrangement of heat recovery sections in the direction of gas flow is as follows:

Evaporator	50-1-B-1:E-1
Superheater	50-1-B-1:E-2
Reheater	50-1-B-1:E-3

Preheated boiler feedwater enters at the bottom of the boiler vessel and flows to the evaporator header.

The boiler feedwater is evaporated to dryness in the evaporator tubes and exits to a moisture separator located outside the boiler vessel. The saturated HP steam combines with HP steam from the gasification area (in Case EXTS only) and the combined flow is then superheated to 997°F (525°C) in the steam superheat coil located inside the vessel (for Case EATS the superheat temperature is 995°F). The superheated HP steam then flows to the main power turbine. IP turbine exhaust from the power turbine returns to the pressurized boiler where it is reheated to 977°F (525°C) in the reheat coil. Temperature control is provided by one water injection point ahead of the HP Superheater 50-1-B-1:E-2 and one water injection point ahead of the Reheater 50-1-B-1:E-3. These water injection points are both located outside the boiler shell.

All internal heating surfaces are tube coils. The shell of the combustion chamber consists of steam generation tubes welded together to form gas-tight panels,

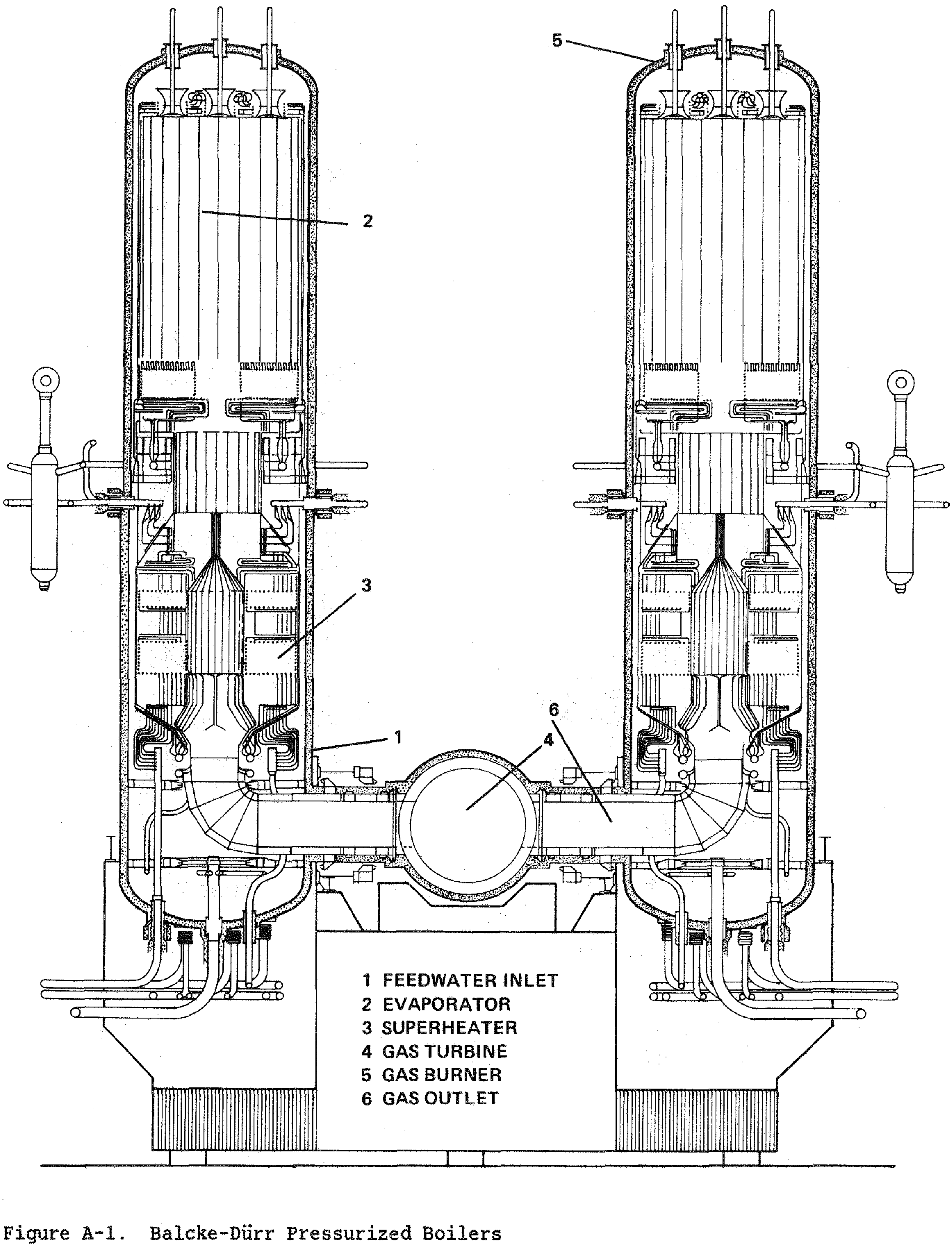

Figure A-1. Balcke-Dürr Pressurized Boilers

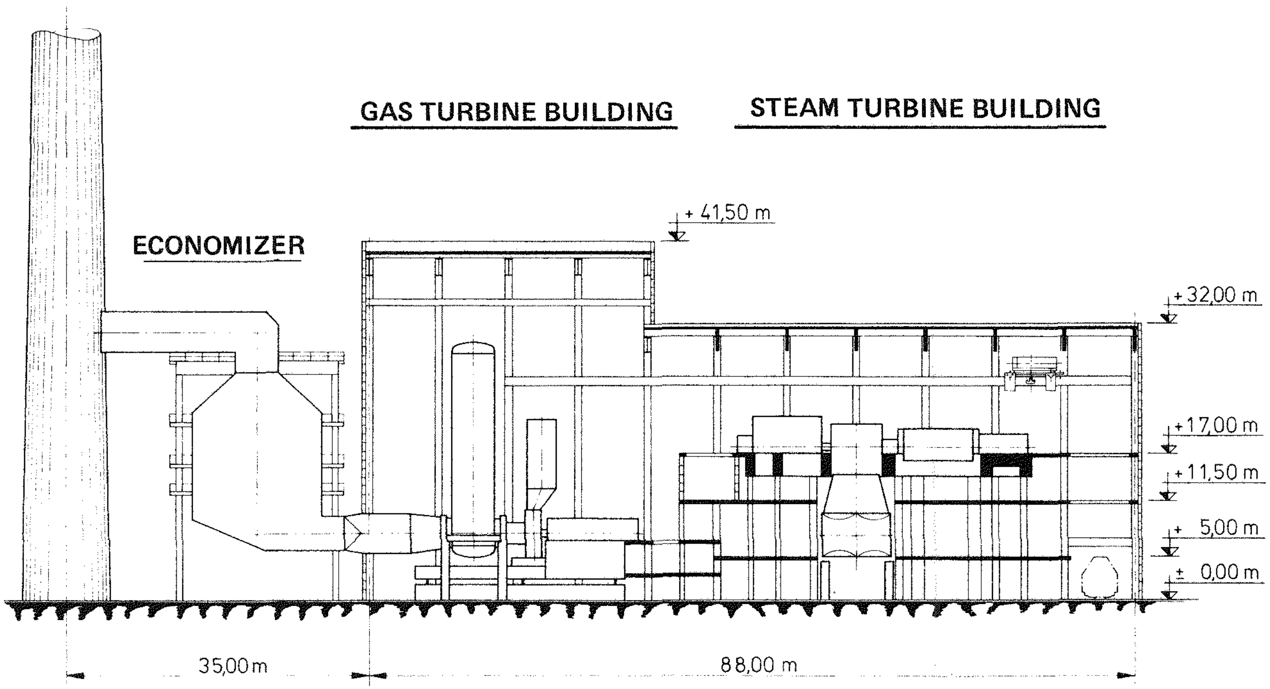

Figure A-2. STEAG Pressurized Boiler Combined Cycle Layout - Side Elevation

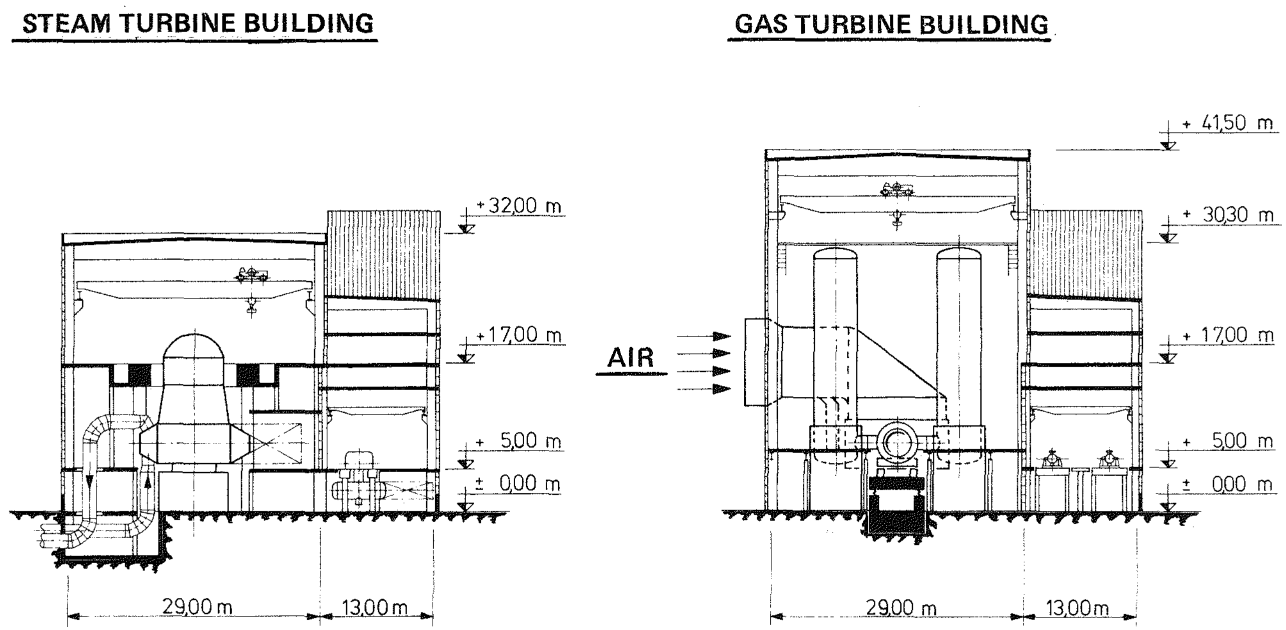

Figure A-3. STEAG Pressurized Boiler Combined Cycle Layout - End Elevation

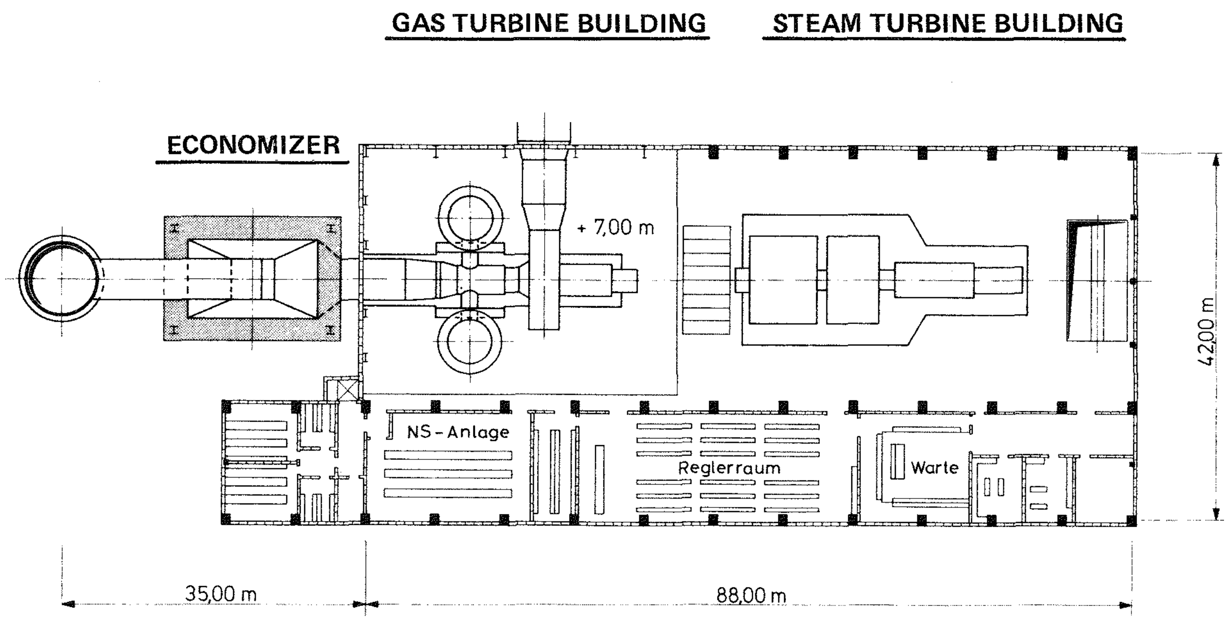

Figure A-4. STEAG Pressurized Boiler Combined Cycle Layout - Plan View which are then hydraulically tested to hold internal pressure. Since the combustion chamber is gas-tight panels of coils, no internal refractory is necessary to protect the outer carbon steel pressure vessel. The pressurized boilers are shop-fabricated devices and require only field erection.

The pressurized boiler described above derives from the compact, high velocity Velox type boilers originally developed in the 1930's for operation at pressures slightly above atmospheric. By 1965, over 100 Velox type boiler plants had been constructed, primarily in Europe. The Balcke-Dürr pressurized boiler is similar to a Velox type boiler, but operates at a higher gas pressure.

Steam Turbine (50-1-T-1 and 50-1-T-2)

The STEAG power block design consists of two complete identical combined-cycle systems in parallel. Therefore, each power block has its own steam turbine bottoming cycle. The steam turbines 50-1-T-1 and 50-1-T-2 are tandem compound, reheat machines. The design is based on KWU HMN Series 400 MW turbines of 3 casing design with a two-pass intermediate-pressure and a two-pass low-pressure casing.

The HP end of 50-1-T-1 receives superheated steam at 2669 psig, 977°F (2813 psig, 995°F for Case EATS) and exhausts to the IP steam header operating at approximately 595 psig. The IP steam available after meeting the process IP steam demand is reheated to 977°F in the pressurized boiler (50-1-E-1:E-3) and flows to the IP end of 50-1-T-1. The inlet and exhaust conditions for the IP end of 50-1-T-1 are 525 psig, 977°F and 86 psig, respectively.

The LP end (50-1-T-2) is a condensing-type unit receiving steam at 86 psig and exhausting at 2-1/2" Hg abs. The turbine is provided with the capacity for steam extraction at several points to preheat boiler feedwater and to satisfy the overall steam balance of the plant. The surface condenser associated with 50-1-T-2 is designed for an 80°F CW inlet temperature and 19°F CW temperature rise.

Steam Turbine Generator (50-1-G-2)

The steam turbine (50-1-T-1&2) drives an appropriately rated electric generator suitable for North American applications. A summary tabulation of steam turbine performance and generator output is given in Table A-3.

Boiler Feedwater System

The design of the boiler feedwater system for the pressurized boiler combined cycle requires some distinctive features that differ from atmosperhic boiler combined cycles. Accordingly, the major equipment items are discussed below.

Boiler Feedwater Treating

The evaporation of boiler feedwater to dryness in the tubes of the pressurized boiler is typical of Benson-type boilers. This requires demineralization of all boiler feedwater fed to the boiler. To accomplish this, two parallel 50 percent demineralization units are used on the boiler feedwater upstream of preheating. When tube leakage in the surface condenser causes infiltration of cooling water into the boiler feedwater, the demineralization units allow the plant to be operated without damaging the boiler and the turbine.

Hydrazine treatment is provided for the removal of residual oxygen which may be present in the turbine condensate.

Deaerator (50-DA-1)

The deaerator in the STEAG cycle holds about 220 tons of available feedwater. The design is based upon spray-type deaeration such as in the STORK design. The deaerator is equipped with a high level alarm. In addition, 50-1-E-5 Condensate Subcooler is automatically bypassed at rising water level. If the water level rises further, then the steam extraction valve will be closed.

High-Pressure Boiler Feedwater Pump (50-1-P-1)

Each train in the STEAG power block is provided with two 50 percent High-Pressure Boiler Feedwater Pumps (50-1-P-1), each discharging at 3340 psig. These are typically Halberg HDV high-pressure pumps provided with Voith variable drives and electric motors. In order to meet NPSH requirements, the feed pump is divided into a low-speed section and a high-speed section. The low-speed section is directly coupled to the motor at approximately 1450 rpm. The high-speed section is driven at maximum 5000 rpm by a hydraulic drive. The startup of each train is accomplished with one 50 percent feed pump. The second pump is engaged as the load increases to complete the startup.

Boiler Feedwater Steam Preheaters (50-1-E-1 through 50-1-E-6)

Boiler feedwater to the Pressurized Boiler 50-1-B-1 is primarily heated by gland steam leakage and by extraction steam from Power Turbine 50-1-T-2, supplemented with low-pressure steam received from the gasification area. This preheating with steam is accomplished in multiple stages of steam preheaters. Each of these are separate shell and tube units. In the direction of boiler feedwater flow, the steam preheaters are as follows:

- 50-1-E-1 Turbine Gland Steam Preheater
- 50-1-E-2 First Extraction Steam Preheater
- 50-1-E-3 Second Extraction Steam Preheater
- 50-1-E-4 Third Extraction Steam Preheater
- 50-1-E-5 Condensate Subcooler
- 50-1-E-6 Fourth Extraction Steam Preheater

Turbine Gland Steam Preheater 50-1-E-1 is heated with gland steam leakage from turbines 50-1-T-1 and 50-1-T-2.

Second and Third Extraction Steam Preheaters 50-1-E-2 and 50-1-E-3 are heated with turbine extraction steam, which is supplemented with process steam from the gasification area for Cases MASW and EATS.

Fourth Extraction Steam Preheater 50-1-E-6 is also heated with turbine extraction steam. The hot condensate from 50-1-E-6 is subcooled in 50-1-E-5 to avoid excessive condensate flashing in the deaerator.

Flue Gas Boiler Feedwater Preheaters (50-1-E-8, 50-1-E-9 and 50-1-E-10)

The flue gas boiler feedwater preheater contains preheaters 50-1-E-8, 50-1-E-9, and 50-1-E-10 in a common housing similar to an HRSG. This device utilizes gas turbine exhaust to preheat boiler feedwater (BFW) for the pressurized boiler. The preheater is divided into two high-pressure BFW sections (50-1-E-8 and 50-1-E-9) and one low-pressure BFW section (50-1-E-10). The first high-pressure section (50-1-E-8) is connected in series while the other sections (50-1-E-9 and 50-1-E-10) are connected in parallel with the steam heated feed heaters (50-1-E-4 and 50-1-E-6).

The three sections of the flue gas boiler feedwater preheater are vertically arranged in an outdoor type enclosure. All heating surfaces are made of extended tubes. The entire STEAG Kombi-block can be isolated from the stack by a tight closing damper which is installed behind the flue gas preheater and which is automatically controlled and interlocked.

Gas Turbine and Air Compressor (50-1-GT-1 and 50-1-C-1)

Overall engine performance was calculated by STEAG, based on site conditions of 88°F and 14.4 psia, and integration with the pressurized boiler.

The STEAG turbine designs are based on a Kraftwerke Union (KWU) type V94 gas turbine, with the pressurized boiler replacing the standard combustor. The compressor section contains one extra stage over the standard design, in order to obtain a higher pressure ratio. The standard V94 design is rated at 85 MW when using natural gas or oil as the fuel. In the STEAG design the turbine output increases substantially.

Auxiliary Air Compressor (50-1-C-2)

For Case EATS only, an auxiliary air compressor is coupled to the main gas turbine shaft to provide additional air needed by the gasifiers. This axial compressor has an air intake and lube oil system in common with the main gas turbine aggregate. An air intercooler is not provided, and a partial load regulating device is not necessary in the absence of shutoff valves on the air discharge system.

Gas Turbine Generator (50-1-G-1)

The gas turbine (50-1-GT-1) drives an appropriately rated electric generator suitable for North American applications. A summary tabulation of gas turbine performance and generator output is given in Table A-4.

Start-up Boiler

The steam required for starting each train of the STEAG cycle in the power block depends on the startup method selected. If the pressurized boiler is fired with oil at startup, then an auxiliary startup boiler capable of supplying about 133,000 lb/hr of superheated 350 psig steam is sufficient. This can be a conventional package unit. The auxiliary steam is used for heating and initial operation of the steam cycle equipment.

EQUIPMENT STATE OF THE ART

The only equipment required by the STEAG combined cycle which has not been fabricated and operated in the size ranges needed for Cases MASW, EXTS, and EATS is the pressurized boiler. This statement by STEAG is based on current West German power plant technology.

Very good operating experience has been obtained with the prototype pressurized boilers in the 170 MW plant at Lünun, West Germany, which are producing 750,000 lb/hr steam at 1885 psia and 977°F. These boilers have been in operation over 10,000 operating hours since 1972. Apart from minor welding imperfections which had caused some leaks during the initial operating hours and were immediately corrected, there have been no problems with these boilers. STEAG anticipates that larger units of this type of boiler will similarly perform well.

STEAG has no reservations concerning the reliability of operation of other power block equipment, as these equipment items are proven in commercial operations.

Table A-1

POWER BLOCK PERFORMANCE SUMMARY - STEAG CASES

	Case MASW	Case EXTS	Case EATS
GENERATION			
Steam Turbines, kW	716,720	814,400	918,385
Gas Turbines, kW	215,177	210,790	204,638
Fuel Gas Expander, kW	--	21,020	--
Total Power Block, kW	931,897	1,046,210	1,123,023
HEAT REJECTION			
Process Cooling Rejection to Tower, MM Btu/hr	289	1,265	359
Power Block Heat Rejection,* MM Btu/hr	3,510	3,845	5,073
Total Heat Rejection to Tower, MM Btu/hr	3,799	5,110	5,432

*Includes mechanical and electrical losses of the power block

Table A-2

PRESSURIZED BOILER PERFORMANCE SUMMARY - STEAG CASES

DESCRIPTION	Case MASW	Case EXTS	Case EATS
DIMENSIONS			
Pressure Vessel Height, ft	82	82	82
Pressure Vessel Diameter (OD), ft	16.4	16.4	16.4
WEIGHT			
Weight Per Unit			
shell, tons	162	165	143
pressure parts, tons	192	198	170
structure plus platforms, tons	96	99	88
total tons	450	463	401
Weight: two boilers, tons	900	926	802
COMBUSTION AIR - TOTAL PLANT			
Flow, lb/sec	1703	1559	1143
Temperature, °F	686	667	686
Percent Excess	26	26	20
BOILER EXHAUST GAS - TOTAL PLANT			
Flow, lb/sec	2406	1846	2253
Temperature, °F	1742	1742	1742
Pressure, psig	120	148	148
EVAPORATOR-SUPERHEATER SECTION - TOTAL PLANT			
HP BFW Flow, lb/sec	1030	759	482
Inlet Temperature, °F	645	596	645
Inlet Pressure, psig	3517	3517	3517
Saturated Steam Flow from Process, lb/sec	0	745	0
HP Water Injection to Superheater, lb/sec	54	79	56
Total SH Steam Outlet Flow, lb/sec	1084	1583	538
Outlet Temperature, °F	977	977	995
Outlet Pressure, psig	2669	2669	2813
Evaporator Pressure Drop, psi	392	392	392
Superheater Pressure Drop, psi	203	203	203
Interconnecting Piping Pressure Drop, psi	109	109	109
Total Pressure Drop, psi	704	704	704
Absorbed Duty, MM Btu/hr	3040	3763	1522

Table A-2 (Continued)

PRESSURIZED BOILER PERFORMANCE SUMMARY - STEAG CASES

DESCRIPTION	Case MASW	Case EXTS	Case EATS
REHEATER SECTION - TOTAL PLANT			
Inlet Steam Flow, lb/sec	1077	1479	1403
HP Water Injection, lb/sec	19	26	24
Outlet Steam Flow, lb/sec	1096	1505	1427
Inlet Temperature, °F	590	590	590
Outlet Temperature, °F	977	977	977
Pressure Drop, psi	70	70	70
Total Internal Pressure Drop, psi	58	58	58
Outlet Pressure, psig	525	525	525
RH Duty, MM Btu/hr	949	1302	1273

Table A-3

HP/IP STEAM TURBINE PERFORMANCE SUMMARY - STEAG CASES

	Case MASW	Case EXTS	Case EATS
HP ELEMENT			
Throttle Conditions:			
Throttle Flow, lb/sec	1263.3	1583.2	1369.5
Steam Enthalpy In, Btu/lb	1436.5	1436.5	1444.1
Exhaust Flow to Process, lb/sec	181.5	100.0	92.6
Exhaust Enthalpy, Btu/lb	1284.3	1284.3	1284.3
IP ELEMENT			
Inlet Conditions:			
Inlet Flow, lb/sec	1095.8	1182.5	1427.6
Inlet Enthalpy, Btu/lb	1508.0	1508.0	1508.0
LP Extraction #1			
Pressure, psig	86.2	-	86.2
Flow, lb/sec	54.2	-	0.9
Enthalpy, Btu/lb	1314.0	-	1314.0
LP Extraction #2			
Pressure, psig	-	26.5	-
Flow, lb/sec	-	42.5	-
Enthalpy, Btu/lb	-	1236.6	-
LP Extraction #3			
Pressure, psig	24.2	8.8	-
Flow, lb/sec	13.3	42.8	-
Enthalpy, Btu/lb	1193.0	1193.0	-
LP Extraction #4			
Pressure, psia	4.3	4.3	30.0
Flow, lb/sec	29.5	9.3	-20.9**
Enthalpy, Btu/lb	1089.4	1089.4	1165.0
Quality, %	96.0	96.0	100.0
Exhaust Flow to Condenser			
Pressure, in - Hg	2.5	2.5	2.5
Flow, lb/sec	1001.4	1090.5	1450.0
Enthalpy, Btu/lb	1037.0	1037.0	1035.6
Quality, %	92.9	92.9	92.9
Cond. Circ. Water Flow, gpm	289,000	317,700	418,700
Power Output, kW*	716,720	814,400	918,385
Generator Voltage, kV	21	21	21

*At generator terminals

**Steam admission to turbine from gasification plant

Table A-4

GAS TURBINE PERFORMANCE SUMMARY - STEAG CASES

DESCRIPTION	Case MASW	Case EXTS	Case EATS
Inlet Air Temperature, °F	88	88	88
Inlet Air Pressure, psia	14.4	14.4	14.4
Relative Humidity, percent	54	54	54
Compressor Discharge Pressure, psia	139.2	166.3	167.0
Compressor Discharge Temperature, °F	686	667	686
Compressor Pressure Ratio	9.7	11.5	11.6
Compressor Inlet Air Flow, lb/sec	2147	1648	1949
Air to Process, lb/sec	445	89.4	1046
Exhaust Gas Flow, lb/sec	2406	1960	2392
Exhaust Gas Temperature, °F	950	924	893
Generator Power Output, kW*	215,177	210,790	204,638
Generator Voltage, kV	10.5	10.5	10.5
Weight of Gas Turbine with Accessories**, tons	269	297	269
Heaviest Piece of Transport, tons	126	139	115
Heaviest Piece to Lift (for erection), tons	126	139	115
Heaviest Piece to Lift (generator rotor), tons	43.2	47.3	47
Generator Rotor, tons	29	32	26

*At generator terminals

**Based on standard KWU Model V94 scaled to application

Appendix B

ATMOSPHERIC HRSG COMBINED-CYCLE SYSTEM DETAILS

GENERAL

The design of the combined-cycle system for Case EXTC78 was supplied by United Technologies Corporation (UTC), Power Systems Division, South Windsor, Connecticut. Design interface conditions between the fuel processing area and power block were provided by Fluor.

The UTC combined-cycle design consists of twelve gas turbine generating "units" operating in parallel. Each "unit" consists of two parallel gas turbines exhausting to a common heat recovery steam generator (HRSG). Steam from the twelve parallel HRSG's is used to drive a steam turbine bottoming cycle. All equipment designs which comprise the overall combined cycle are based on current (1978) state-of-the-art technology.

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 B-1. The power output is calculated at the generator terminals without margins for design or manufacturing tolerances except in the gas turbine unit, which is based on average field data. The calculated power outputs include approximately 2.5 percent deduction for mechanical and electrical losses including lube and seal oil pumps.

TECHNICAL INPUT DATA

Gas Turbine

The combined-cycle design is based on a typical gas turbine currently in production and available for delivery in 1978. Design combustor exit temperature is 1980°F. At this combustor exit temperature, the gas turbine design compression ratio is 13.4:1. Based on suction air at the site conditions of 14.4 psia, the clean fuel gas must therefore be delivered to the gas turbine fuel valve stations at a pressure greater than 193 psia (178.6 psig) in order to enter the machine. As shown on Process Flow Diagram EXTC78-50/51-1, clean fuel gas exits the expander 50-1-EX-1 at 228 psig, 767°F upstream of the gas turbine.

A combustor outlet temperature of 1980°F is within the design limits for current gas turbines in peaking service. Some minor upgrading of rotor materials in the expander section may be necessary for baseload operation at this temperature.

Steam Conditions

Steam conditions for the combined cycle are:

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

Heat Recovery Steam Generator (HRSG) Conditions

Low pressure flash gas from the Acid Gas Removal units is burned in the exhaust stream from the gas turbines. The resulting combined stream enters the HRSG at 948°F. Overall energy recovery in the HRSG was calculated using a flue gas stack temperature of 275°F. Based on the use of demineralized boiler feedwater, the boiler blowdown was assumed to be zero.

Process Interface

Heat integration between the fuel processing units and power block is utilized to increase the overall thermal efficiency of the system. Interface conditions of temperature, pressure, composition and flow rate for the streams involved in heat integration are based on the design of the process units.

The Case EXTC78 gasification-combined cycle design is not fully optimized from the standpoint of maximizing power generation. Studies directed toward development of optimal design criteria are presently being performed under another EPRI contract. It is anticipated that a future application of the resulting optimal design criteria will improve the overall thermal efficiency of Case EXTC78.

Steam Drivers

Condensing turbines are used for process drivers which develop more than 15,000 hp.

POWER BLOCK

Gas Turbine-Generator Unit

Gas Turbine (50-1-GT-1A&B). United Technologies has provided the overall gas turbine performance summary contained in Table B-2. The inlet air and exhaust duct pressure losses are comprised of allowances for silencers, ducting, after-burners and the heat recovery sections of the HRSG. The total power output of 705,390 kW is produced by 24 parallel turbines arranged in 12 parallel "units."

Generator (50-1-G-1). The gas turbine "unit" design uses two modular industrial turbines to drive a single air-cooled, two-pole electric generator, operating at 60 Hertz. Each generator has a double-ended drive incorporating flexible diaphragm couplings to accommodate thermal expansion. The generators utilize brushless excitation with solid state voltage regulators and integral lubrication systems.

Steam Cycle Selection

HRSG. A single HRSG 51-1-B-1 is coupled with each pair of gas turbines, 50-1-GT-1A&B, to recover heat from the turbine exhaust gases. Flash gas from the process plants is also burned in the gas turbine exhaust before entry to the HRSG. Performance of the HRSG for Case EXTC78 is summarized in Table B-3.

The HRSG generates saturated steam at three pressure levels; intermediate-pressure (IP) at 445 psig, low-pressure (LP) at 50 psig and low-pressure at 15 psig. The 15 psig steam generated in the HRSG is used as deaerating steam in Deaerator 51-DA-1.

No high-pressure (HP) steam is generated in the HRSG. The HRSG for Case EXTC78 differs in this respect from that for Case EXTC, presented in EPRI report AF-642. A mix of IP and HP steam generation in the Case EXTC78 HRSG may be possible, and could lead to a plant with higher overall efficiency.

As shown in Figure B-1, the arrangement of heat recovery sections of the HRSG in the direction of gas flow is as follows:

Reheater and Superheater	51-1-B-1:E-1 and 51-1-B-1:E-2
Economizer One	51-1-B-1:E-3
IP Evaporator	51-1-B-1:E-4
Economizer Two	51-1-B-1:E-5
LP (50 psig) Evaporator	51-1-B-1:E-6
Deaerator LP (15 psig) Evaporator	51-1-B-1:E-7

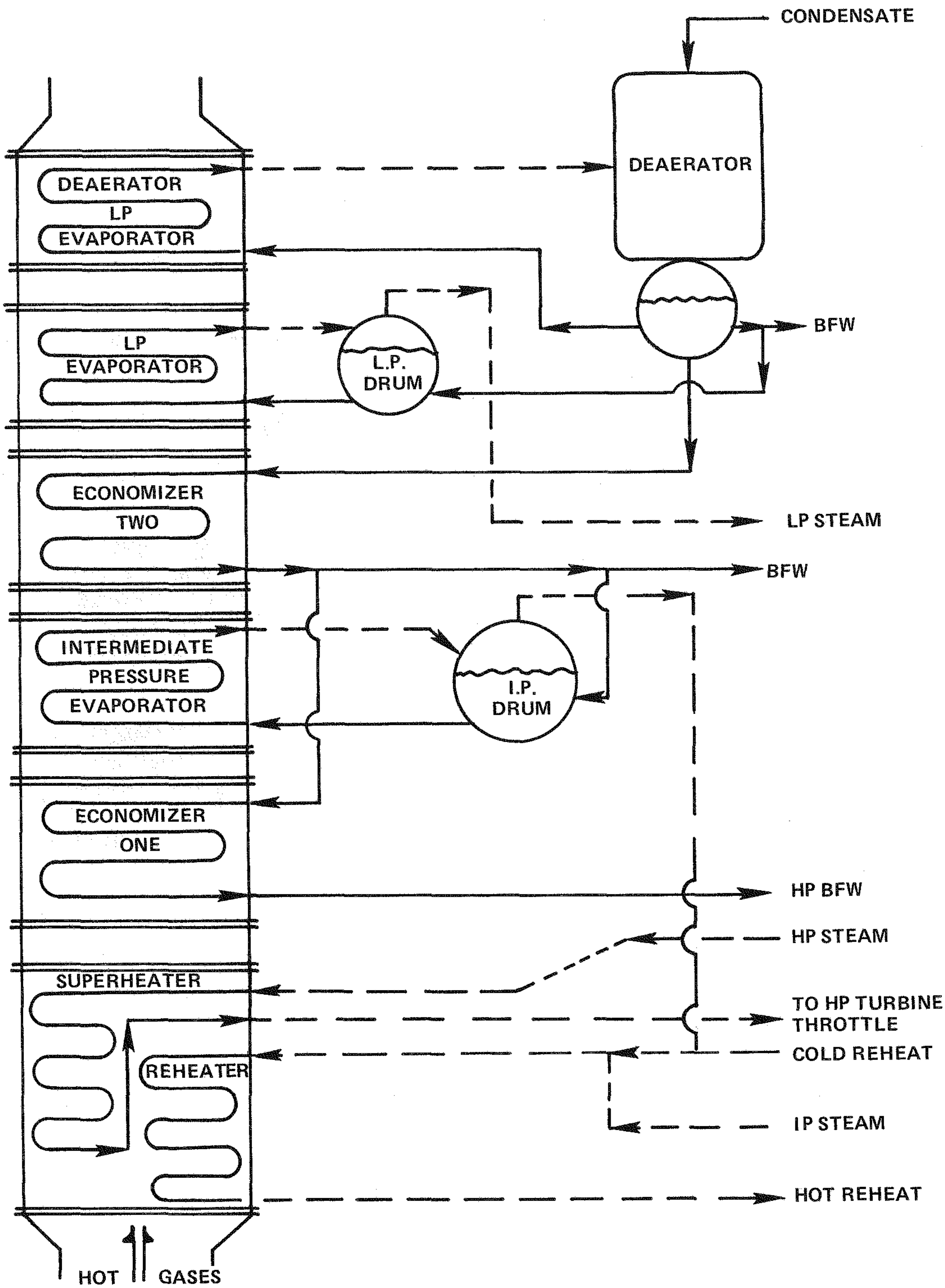

Figure B-1. Heat Recovery Steam Generator

High-pressure saturated steam from the gasification unit is heated to 900°F in superheater 51-1-B-1:E-2 and flows to the HP back pressure end of turbine 51-T-1. Saturated steam produced in the IP evaporator combines with both intermediate-pressure steam from the process generators and cold reheat steam from high-pressure steam turbine, 51-T-1, and is subsequently reheated to 900°F by passing through the reheater, 51-1-B-1:E-1. This superheated IP steam is then used in condensing turbines 11-1-T-1 and 11-1-T-2 and in the IP condensing end of turbine 51-T-1.

The 50 psig saturated steam produced in LP evaporator 51-1-B-1:E-6 combines with excess 100 psig steam from the process units and is used in condensing turbine 51-T-2.

LP evaporators 51-1-B-1:E-7 located in the topmost section of the HRSG's supply 15 psig deaerating steam to tray type deaerator, 51-DA-1. One common deaerator is provided for the multiple HRSG's and the process steam generators.

The boiler feedwater (BFW) from the deaerator is first preheated to the IP steam saturation temperature (458°F) in Economizer Two 51-1-B-1:E-5. A portion of the BFW flows to IP evaporator 51-1-B-1:E-4 and to the process IP steam generators to meet the makeup water demand of the IP steam generation system. The balance of the BFW is further preheated to the HP steam saturation temperature (598°F) in Economizer One 51-1-B-1:E-3 and then flows to the gasification unit waste heat boilers for the generation of HP steam.

Each HRSG is provided with its own LP and IP steam drums and corresponding BFW circulation pumps.

The HRSG exhaust (stack) gas temperature of 275°F, established in conjunction with the design of the low-pressure (15 psig) evaporator section, allows the gas side surface of the LP evaporator to operate safely above the dew point of acidic SO_2-bearing mist in the exhaust gas.

Steam Turbine-Generator Units

Steam Turbine (51-T-1). Steam turbine 51-T-1 is a single turbine consisting of HP back pressure and IP condensing ends. 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 445 psig. The total flow from the IP steam header is reheated to 900°F in the HRSG reheaters (51-1-B-1:E-1). The reheated IP steam available after meeting the process unit demands is sent to the IP end of 51-T-1.

The IP end of 51-T-1 is a condensing type unit receiving reheat steam at 400 psig and 900°F and exhausting at 2-1/2" Hg abs. The surface condenser, 51-E-8, associated with 51-T-1 is designed for cooling water (CW) flow in the tube side with 80°F CW inlet temperature and 15°F CW temperature rise.

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

Steam Turbine (51-T-2). Steam turbine 51-T-2 is a single LP condensing type unit. The machine receives 50 psig steam from LP Evaporator 51-1-B-1:E-6 and exhausts at 2-1/2" Hg abs. The surface condenser, 51-E-9, associated with 51-T-2 is designed for cooling water (CW) flow in the tube side with 80°F CW inlet temperature and 15°F CW temperature rise.

A summary tabulation of steam turbine performance and generator output is given in Table B-5.

Fuel Gas Expander-Generator Unit

Fuel Gas Expander 50-1-EX-1 is a single casing machine which receives fuel gas at 514 psig and 1000°F and exhausts to the combustion turbine, 50-1-GT-1, at 228 psig and 767°F. This expander drives a small electric generator.

EQUIPMENT STATE OF THE ART

Gas Turbine

Operating parameters for the major equipment items in the power block combined-cycle system of Case EXTC78 have been established based upon current industry practice. In particular, the combustion turbine inlet temperature of 1980°F permits the use of 1978 technology machines as opposed to the projected 1985 state-of-the-art equipment required for the 2400°F inlet temperature described in EPRI Report AF-642, Case EXTC (January 1978, page 297). Some minor upgrading of rotor materials may be necessary to accommodate a 1980°F temperature in baseload operation.

HRSG

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

Steam Turbines

The selected throttle steam conditions of 1450 psig/900°F/900°F reheat for turbine 51-T-1 present no problem to the state of the art.

Steam turbine 51-T-2 is commercially available.

Fuel Gas Expander

This equipment is commercially available.

Table B-1

POWER BLOCK PERFORMANCE SUMMARY - CASE EXTC78

GENERATION	
Gas Turbines, kW	705,390
Steam Turbines, kW	330,798
Fuel Gas Expander, kW	43,460
Total, Power Block, kW	1,079,648
HEAT REJECTION	
Process Cooling Rejection to Tower, M^2 Btu/hr	187.0
Power Block Heat Rejection,* M^2 Btu/hr	3,781.0
Total Heat Rejection to Tower, M^2 Btu/hr	3,886.0

*Includes mechanical and electrical losses of the power block

Table B-2

GAS TURBINE PERFORMANCE SUMMARY - CASE EXTC78

DESCRIPTION	
Compressor Inlet Air Duct Loss, in H_2O	4"
Turbine Exhaust System Loss, in H_2O	31.6
Compressor Air Flow, lb/sec	6,253.0
Air to Process, lb/sec	0.0
Fuel Flow, lb/sec	447.62
Turbine Exhaust Temperature, °F	943
Power Output, kW*	705,390
Flash Gas Fuel Flow, lb/sec	5.10
Total Exhaust Gas Flow, lb/sec**	6,685.7
Exhaust Gas Temperature, into HRSG, °F	948

*At generator terminals

**This flow is based on an estimated total leakage of 20 lb/sec

Table B-3

HRSG PERFORMANCE SUMMARY - CASE EXTC78

Exhaust Gas Flow, lb/sec	6,685.7
HP, SH AND RH SECTIONS	
Exhaust Gas Temperature In, °F	948
SH Temperature Out, °F	900
SH Pressure Out, psig	1,450
SH Enthalpy Out, Btu/lb	1,431
Saturated Steam from Process, lb/sec	323.36
SH Outlet Flow, lb/sec	323.36
SH Duty, MM Btu/hr	294
RH Enthalpy In, Btu/lb	1,241.1
RH Temperature Out, °F	900
RH Pressure Out, psig	400
RH Enthalpy Out, Btu/lb	1,469.8
RH Flow, lb/sec	998.3
RH Flow to Process Drivers, lb/sec	378.8
RH Duty, MM Btu/hr	822
Exhaust Gas Temperature Out, °F	778
ECONOMIZER NO. 1 SECTION	
HP BFW Enthalpy In, Btu/lb	440
HP BFW Flow, lb/sec	323.36
HP BFW Flow to Process, lb/sec	323.36
Duty, MM Btu/hr	206
Exhaust Gas Temperature Out	744
IP EVAPORATOR SECTION	
IP Drum Temperature, °F	458
IP Drum Pressure, psia	460
IP Steam Enthalpy Out, Btu/lb	1,204
IP Steam Evaporator, lb/sec	587.2
IP Evaporator Duty, MM Btu/hr	1,616
IP Steam to (from) Process, lb/sec	(87.75)

Table B-3 (Continued)

HRSG PERFORMANCE SUMMARY - CASE EXTC78

IP Steam to Cold RH, lb/sec	674.95
IP BFW Enthalpy In, Btu/lb	440
Exhaust Gas Temperature Out, °F	488
ECONOMIZER NO. 2 SECTION	
Water Enthalpy In, Btu/lb	221.6
Total BFW Flow, lb/sec	998.3
Outlet BFW Flow to Process, lb/sec	87.75
Duty, MM Btu/hr	771
Exhaust Gas Temperature Out, °F	361
50 PSIG EVAPORATOR SECTION	
50 psig Drum Temperature, °F	297
50 psig Drum Pressure, psia	65
50 psig Steam Enthalpy Out, Btu/lb	1,179
50 psig Steam Evaporator, lb/sec	67.4
50 psig Evaporator Duty, MM Btu/hr	233
Net 50 psig Steam from Process, lb/sec	6.3
50 psig Steam to 51-T-2, lb/sec	73.7
Water Enthalpy In, Btu/lb	218
Exhaust Gas Temperature Out, °F	323
LP EVAPORATOR AND DA SECTION	
LP Drum Temperature, °F	250
LP Drum Pressure, psia	30
Cond. Flow In, lb/sec	1,072.0
Cond. Enthalpy In, Btu/lb	141
Process Flows In, lb/sec	28.7
BFW Flow to Process, lb/sec	103.0
Duty, MM Btu/hr	291
Exhaust Gas Temperature Out, °F	275

Table B-4

HP/IP STEAM TURBINE
PERFORMANCE SUMMARY - CASE EXTC78

HP BACK PRESSURE ELEMENT

Throttle Conditions:	1,450 psig/900°F TT
Steam Enthalpy In, Btu/lb	1,431
Throttle Flow from HRSG, lb/sec	323.36
Throttle Flow from Process, lb/sec	0.0
Total Throttle Flow, lb/sec	323.36
Exhaust Flow to Process, lb/sec	0.0
Exhaust Enthalpy, Btu/lb	1,317.8

IP CONDENSING ELEMENT

Inlet Conditions:	400 psig/900°F TT
Inlet Enthalpy, Btu/lb	1,469.8
Inlet Flow, lb/sec	619.5
Exhaust Enthalpy, Btu/lb	1,027.3
Exhaust Flow to Process, lb/sec	0.0
Exhaust Flow to BFP, lb/sec	0.0
Total Flow to Condenser, lb/sec	619.5
Cond. Circ. Water Flow, gpm	214,100
Power Output, kW* (Total)	317,320

*At generator terminals

Table B-5

LP CONDENSING STEAM TURBINE
PERFORMANCE SUMMARY - CASE EXTC78

Inlet Conditions:	50 psig/299°F TT
Inlet Enthalpy, Btu/lb	1,179.9
Inlet Flow, lb/sec	73.7
Exhaust Enthalpy, Btu/lb	1,004.3
Total Flow to Condenser, lb/sec	73.7
Cond. Circ. Water Flow, gpm	24,100
Power Output, kW*	13,478

*At generator terminals

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 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	Power Generation (Cases MASW, EXTS, and EATS)
		50	Gas Turbine and Power Generation (Case EXTC78 only)
		51	Heat Recovery and Power Generation (Case EXTC78 only)

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

Appendix D

CASE DESIGNATIONS

A letter code has been prepared to shorten and simplify the references to the various cases given in this report, as follows:

Case MASW - Moving-bed, Air-blown, STEAG cycle plant operating on Western coal. This plant uses the Lurgi dry ash gasifier.

Case EXTS - Entrained-bed, oXygen-blown, Texaco gasifier, with a STEAG cycle power plant.

Case EATS - Entrained-bed, Air-blown, Texaco gasifier, with a STEAG cycle power plant.

Case EXTC78 - Entrained-bed, OXygen-blown, Texaco gasifier, with a Combined-cycle power plant using technology available in 1978. This case uses a combined-cycle power plant designed by United Technologies Corporation.

The letter codes for analogous cases in other EPRI reports are as follows:

Case MACW - Moving-bed, Air-blown, Lurgi gasifier, Combined-cycle power plant using mid-1980 technology and Western coal.

Case EXTC - Entrained-bed, oXygen-blown, Texaco gasifier, Combined-cycle power plant using mid-1980 technology.

Case EATC - Entrained-bed, Air-blown, Texaco gasifier, Combined-cycle power plant using mid-1980 technology.