Economics of Fuel Gas From Coal—An Update

EPRI

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Prepared by
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Irvine, California

ELECTRIC POWER RESEARCH INSTITUTE

Economics of Fuel Gas From Coal—An Update Including the British Gas Corporation's Slagging Gasifier

AF-782
Research Project 239

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ABSTRACT

This report presents the results of an economic screening study for British Gas Corporation's oxygen blown slagging coal gasification process to produce intermediate Btu fuel gas, and an update of the economic sections of an earlier report (EPRI AF-244) which included air and oxygen blown Lurgi moving bed, U-Gas fluidized bed and Combustion Engineering's entrained processes for fuel gas production.

The process arrangement presented in this report (Case MXS) extends the work covered in EPRI AF-244 to include the moving bed slagging ash gasifier. All of the processes investigated produce fuel gas which could be used in fossil fired power plants.

This evaluation was based on a complete "grass roots" facility sized to conform to present electric utility practice of building units of approximately 1000 MW capacity.

The conclusion reached in this supplement report is that, within the accuracy of the study, fuel gas costs projected for the moving bed process, using the BGC slagging gasifier, are competitive with costs projected in earlier studies based on fluidized bed and entrained processes. The major assumption underlying this conclusion is that the BGC slagging gasifier will operate successfully on a commercial scale in exactly the same manner as is represented by the performance estimates used for this study.

It is recommended that further development is required to obtain better data to confirm the cost projections reported here. If such pilot plant data confirms these cost projections further, development of the slagging gasifier should be encouraged.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This final report, Economics Of Fuel Gas From Coal - An Update Including The British Gas Corporation's Slagging Gasifier, is the second in a series of evaluations aimed at determining the cost of producing clean low Btu and intermediate Btu fuel gas from coal. The first report, EPRI AF-244, Economics Of Current And Advanced Gasification Processes For Fuel Gas Production, presented an economic evaluation of fuel gas produced from Lurgi moving bed, dry ash gasifiers, U-Gas fluidized bed systems and Combustion Engineering two-stage entrained coal gasifiers. This report was published in July 1976 and was based on economic criteria that were somewhat different than those now being employed by EPRI.

PROJECT OBJECTIVE

This 9 month project had two major purposes. The first objective was to evaluate the cost of producing clean, intermediate Btu fuel gas from an oxygen blown British Gas Corporation slagging bottom gasifier based on Illinois #6 coal. The second objective was to put the previously developed fuel gas results (EPRI AF-244) on a consistent basis with the new BGC slagger evaluation such that results could be compared and incentives for further development could be evaluated.

CONCLUSIONS AND RECOMMENDATIONS

The results indicate that, within the accuracy of the study, fuel gas costs projected for the BGC slagger are competitive with similar costs projected for both the U-Gas gasifier and Combustion Engineering's two-stage entrained system. The results also indicate a substantial decrease in fuel gas costs for the BGC slagging technology over gas produced from existing dry ash Lurgi gasifiers. Such results must be treated with extreme caution, as of all the gasifiers evaluated, the Lurgi system is the only one that has been successfully demonstrated at full commercial scale. Therefore, data used for the Lurgi evaluation can be considered to be reliable and defensible. Operating information for all other gasifiers studied was either supplied by the process developer or estimated by EPRI. None of these data can be defended on the basis of large scale operation. Some reservations concerning projected operating data for the U-Gas and the Combustion Engineering gasifiers were discussed in the earlier fuel gas report (EPRI AF-244). Projections for the performance of the BGC Slagger on Illinois #6 coal have been based on pilot plant operations using Scottish coal. The cost estimates presented in this report for the BGC Slagger-Illinois #6 coal case must be treated cautiously until such time as the assumptions concerning capacity, tar recycle, oxygen and steam requirements, fines handling capabilities and operability with a caking coal have been demonstrated, at least at the pilot plant level. The results of this study indicate that if the design assumptions used can be confirmed by pilot plant data, further development of the BGC slagging gasifier should be encouraged.

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SUMMARY

This study is an extension and update of earlier work[1] on production of fuel gas from coal using the gasification route. The objective of this study was to evaluate the moving bed slagging gasification technology for fuel gas production from coal. British Gas Corporation's oxygen blown slagging gasifier was selected to represent the moving bed oxygen blown slagging technology. This case is identified as Case MXS (Moving bed oXygen blown Slagging gasification). The report presents the results of the study performed to establish capital costs and cost of services for a complete grass roots plant for Case MXS. In Appendix B, updated economic sections of EPRI AF-244[1] are presented reflecting escalation to a mid-1976 pricing basis plus other economic criteria revisions required to put all the cost estimating on a consistent basis with Case MXS.

The plant design basis for Case MXS was consistent with other fuel gas cases reported previously,[2] i.e., 10,000 tons/day of Illinois No. 6 coal. The plant was designed to deliver product fuel gas at 25 psig. This delivery pressure could be increased with minor economic penalties because the gasifier in this process operates at relatively high pressure.

Fuel gas purification systems were specified such that the product gas should contain no more than 1.0 pound of sulfur dioxide equivalent per million Btu of coal charged and should be essentially free of other contaminants such as particulates, ammonia, tars, oils, naphthas and phenols. One difference between Case MXS and other moving bed cases (Lurgi) reported previously[3] is that in Case MXS the liquid hydrocarbons and coal fines separated from the fuel gas are recycled back to the gasifiers to extinction. As a result no liquid hydrocarbon byproducts are produced. Ammonia was recovered and was assigned a product value of $100/ton. Sulfur removal and recovery were achieved by employing Allied Chemical Corporation's Selexol process for removal of hydrogen sulfide from the crude gas followed by a Claus sulfur recovery plant and a Beavon-Stretford tail gas treating unit.

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

As in other technologies reported earlier[1], Case MXS produced substantial quantities of heat that had to be utilized efficiently. Sources of this heat were oxidant compression intercooling, gasifier jacketing or temperature control, gasifier effluent cooling and stack gas heat recovery. This heat was used to generate steam and supply process heat. Excess steam was expanded through a turbogenerator to produce electric power. The bottoming cycle using isobutane in a low temperature Rankine cycle scheme incorporated in other fuel gas cases[2] was not used for Case MXS. The cost of this equipment represents such a small part of the plant designs, that this minor design difference does not alter the conclusions of the cost study. The amount of by-product power lost is insignificant. A subsequent Fluor study established that low level heat recovery in a bottoming cycle would be uneconomical for the coal gasification plants studied. Case MXS produces by-product electric power. Net electric power exported or consumed was evaluated at 25 mils/kWh plus the cost of coal at a heat rate of 9000 Btu/kWh.

The technical criteria used in the plant design for Case MXS are given in the Criteria section of this report. Table S-1 shows some of the important operating features of this case, along with similar data[3] from the earlier study.

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

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

1. "Economics of Current and Advanced Gasification Processes for Fuel Gas Production," EPRI AF-244, July 1976.
2. Ibid.
3. Op. Cit., Table S-1.

TABLE S-1

SUMMARY OF OPERATING RESULTS

	BGC SLAGGER MXS	LURGI Moving Bed		U-GAS Fluidized Bed		COMBUSTION ENGINEERING Entrained Bed	
	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen
GASIFICATION							
Air or Oxygen, lb/lb maf coal (1)	0.534	2.56	0.541	3.44	0.705	4.88	0.932
Oxidant Temperature, °F	214	342	321	900	480	600	800
Steam, lb/lb maf coal	0.34	1.65	2.58	0.646	0.586	None	None
Exit Pressure, psig (in. water)	320	302	302	325	325	(-0.5)	(-0.5)
Exit Temperature, °F	820	955	1078	1660	1550	1700	1700
Coal Carbon Converted to CH_4, atom %	11.2	10.6	14.0	10.3	10.8	0.001	0.0824
Cold Gas Efficiency, % of coal HHV (2)	96.37	90.9	91.2	83.3	89.4	69.8	82.4
OVERALL PLANT							
System Cold Efficiency, % of coal HHV (3)	81.3	70.4	68.3	79.7	85.1	77.2	77.6
Net Fuel Gas, mscf/ton maf coal (1)	60	86.5	50.3	136.0	71.9	170.5	71.6
Fuel Gas HHV, Btu/scf (1)	379	179	302	158	323	113	312
Liquid Hydrocarbons, #/ton maf coal	None	161	161	None	None	None	None
Net By-Product Electric Power, kwh/ton maf coal	35.6	213	177	202	140	294	(35)
By-Product Ammonia, #/ton maf coal	28.7	28.5	28.5	2.37	1.42	None	None
Process Makeup Water, mgal/ton maf coal	0.16	0.469	0.734	0.215	0.180	0.060	0.051
Cooling Tower Heat Rejection, % of coal HHV	8.6	21.3	20.8	19.1	12.8	12.2	9.8
Cooling Tower Makeup Water, mgal/ton maf coal	0.20	0.635	0.501	0.666	0.460	0.552	0.417

NOTES

(1) Dry Basis.

(2) (HHV of crude gas including by-product liquid hydrocarbons where applicable)(100)/(HHV of coal).

(3) [(HHV of clean fuel gas) + (HHV of by-product liquid hydrocarbons) + (Net Power)(9000 Btu/kwh)](100)/(HHV of coal).

Total capital requirement for Case MXS was determined by adding capital related charges such as preproduction costs, paid-up royalties, initial chemical and catalyst costs, construction loan interest and working capital to the estimated plant investments.

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

The cost of services for both Case MXS and the update (Appendix B) of the AF-244 results were calculated on a consistent basis by adding capital related charges such as depreciation bond interest, return on equity and income tax to the operating charges. The values of by-product power and by-product ammonia were credited against other operating charges.

A summary of economic results for this study and the updated results from Appendix B are presented in Table S-2. Process contingencies (see Table C-6) for the updated cases from AF-244, have been changed to be consistent with the combined cycle study.[1]

As in the previous study, AF-244[2], fuel gas costs have been projected for operating load factors of 70 and 90 percent. The economic results presented in Table S-2 must be used with the utmost caution. The original purpose of the fuel gas

1. "Economics Studies of Coal Gasification Combined Cycle Systems for Electric Power Generation," EPRI AF-642, January 1978.
2. "Economics of Current and Advanced Gasification Processes for Fuel Gas Production," EPRI AF-244, July 1976.

TABLE S-2

SUMMARY OF ECONOMIC RESULTS

	BGC SLAGGER MXS	LURGI Moving Bed		U-GAS Fluidized Bed		COMBUSTION ENGINEERING Entrained Bed	
	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen
PRODUCTION AT DESIGN CAPACITY							
Net Fuel Gas, MM Btu/day (1)	196,937	134,376	131,880	184,872	201,432	185,664	196,920
Liquid Hydrocarbons, MM Btu/day (2)	-	21,912	21,912	-	-	-	-
By-Product Power, MW	12.77	76.4	63.7	72.5	50.2	106.0	(12.4)
TOTAL CAPITAL (3)							
Total Capital @ 70% Operation Factor, $1,000	391,496	582,652	704,393	474,790	459,445	397,254	390,053
Total Capital @ 90% Operating Factor, $1,000	391,722	583,731	704,832	475,084	459,701	397,525	390,278
AVERAGE COST OF SERVICES (3) (4)							
Annual Cost @ 70% Operating Factor, $1,000/year	140,369	169,077	199,722	149,182	150,780	128,036	149,358
Per Unit @ 70% Operating Factor, $/MM Btu	2.79	4.23	5.08	3.16	2.93	2.70	2.97
Annual Cost @ 90% Operating Factor, $1,000/year	157,768	183,445	215,143	164,176	166,962	141,000	169,149
Per Unit @ 90% Operating Factor, $/MM Btu	2.44	3.57	4.26	2.70	2.52	2.31	2.61

NOTES

(1) Heating value plus sensible heat.

(2) Higher heating value only.

(3) Mid-1976 dollars and $1/MM Btu coal.

(4) Capital includes the cost of generating equipment required to produce by-product power.

studies was to compare the economic advantages associated with either air or oxygen blowing a particular gasifier. Caution should be exercised in comparing the economics of the different technologies due to the fact that they are all at very different stages of development. Within the accuracy of this study, the BGC slagger apparently reduces the cost of gas from the moving bed gasifier to levels competitive with the entrained processes and the fluidized bed processes.

Some comments as to the present suitability of the moving bed slagging technology for commercialization seem in order. The BGC oxygen blown slagging gasifier has been demonstrated in fairly large scale equipment by the British Gas Corporation at the Westfield Development Center. However, it is important to realize that the gasifier operating characteristics used for the performance of this study (based on Illinois #6 coal) were projected from reported pilot plant operating data based on Scottish coal. Until such time as the assumed performance characteristics (steam/coal, oxygen/coal, liquid hydrocarbon recycle, fines recycle, capacity and ability to handle caking coal) are confirmed for Illinois #6 coal, the cost projections presented in this report are to be considered to be preliminary and should be treated with caution. A DOE sponsored 60 MM scfd demonstration plant for SNG production in Ohio based on this technology is currently in the design phase.

INTRODUCTION AND SCOPE - CASE MXS

The work reported here presents an extension and update of the earlier study[1] done for EPRI by Fluor. Lurgi's gasification process was studied as representative of the moving bed technology in the previous work.[1] This report presents results of studies performed to establish capital costs and cost of services for a complete grass roots plant for fuel gas production from coal, based on another moving bed technology, British Gas Corporation's oxygen blown slagging gasifiers. This case is identified as Case MXS.

The basis for the preliminary plant design for Case MXS is the same as other cases reported previously[2], i.e., 10,000 short tons/day of Illinois No. 6 coal at design capacity. Average cost of services is computed at 70 and 90 percent operating factors. The gasifier material balances were supplied by Electric Power Research Institute. Preliminary plant design and plant cost estimates were developed by Fluor.

For economic evaluation, revised EPRI criteria were used. These are consistent with the ones used in a recent EPRI report.[3] Included in the Appendix B are the economic sections of the earlier fuel gas report[1] as recalculated using the revised economic data supplied by EPRI. This update includes using revised process contingencies (see Economic Criteria, Table C-6) to be consistent with the combined cycle studies.[4]

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1. "Economics of Current and Advanced Gasification Processes for Fuel Gas Production," EPRI AF-244, July 1976.
2. Ibid.
3. "Economic Studies of Coal Gasification Combined Cycle Systems for Electric Power Generation," AF-642, January 1978.
4. Ibid.

CRITERIA - CASE MXS

TECHNICAL CRITERIA - CASE MXS

Plant designs were based on criteria established by the Electric Power Research Institute (EPRI). These criteria included coal data, site location, gasifier material and heat balances, gasifier equipment requirements and general plant requirements and are consistent with earlier work.[1]

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

Table C-1 gives the analysis of Illinois Number 6 coal used for this case. Coal was assumed delivered to the site washed and sized. If experience were to demonstrate that this assumption was not reliable, then additional coal handling equipment would be required. This would slightly affect overall plant costs, but would not alter the comparisons between Case MXS and other cases.[2]

Fuel gas is delivered at the plant battery limits at 25 psig. Net plant products were restricted to fuel gas, electricity, sulfur and ammonia. Hydrocarbon by-products are totally recycled back to the gasifiers. Plant sulfur emissions are restricted to 1 lb SO_2/MM Btu (HHV) of coal fired.

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

- Power Recovery
- Cooling tower
- Plant and instrument air

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

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

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

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

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 unit, and on the flow diagrams. The degree of redundancy is compatible with a 90 percent onstream factor in the early years of plant life. The plant designs depicted here are intended to represent what is possible when the technology is fully established, and not to necessarily reflect the approach to be taken on a "first of a kind" plant.

TABLE C-1

COAL ANALYSIS

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

TABLE C-2

SITE CONDITIONS

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

TABLE C-3

WATER ANALYSIS

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

ECONOMIC CRITERIA - FUEL GAS PLANTS

A set of criteria for estimating capital requirement and cost of services was supplied by EPRI. These criteria are summarized in Tables C-4 and C-5. These criteria were applied to the new Case MXS being reported, as well as to the updating of the cases previously reported in AF-244.[1]

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

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

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

The capital costs in this report are based on mid-1976 pricing, whereas the earlier work (EPRI AF-244) was based on mid-year 1975 pricing and different economic criteria. The economic sections of the earlier work (EPRI AF-244) which have been updated to mid-1976 pricing and have been put on a consistent basis with Case MXS are reported here in Appendix B.

TABLE C-4

CAPITAL INVESTMENT BASIS FOR GASIFICATION BASED FUEL GAS PLANTS

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

ITEM	BASIS
	result from a more detailed design of a definitive project at an actual site. An allowance of 15% of the sum of the Process Plant Investment and the General Facilities cost is used.
Process Contingency	- This contingency factor is to be applied to unproven technology in an effort to quantify the uncertainty in the design, performance and cost of the commercial scale equipment. Process contingency allowances are shown in Table C-6.
Total Capital Requirement	- The total capital requirement includes all capital necessary to complete the entire project. These items include: (a) Total plant investment. (b) Royalties. (c) Preproduction costs. (d) Construction loan interest. (e) Initial chemical and catalyst charge. (f) Working capital.
Paid-up Royalties	- 0.5% of total plant investment.
Preproduction Costs	- One month variable operating costs excluding coal. Variable costs are catalysts and chemicals, utilities, and maintenance materials. - Two month's fixed costs excluding income taxes. Fixed costs are operating and maintenance labor, administrative and support labor, general and administrative expense, and property taxes and insurance.

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

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

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

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

TABLE C-5

COST OF SERVICES BASIS FOR GASIFICATION BASED FUEL GAS PLANTS

ITEM	BASIS
Operating Load Factor	- 70% and 90%
Cost of Coal Delivered	- \$1.00/MMBtu and \$2.00/MMBtu
Chicago City Water	- 40 cents/1,000 gallons
Ash Disposal	- \$1.00/ton
By-Product Ammonia Credit	- \$100/ton
By-Product Sulfur Credit	- None
By-Product Electric Power Credit	- 25 mils/kWh, plus coal cost at 9,000 Btu/kWh heat rate
By-Product Hydrocarbons Credit	- Credited at coal cost, i.e., either \$1.00/MM Btu or \$2.00/MM Btu
	- Coal cost at 9000 Btu/kWh heat rate
Maintenance	- Annual maintenance costs are normally estimated as a percentage of the total installed plant cost of the facilities. The percentage varies widely depending on the nature of the processing conditions and the type of design. Maintenance costs shown below were used.

Process Unit	Maintenance % of Total Plant Investment/Yr
Coal Handling	3.0
Oxidant Feed	2.0
Gasification & Ash Handling	4.5
Gas Cooling	3.0

ITEM	BASIS
	(see maintenance table below)

Process Unit	Maintenance % of Total Plant Investment/Yr
Acid Gas Removal & Sulfur Recovery	2.0
Fuel Gas Compression	3.0
Process Condensate Treating	3.0
Steam, Condensate & BFW	1.5
Power Equipment	1.5
Support Facilities	1.5

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

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

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

TABLE C-6

PROCESS CONTINGENCIES

		PERCENT						
		BGC SLAGGER	LURGI		U-GAS		COMBUSTION ENGINEERING	
UNIT	CASE	MXS	MA	MX	FA	FX	EAL	EXL
Coal Handling		0	0	0	0	0	5	5
Oxidant Feed		0	0	0	0	0	0	0
Gasification		15	5	5	50	50	35	35
Ash Handling		5	5	5	15	15	5	5
Gas Cooling		0	0	0	0	0	10	10
Acid Gas Removal		5	5	5	5	5	5	5
Sulfur Recovery (Claus)		0	0	0	0	0	-	-
Tail Gas Treating		15	15	15	15	15	-	-
Process Condensate Treatment		0	0	0	0	0	-	-
Steam, Condensate and BFW		0	0	0	0	0	0	0
Support Facilities		0	0	0	0	0	0	0
Power Equipment		0	0	0	0	0	0	0

PLANT DESCRIPTION - CASE MXS

A grass roots plant for fuel gas production is shown schematically on Block Flow Diagram MXS-1-1. This plant is based on gasifying 10,000 ST/day of Illinois No. 6 coal in a moving bed slagging bottom gasifier. The block flow diagram represents Case MXS, which is based on using oxygen blown slagging bottom gasifiers currently under development by British Gas Corporation. Major units in the plant, the number of operating trains, major stream flows at 100 percent of capacity operation and certain key stream heat contents are shown on the block flow diagram.

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

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

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

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

TABLE MS-1

MAJOR EQUIPMENT SECTIONS - CASE MXS

Unit No.	Name	Case MXS Operating	Spare
10	Coal Preparation	1	0
11	Oxidant Feed System		
	. Air Compression	3	0
	. Air Separation	3	0
	. Oxygen Compression	3	0
	. Gas Expansion	3	0
20	Gasification	3*	1*
20	Ash Handling	1	0
21	Gas Cooling	3	0
22	Acid Gas Removal	3	0
23	Sulfur Recovery & Tail Gas Treating	2	1
24	Process Condensate Treating		
	. Tar Oil Separation	2	1
	. Phenol Extraction & Ammonia Recovery	1	0
30	Steam, BFW and Condensate System		
	. High Pressure Boiler	3	1
	. Condensate Collection & Deaeration	1	0
	. Water Treating	1	0
31	Power Recovery	1	0
32	Cooling Water System	1	0
40	Effluent Water Treating	1	0

*Each operating train includes two parallel gasifiers.

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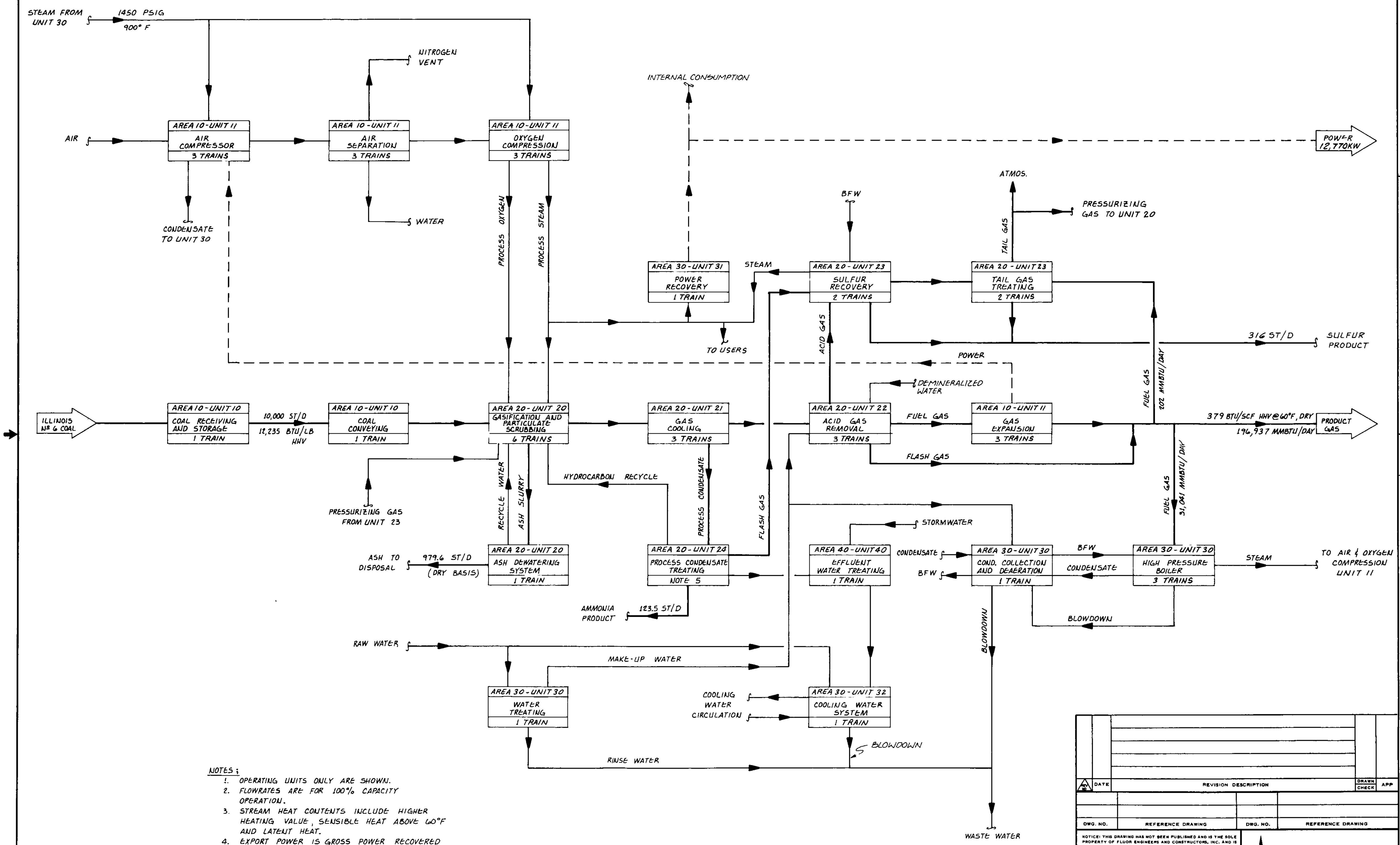

NOTES:
1. OPERATING UNITS ONLY ARE SHOWN.
2. FLOWRATES ARE FOR 100% CAPACITY OPERATION.
3. STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE, SENSIBLE HEAT ABOVE 60°F AND LATENT HEAT.
4. EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
5. PROCESS CONDENSATE TREATING CONSISTS OF TWO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

DATE	REVISION DESCRIPTION	DRAWN CHECK	APP

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

FLUOR

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

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

SCALE NONE | DRAWING NUMBER MXS-1-1 | REVISION 1

4483A-MXS-1-1

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COAL PREPARATION

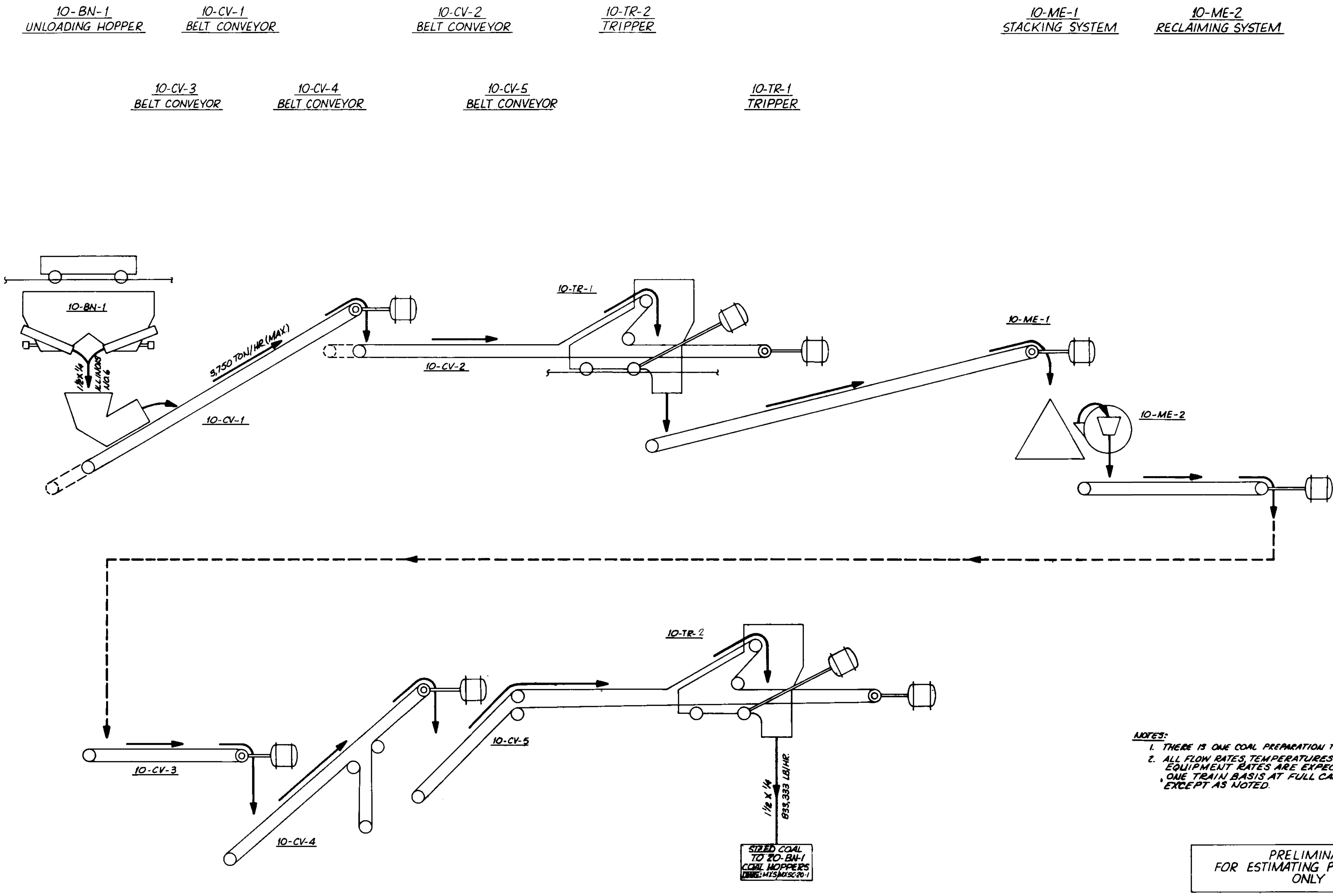
Process Flow Diagram MXS/MXSC-10-1 depicts the process arrangement of equipment in this section.

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

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

Equipment Notes

All equipment is commercially available.



NOTES:

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

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REV.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

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

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

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

MXS/MXSC-10-1

OXIDANT FEED

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

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

The 31,100 hp required by each compressor is supplied by a combination of a steam turbine, 11-1-T-1, and a product gas expander, 11-1-EX-1. Both the drivers are on a common shaft. The fuel gas expander, 11-1-EX-1, extracts about 17,600 hp from the hot fuel gas at 400°F and 280 psig by expanding it to 25 psig. The fuel gas subsequently flows to the fuel gas header. The balance of the compressor horsepower (13,500 hp) is supplied by the steam turbine, 11-1-T-1. The steam turbine is a condensing type machine operating at inlet steam conditions of 1450 psig and 900°F with exhaust pressure of 2-1/2" Hg abs. The drivers are designed to handle the turndown or upset conditions.

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

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

The 9,950 hp oxygen compression requirement is supplied by a backpressure turbine, 11-1-T-2. The inlet steam condition is 1450 psig, 900°F with backpressure at 400 psig. The exhaust steam flows to the medium pressure steam header.

Equipment Notes

All the equipment is commercially available.

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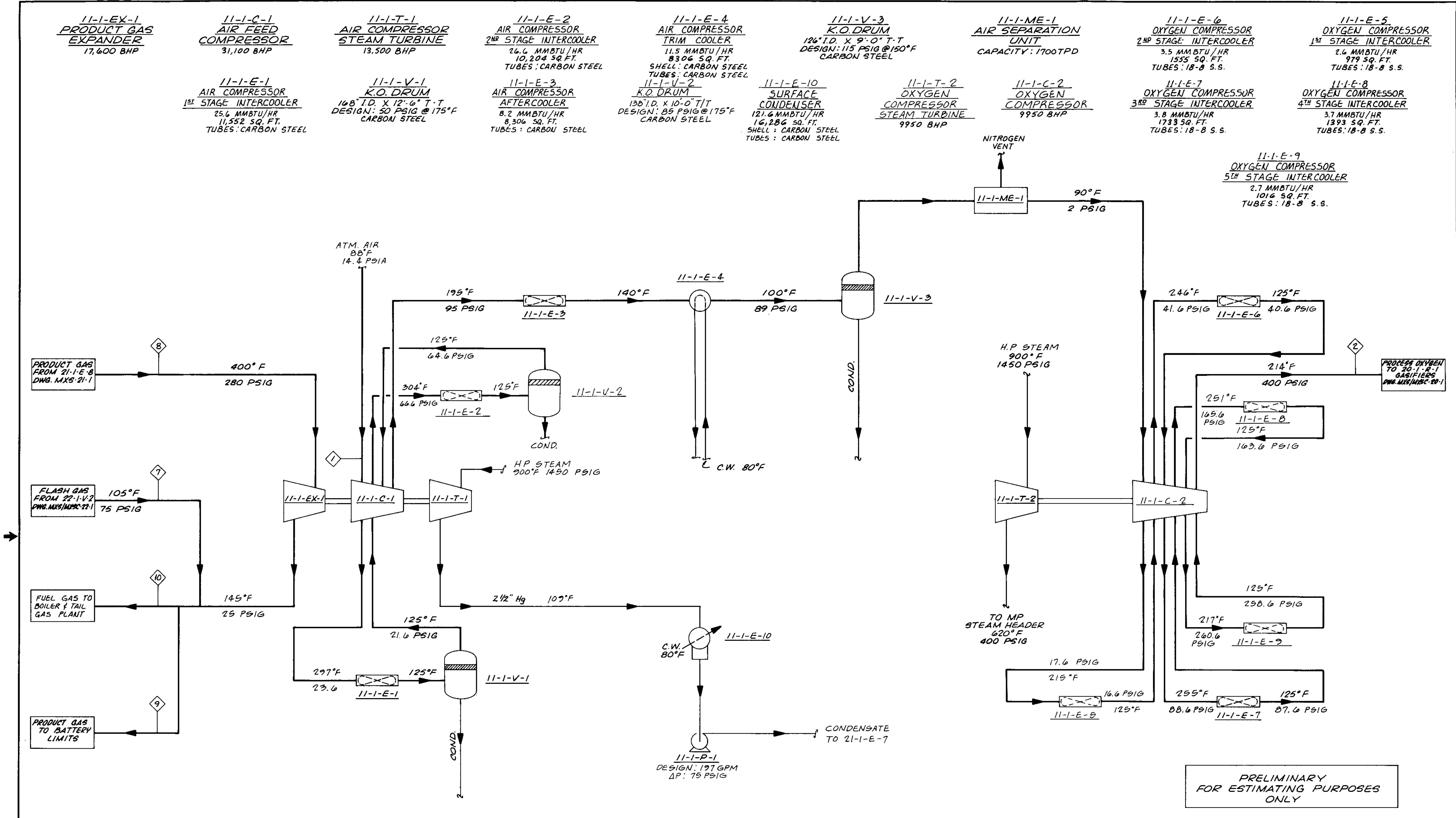

MATERIAL BALANCE												
STREAM NUMBER	1		8		7		10		9		2	
COMPONENT	AIR FEED		PRODUCT GAS		FLASH GAS		FUEL GAS TO BOILERS		PRODUCT GAS		PROCESS OXYGEN	
	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %	MPH	MOL %
CH_4	—	—	1,702.5	7.78	14.2	12.36	233.8	7.80	1,482.9	7.80	—	—
C_2H_4	—	—	47.8	0.22	0.8	0.70	6.6	0.22	42.0	0.22	—	—
C_2H_6	—	—	70.5	0.32	1.1	0.96	9.8	0.33	61.8	0.33	—	—
H_2	—	—	6,755.0	30.87	16.2	14.10	922.3	30.78	5,848.9	30.78	—	—
CO	—	—	12,851.5	58.73	58.0	50.48	1,758.3	58.69	11,151.2	58.69	—	—
CO_2	—	—	234.1	1.07	18.3	15.93	34.4	1.15	218.0	1.15	—	—
H_2S	—	—	34.6	0.16	4.5	3.92	5.3	0.18	33.8	0.18	—	—
COS	—	—	8.6	0.04	0.2	0.17	1.2	0.04	7.6	0.04	—	—
NH_3	—	—	—	—	—	—	—	—	—	—	—	—
N_2	16,264.2	77.05	99.1	0.45	0.3	0.26	13.5	0.45	85.9	0.45	81.6	2.00
O_2	4,323.4	20.48	—	—	—	—	—	—	—	—	3,999.4	98.00
H_2O	521.4	2.47	78.8	0.36	1.3	1.12	10.9	0.36	69.2	0.36	—	—
TOTAL MPH	21,109.0	100.00	21,882.5	100.00	114.9	100.00	2,996.1	100.00	19,001.3	100.00	4,081.0	100.00
TOTAL LB/HR	603,380		420,560		2,940		57,680		365,775		130,270	
MOL. WT.	28.58		19.22		25.61		19.25		19.25		31.92	

NOTES:

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

DRAWN BY: T. BUJACICH	PROCESS FLOW DIAGRAM
CHECKED BY: R. RICE	OXIDANT FEED SYSTEM
SUPERVISOR: R. RICE / RELEASE DATE	MOVING BED SLAGGING - OXYGEN BLOWN
SUPERVISING ENGR: T.H. TAN / INITIALS	ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIF.
PROJECT ENGR. / APP. DATE	SCALE: NONE — DRAWING NUMBER: MXS-11-1 — REVISION: 1
CLIENT / APP. DATE	

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4483A-MXS-11-1 (S.X.3)

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GASIFICATION AND ASH HANDLING

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

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

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

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

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

Oxidant and steam, together with recycle coal fines and hydrocarbon products, enter the gasifier near the bottom through an array of tuyeres. The intense heat created by the exothermic reactions in the "raceway" between opposing tuyeres maintains a temperature of approximately 3500°F in the bottom of the gasifier allowing ash to be withdrawn as a molten slag. Oxidant flow rate is controlled to accomplish complete gasification of coal. Steam rate is controlled to allow the ash to form into a molten slag.

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

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

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

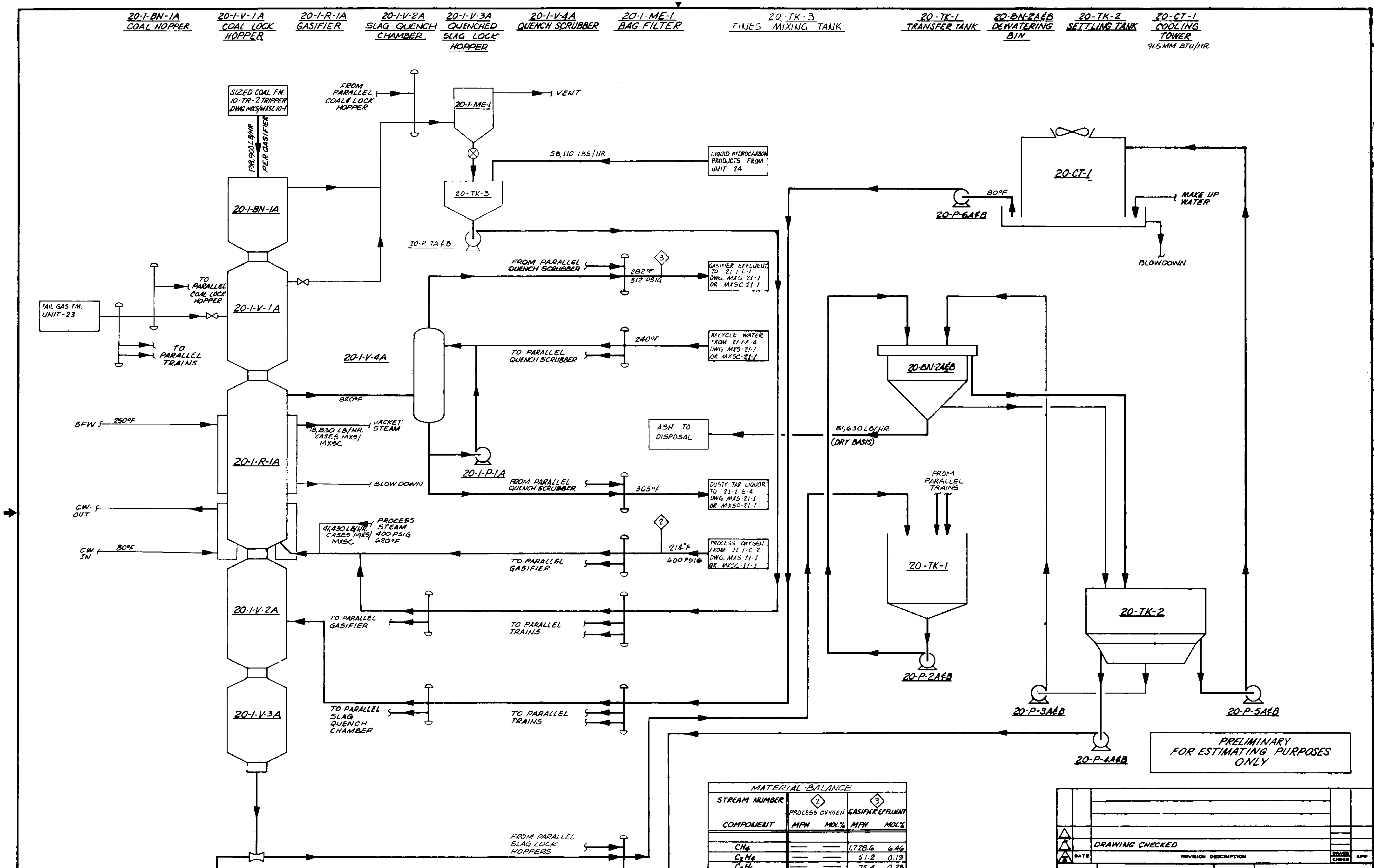
The ash slurry from the transfer tank is pumped to dewatering bins 20-BN-2A&B to produce ash ready for disposal.

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

Equipment Notes

The coal feed, coal distribution, stirrer and gas quench technology associated with moving bed gasifiers is commercially proven for noncaking coals via the Lurgi experience. Operation of this gasifier on Illinois #6 coal (the basis for the study) is yet to be demonstrated. The slagging technology has been under development at the Westfield Development Centre of British Gas Corporation. A slagging bottom gasifier has been operating at Westfield for the past three years supported by a consortium comprising fourteen gas companies and the Electric Power Research Institute. The success of this pilot program has resulted in a DOE contract for the design of a 60 MM SCFD demonstration plant in Ohio for SNG production based on the slagging gasifier technology.

The ash slurry system is a commercially available system.



NOTES:

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

MATERIAL BALANCE

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

DRAWING CHECKED

FLUOR

R. CANNON
R. RICE
BARGER
T.H. TAN

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

ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CALIFORNIA

SCALE NONE

MXS/MXSC-20-1

GAS COOLING

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

Gasifier quench scrubber effluent is cooled to approximately 264°F in a shell and tube exchanger, 21-1-E-1, by cold product gas from 22-1-V-4 in the Acid Gas Removal Unit. The condensate from the crude gas is separated in a knockout drum, 21-1-V-1.

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

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

Condensate from 21-1-V-1 is combined with cooled dusty tar liquor from a shell and tube exchanger, 21-1-E-4. The dusty tar liquor is then cooled by heat exchange against recycle water in 21-1-E-5. It is further cooled to 220°F by a cold demineralized water and condensate stream in exchangers 21-1-E-6 and 21-1-E-7, and flows to the Process Condensate Treating Unit for further processing.

The recycle water is further heated to 240°F against dusty tar liquor stream from the quench scrubber in exchanger 21-1-E-4 and flows to the gasification unit.

Equipment Notes

All the equipment is commercially available.

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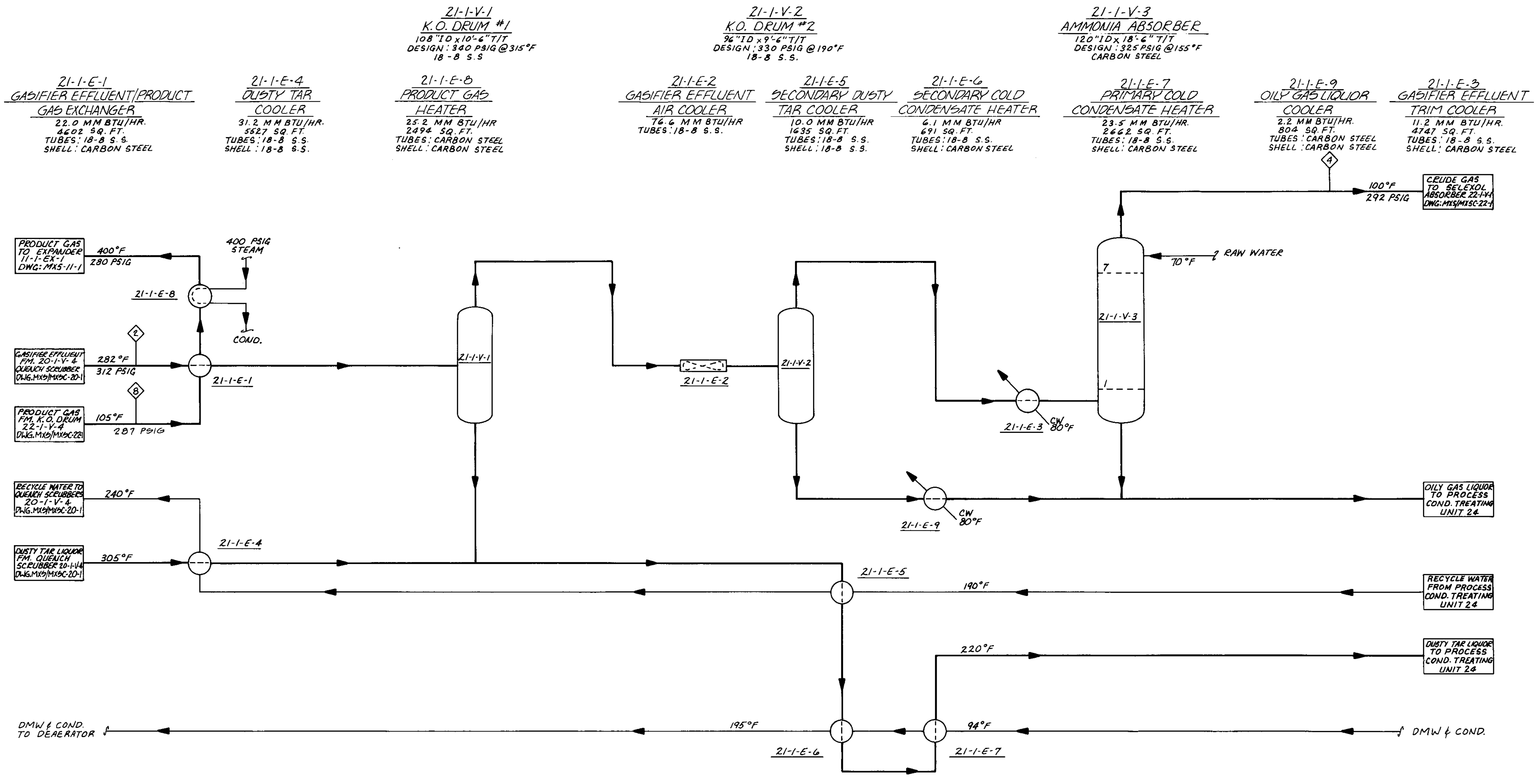

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MATERIAL BALANCE						
STREAM NUMBER	2		8		4	
	GASIFIER EFFLUENT		PRODUCT GAS		CRUDE GAS	
COMPONENT	MPH	MOL%	MPH	MOL%	MPH	MOL%
CH_4	1,728.6	6.46	1,702.5	7.78	1,728.6	7.69
C_2H_4	51.2	0.19	47.8	0.22	51.2	0.23
C_2H_6	75.4	0.28	70.5	0.32	75.4	0.33
H_2	6,773.7	25.32	6,755.0	30.87	6,773.7	30.13
CO	12,928.9	48.33	12,851.5	58.73	12,928.9	57.52
CO_2	431.0	1.61	234.1	1.07	431.0	1.92
H_2S	306.9	1.15	34.6	0.16	306.9	1.37
COS	13.4	0.05	8.6	0.04	13.4	0.06
NH_3	201.9	0.74	—	—	—	—
N_2	99.5	0.37	99.1	0.45	99.5	0.44
H_2O	4,094.9	15.31	78.8	0.26	69.7	0.31
TOTAL MPH	26,705.4	100.00	21,882.5	100.00	22,478.3	100.00
T.O.P.N.& OTHER LB/HR	4,840		—		—	
TOTAL LB/HR	522,310		420,560		441,510	
MOL. WT.	19.52		19.22		19.64	

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

REV. NO.	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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

FLUOR

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CHECKED BY R. RICE
SUPERVISOR R. RICE
RELEASE DATE
SUPERVISING ENGR. S. FEERRAR
INITIALS
PROJECT ENGR.
APP. DATE
CLIENT
APP. DATE

PROCESS FLOW DIAGRAM
GAS COOLING
MOVING BED SLAGGING-OXYGEN BLOWN
ELECTRIC POWER RESEARCH INSTITUTE PALO ALTO, CA.

SCALE: NONE
DRAWING NUMBER: MXS-21-1
REVISION: 1

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AAE3844-SMX-21-1 (S.S.E)

ACID GAS REMOVAL

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

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

The cooled ammonia-free gas flows through an acid gas absorber, 22-1-V-1, where it contacts Selexol® solvent countercurrently in a packed tower. The treated gas from the top of the absorber flows through knockout drum 22-1-V-4 back to the upstream gas cooling unit (Flow Diagram: MXSC-21-1) where it exchanges heat with the hot gasifier effluent in exchanger 21-1-E-1. Product gas is then heated to 400°F in a product gas heater, 21-1-E-8, by exchanging heat against 400 psig condensing steam and flows to the oxidant feed section (Flow Diagram MXS-11-1). The rich solvent from the bottom of the absorber is let down through a hydraulic turbine, 22-1-HT-1, which supplies a portion of the power required by the lean solution pump, 22-1-P-1. It then flows to a flash drum, 22-1-V-2, where most of the dissolved hydrocarbon gases in the solvent flash off. Approximately 98% of the dissolved H_2S and most of the dissolved COS are retained in the solvent because of their selective absorption in the Selexol® solvent. The flash gas flows to the sulfur plant.

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

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

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

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

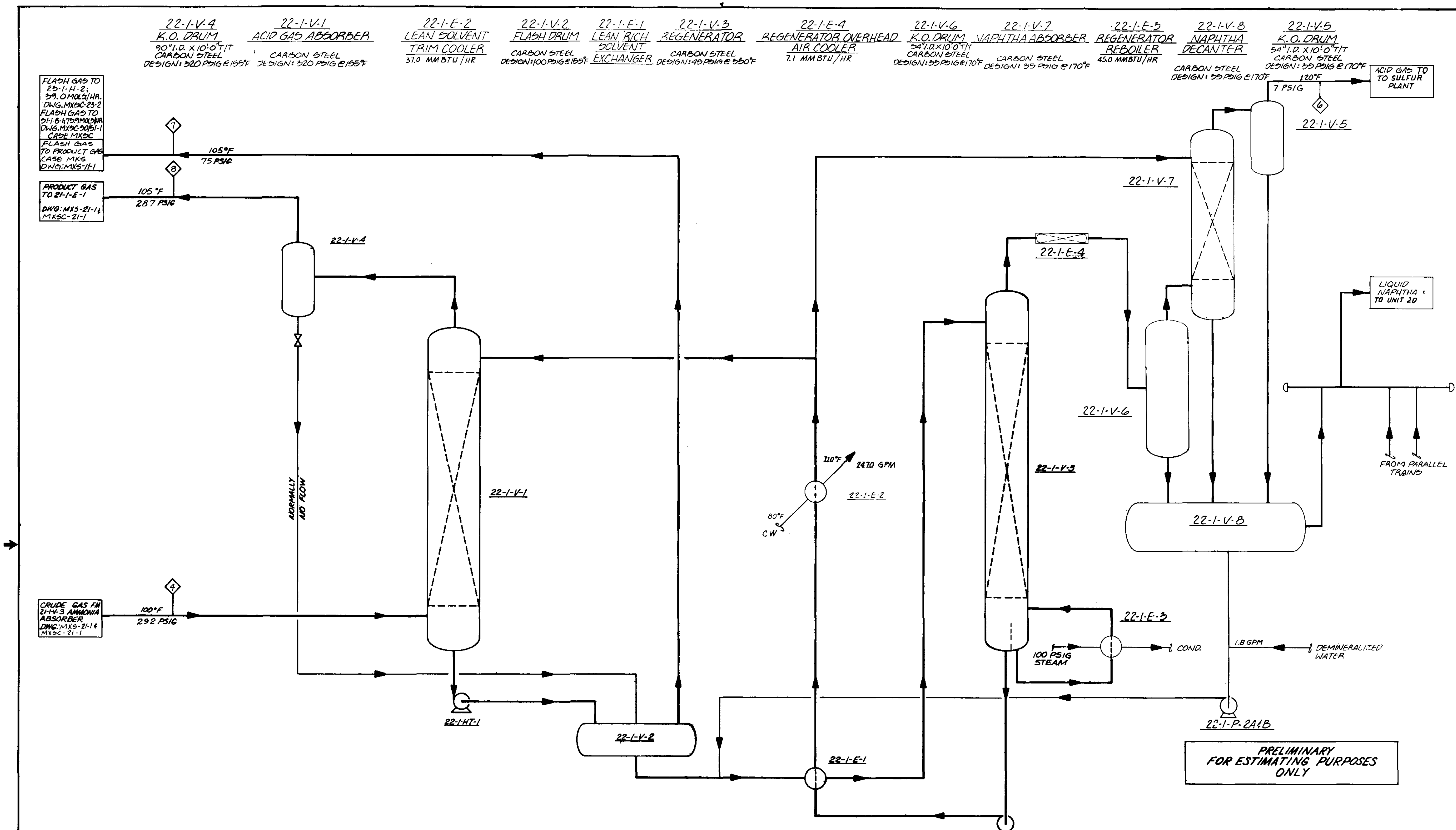
A small quantity of demineralized water is added to the Selexol® solvent at the suction of 22-1-P-2A&B to maintain the water balance in the absorption system.

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

Equipment Notes

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

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



MATERIAL BALANCE

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

NOTES:

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

	DATE	REVISION DESCRIPTION	DRAWN / CHECK	APP

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FLUOR

DRAWN BY: BOB MCAULEY
CHECKED BY: R. RICE
J. BARBER
T.H. TAN

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

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

MXS/MXSC-SS-1

SULFUR RECOVERY AND TAIL GAS TREATING

The processing schemes used for these units are similar to ones described in the Fuel Gas Report.[1] Refer to Cases MA/MX and Process Flow Diagrams MA/MX - FA/FX-23-1 and MA/MX - FA/FX-23-2 in this report for a detailed process description of these units.

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

---

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

PROCESS CONDENSATE TREATING

The processing schemes for these units are similar to the ones described in the Fuel Gas Report.[1] Refer to Cases MA/MX and Process Flow Diagrams MA/MX-24-1 and MA/MX-24-2 in this report for a detailed process description of these units.

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

An anhydrous ammonia stream of 123.5 tons/day suitable for fertilizer and commercial uses is recovered from dephenolized process condensate in an ammonia recovery unit using U.S. Steel's PHOSAM-W process.

---

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

STEAM, BOILER FEEDWATER AND CONDENSATE

Process Flow Diagram, MXS-30-1, schematically represents steam, boiler feedwater and condensate system.

The steam system operates at four pressure levels:

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

Major steam generation is carried out in three boilers operating at 1450 psig. Boiler feedwater from a deaerator is pumped to boilers 30-B-1 through pumps 30-P-2 A, B & C to generate superheated 1450 psig, 900°F steam. Each boiler is designed with 10% overcapacity. There is a fourth standby boiler.

All high pressure steam is used to drive the air feed compressors and the oxidant feed compressors. The air compressors use condensing turbines 11-1-T-1. The turbines (11-1-T-2) for oxidant feed compressors are back pressure turbines exhausting at 400 psig.

The medium pressure steam level at 400 psig is primarily supplied by the back pressure turbines 11-1-T-2. The balance comes from process waste heat genration, jacket steam from the gasifiers, and steam generated in the waste heat boilers in the Sulfur Recovery Unit.

The medium pressure steam header is controlled by feeding the excess steam to a power recovery turbine 31-T-1. In addition to recovering power, the turbogenerator acts as the balancing wheel to control the steam system. Swings in steam demand at the different levels are reflected in the power output of the turbogenerator. About 36% of the total medium pressure steam is supplied to the gasifiers. The balance is used for product gas heating (21-E-8) and for pump steam turbine drivers which exhaust to the 50 psig steam level.

The 100 psig steam is mainly used in the Acid Gas Removal and Process Condensate Treating Units. The steam is supplied by extraction from the turbine, 31-T-1.

The 50 psig level is supplied by steam from the small backpressure turbine drivers, Sulfur Recovery Unit, and blowdown flash steam. Blowdown from the high and medium pressure steam generators is combined in a blowdown drum 30-V-3, which vents flash steam to the 50 psig header.

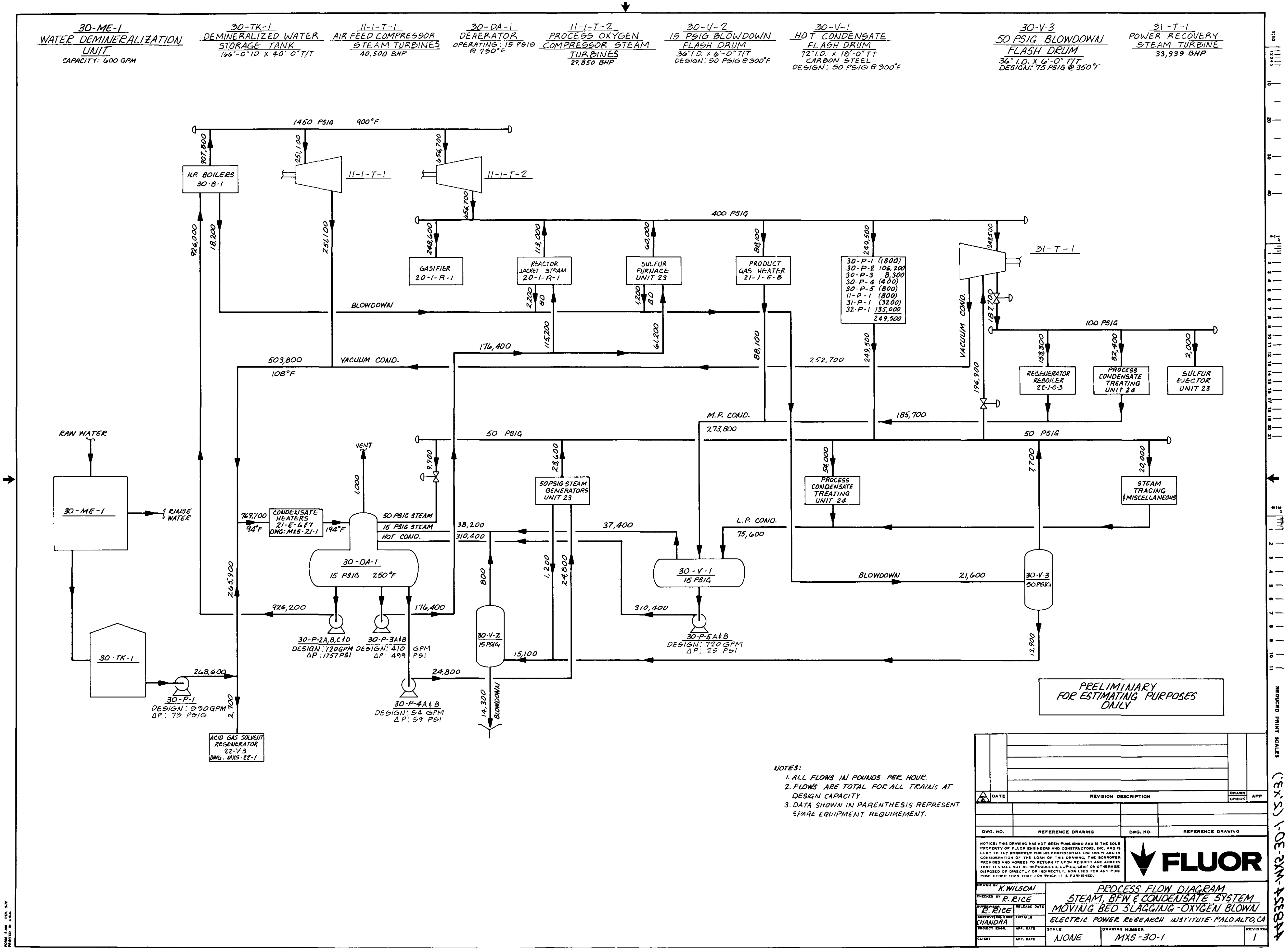
Critical pumps such as boiler feedwater and condensate pumps are steam turbine driven. Other pump services considered important but not critical use electrical motors in normal operation and steam driven spares for emergency service. All turbine drivers operate from 400 psig to 50 psig backpressure.

The steam condensate from the 400 psig, 100 psig and 50 psig steam users is flashed at 15 psig in a hot condensate flash drum, 30-V-1. The flashed steam is vented to the deaerator, 30-DA-1. The condensate from 30-V-1 is pumped to the deaerator trayed section.

The blowdown from 30-V-3 and the 50 psig steam generators (23-E-1 and 2) in the Sufur Recovery Unit is flashed in a 15 psig blowdown flash drum, 30-V-2. The flashed steam is vented to the deaerator. The net blowdown losses from the system are estimated at 14,030 lbs/hr.

Raw water is treated in a semiautomatic, resin bed demineralization unit, 30-ME-1, to produce demineralized water suitable for a 1500 psig boiler system. Storage equivalent to 24 hours of demineralized water production is provided. The demineralized water requirement is estimated at approximately 536 gpm. Some demineralized water is also used to satisfy process requirements.

The demineralized water for the steam system is combined with vacuum condensate returned from the surface condenser. The combined stream is heated to 194°F by heat exchange in the Acid Gas Treating Unit. The heated stream is deaerated in a tray type deaerator operating at 15 psig. The deaerator provides 10 minutes storage.


30-ME-1
WATER DEMINERALIZATION UNIT
CAPACITY: 600 GPM
30-TK-1
DEMINERALIZED WATER STORAGE TANK
166'-0" I.D. x 40'-0" T/T
11-1-T-1
AIR FEED COMPRESSOR STEAM TURBINES
40,500 BHP
30-DA-1
DEAERATOR
OPERATING: 15 PSIG @ 250°F
11-1-T-2
PROCESS OXYGEN COMPRESSOR STEAM TURBINES
29,850 BHP
30-V-2
15 PSIG BLOWDOWN FLASH DRUM
36" I.D. x 6'-0" T/T
DESIGN: 50 PSIG @ 300°F
30-V-1
HOT CONDENSATE FLASH DRUM
72" I.D. x 18'-0" TT
CARBON STEEL
DESIGN: 50 PSIG @ 300°F
30-V-3
50 PSIG BLOWDOWN FLASH DRUM
36" I.D. x 6'-0" T/T
DESIGN: 75 PSIG @ 350°F
31-T-1
POWER RECOVERY STEAM TURBINE
33,939 BHP
1450 PSIG 900°F
907,800
251,100
656,700
H.P. BOILERS 30-B-1
11-1-T-1
11-1-T-2
656,700
400 PSIG
926,000
18,200
251,100
248,600
113,000
60,000
88,100
249,500
248,500
GASIFIER 20-1-R-1
REACTOR JACKET STEAM 20-1-R-1
SULFUR FURNACE UNIT 23
PRODUCT GAS HEATER 21-1-E-8
30-P-1 (1800)
30-P-2 106,200
30-P-3 8,300
30-P-4 (400)
30-P-5 (800)
11-P-1 (800)
31-P-1 (3200)
32-P-1 135,000
249,500
31-T-1
BLOWDOWN
2,200
80
1,200
80
187,700
VACUUM COND.
100 PSIG
176,400
115,200
61,200
88,100
249,500
503,800
108°F
VACUUM COND.
252,700
196,900
153,300
32,400
2,000
REGENERATOR REBOILER 22-1-E-3
PROCESS CONDENSATE TREATING UNIT 24
SULFUR EJECTOR UNIT 23
M.P. COND.
273,800
185,700
RAW WATER
50 PSIG
50 PSIG
VENT
9,900
23,600
34,000
7,700
20,000
1,000
50 PSIG STEAM GENERATORS UNIT 23
PROCESS CONDENSATE TREATING UNIT 24
STEAM TRACING & MISCELLANEOUS
30-ME-1
RINSE WATER
769,700
94°F
CONDENSATE HEATERS 21-E-6&7 DWG: MX5-21-1
194°F
50 PSIG STEAM
15 PSIG STEAM
HOT COND.
38,200
310,400
37,400
L.P. COND.
75,600
30-DA-1
15 PSIG 250°F
265,900
1,200
800
24,800
30-V-1
15 PSIG
BLOWDOWN
21,600
30-V-3
50 PSIG
926,200
176,400
310,400
30-P-2A,B,C&D
DESIGN: 720 GPM
ΔP: 1757 PSI
30-P-3A&B
DESIGN: 410 GPM
ΔP: 499 PSI
30-V-2
15 PSIG
15,100
30-P-5A&B
DESIGN: 720 GPM
ΔP: 25 PSI
13,900
30-TK-1
268,600
30-P-1
DESIGN: 590 GPM
ΔP: 73 PSIG
2,700
24,800
30-P-4A&B
DESIGN: 54 GPM
ΔP: 59 PSI
14,300
BLOWDOWN
ACID GAS SOLVENT REGENERATOR 22-V-3 DWG. MX5-22-1
PRELIMINARY FOR ESTIMATING PURPOSES ONLY
NOTES:
1. ALL FLOWS IN POUNDS PER HOUR.
2. FLOWS ARE TOTAL FOR ALL TRAINS AT DESIGN CAPACITY.
3. DATA SHOWN IN PARENTHESIS REPRESENT SPARE EQUIPMENT REQUIREMENT.
DATE
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STEAM, BFW & CONDENSATE SYSTEM
MOVING BED SLAGGING-OXYGEN BLOWN
ELECTRIC POWER RESEARCH INSTITUTE - PALO ALTO, CA
SCALE NONE
DRAWING NUMBER MX5-30-1
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REDUCED PRINT SCALES

PROCESS DISCUSSION - CASE MXS

The table below summarizes pertinent results.

TABLE MS-2

SUMMARY OF RESULTS - CASE MXS

GASIFIER	
Oxygen, lb/lb maf coal (1)	0.534
Oxidant Temperature, °F	214
Steam, lb/lb maf coal	0.34
Exit Pressure, psig	320
Exit Temperature, °F	820
Coal Carbon Converted to CH_4, atom %	11.2
Coal Carbon Converted to C_2H_4 & C_2H_6, atom %	1.64
Steam Decomposition, % (2)	90.0
Gas H_2:CO, mole ratio	0.524
Gas CO:CO_2, mole ratio	30
Gasifier Cold Gas Efficiency, % of coal HHV (3)	96.37
OVERALL PLANT	
System Cold Efficiency, % of coal HHV (4)	81.3
Fuel Gas Sensible Heat, % of coal HHV	0.34
Net Fuel Gas, mscf/ton maf coal (1)	60.0
Fuel Gas HHV, Btu/scf (1)	379
By-Product Liquid Hydrocarbons, #/ton maf coal	0
Power from Steam Cycle, kWh/ton maf coal	67.8
Power from Air Separation Plant, kWh/ton maf coal	2
Power Consumed, kWh/ton maf coal	34.2
By-Product Ammonia, #/ton maf coal	28.70
Process/Steam Make-up Water, mgal/ton maf coal	0.16
Cooling Tower Heat Rejection, % of coal HHV	8.6
Cooling Water Circulation, mgal/ton maf coal	14.4
Cooling Tower Make-up Water, mgal/ton maf coal	0.20

NOTES

(1) Dry basis

(2) $100.0 - \frac{(H_2O \text{ in crude gas} - H_2O \text{ in coal feed})\ (100.0)}{(\text{Steam} + H_2O \text{ in air} + H_2O \text{ in transport gas})}$

(3) (HHV of crude gas including tars, oils and phenols) (100)/(HHV of coal)

(4) $\frac{[(\text{HHV of fuel gas}) + (\text{Net Power})(9000\ \text{Btu/kWh})]100}{\text{HHV of coal}}$

Gasifier Material Balance

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

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

This gasifier, based on the British Gas Corporation's slagging gasifier development, is unique from two points of view. First, 84.94% weight of the carbon in the coal is converted to CO resulting in a fuel gas containing very little CO_2, thus preserving a major fraction of the chemical heat of the coal in fuel gas. The steam to oxygen mole ratio for this case is approximately 1:12. The steam decomposition is 90%. The high utilization of steam in the process allows for smaller equipment in the downstream units.

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

TABLE MS-3

MATERIAL BALANCE - CASE MXS

FEEDS

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

EFFLUENTS

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

NOTES:

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

Acid Gas Removal

A distinct feature of Case MXS is the production of a smaller quantity of CO_2 in the gasifier effluent compared to other cases reported previously [1]. Selective removal of H_2S over CO_2 is therefore not as important for this case. Allied Chemical's Selexol® process was selected for Case MXS to have a common basis for this section with other cases reported earlier[1].

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

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

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

Process Energy Balances

Table MS-4 presents an overall process energy balance for this case at 100% capacity operation. The boundary for the balance encompasses the entire plant.

---

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

TABLE MS-4

ENERGY BALANCE: CASE MXS

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

	MM Btu/hr					
	HHV	SENSIBLE	LATENT	RADIATION	POWER	TOTAL
HEAT IN						
Coal	10,196	5				10,201
Oxidant Compressor Suction Air		12	30			42
Boiler Combustion Air		6	15			21
Demineralized Water		3				3
Auxiliary Power Inputs					42	42
TOTAL	10,196	26	45	0	42	10,309
HEAT OUT						
Net Product Gas	8,171	35	4			8,210
Ash/Slag	23	100				123
Gasifier Heat Loss				21		21
Gasifier Cooling		41				41
Sulfur Product	106	1				107
Ammonia Product	100					100
Generated Power					83	83
Power Turbine Surface Condenser			227(1)			227
Air Compressor Surface Condenser			237(1)			237
Boiler Stack Losses		78	98			176
Gasifier Effluent Cooling		46	224			270
Oxidant Compressors Cooling		262(1)				262
N_2 Vent from Oxygen Plant		12			2	14
Selexol Solvent Cooling		111				111
Selexol Regeneration Overhead Cooling		2	19			21
Process Condensate Cooling		7				7
Steam Heat and Power Losses		56	3			59
Tail Gas Unit Cooling		17				17
Process Condensate Treating Unit		171				171
Spent Tail Gas	22	2	7			31
Waste Water Effluent		14				14
TOTAL	8,422	955	819	21	85	10,302

$$\frac{\text{Input - Output}}{\text{Input}} = 0.07\%$$

(1) Latent plus sensible heat.

Energy content of the stream crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F, and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kWh. The energy balance closes to less than 0.5 percent. Approximations were used for some units and for calculating some heat loads.

Data from Table MS-4 is shown in MM Btu/hr and as percent of coal higher heating value in Table MS-5. The gasifier and the system cold efficiencies are 96.37 and 81.5 percent respectively. These figures are slightly different from the data reported in an earlier EPRI report[1] as the material balance and energy balance have been recalculated. The changes are all minor.

As shown in the tables, Case MXS produces approximately 8,206 MM Btu/hr (HHV plus sensible heat) in the form of product gas from a coal charge of 10,196 MM Btu/hr.

This case also produces by-product electric power. The by-product electric power generation in Case MXS somewhat differs from other moving bed cases (Cases MA and MX) reported earlier.[2] The bottoming cycles using isobutane in a Rankine Cycle to generate electric power provided in Cases MA and MX, were not included in the plant design of Case MXS. A subsequent Fluor study indicated that bottoming cycles are not economically justified in coal gasification plants. The bottoming cycle was therefore eliminated in Case MXS. The net power (including several users and power related inefficiencies not within the boundary of the process balances) to be credited is 12770 kW. If by-product power is credited at its theoretical conversion to heat energy, 3,413 Btu/kWh, the plant thermal efficiency represented by product gas and power is 80.6 percent. Electric power, however, represents a relatively greater investment in energy than the theoretical conversion because of inherent inefficiencies in any power generation scheme. If by-product power is credited at 9,000 Btu/kWh heat rate, the system thermal efficiency improves to 81.3 percent.

---

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

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

TABLE MS-5

ENERGY BALANCE AS PERCENT OF COAL HHV - CASE MXS

	MM Btu/hr	Percent
IN		
Coal HHV	10,196	100.00
OUT		
Product Gas HHV	8,171	80.14
Product Gas Sensible and Latent	39	0.38
Sulfur Product Heat	107	1.05
Ammonia Product HHV	100	0.98
Spent Tail Gas HHV	22	0.22
Selexol Cooling Losses (Sensible and Latent)	132	1.30
Oxidant Interstage Cooling	262	2.57
Ash/Slag Heat	123	1.21
Boiler Stack Gases	176	1.73
Rejected at Surface Condensers	464	4.55
Other Sensible Losses	340	3.33
Other Latent Losses	189	1.85
Gasifier Heat Losses	21	0.20
Net Power	41	0.40
	10,187	99.91

Substantial amounts of liquid hydrocarbon by-products are produced in the form of naphthas, oils, tars and phenols. These by-products are recovered, and this recovery represents an energy penalty due to the process heat required for the tar and phenol units (171 MM Btu/hr). The liquid hydrocarbon by-products are recycled back to the gasifiers' combustion zone through tuyeres. Therefore, no net hydrocarbon product production results.

Fired boilers are required to generate steam. As a consequence, approximately 176 MM Btu/hr (1.73 percent of coal HHV) is lost in the stack. The other major heat loss is at the surface condensers of the air compressor turbine and the power recovery turbine. The surface condensers' duty represents approximately 464 MM Btu/hr or 4.55 percent of coal HHV.

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

ECONOMICS - CASE MXS

Important economic results are summarized below.

TABLE MS-6

SUMMARY OF ECONOMICS - CASE MXS

PRODUCTION AT DESIGN CAPACITY	
Net Fuel Gas, MM Btu/day (1)	196,937
By-product Power, MW	12.77
TOTAL CAPITAL (2)	
Total Capital @ 70% Operating Factor, \$1,000	391,611
Total Capital @ 90% Operating Factor, \$1,000	391,747
AVERAGE COST OF SERVICES (2)	
Annual Cost @ 70% Operating Factor, \$1,000/yr	140,369
Per Unit @ 70% Operating Factor, \$/MM Btu	2.79
Annual Cost @ 90% Operating Factor, \$1,000/yr	157,768
Per Unit @ 90% Operating Factor, \$/MM Btu	2.44

NOTES

(1) Heating value plus sensible heat at 100% operating load factor.
(2) Mid-1976 dollars and \$1/MM Btu coal.

Tables MS-7 and MS-8 give detailed breakdowns of plant investment, capital charges and working capital for both cases at 70 and 90 percent operating factors, respectively. Plant investment is the same at both operating load factors. The accuracy of plant investments 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. Therefore, caution must be exercised in comparing this case with cases representing other gasification technologies reported earlier.[1]

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

Table MS-9 summarizes cost of services at 70 and 90 percent operating load factors. Costs are computed in accordance with criteria given by EPRI (Criteria section). They are presented as averages for the plant. Figure MS-1 shows the variation of cost of services with cost of coal.

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

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

Operators	21
Foremen	2
Lab and Instrument Technicians	4

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

Operating charges constitute 56.3 and 61.1 percent of costs of services at operating load factors of 70 and 90 percent respectively. Coal is the largest single operating charge. Based on coal cost of $1/MM Btu, yearly coal costs represent approximately 79 and 83 percent of the operating charges respectively at 70 and 90 percent operating load factors. The relationships as percentages are summarized below:

Operating Load Factor	70%	90%
Coal as % of Operating Charges	79.3	83.7
Coal as % of Total Cost of Services	44.7	51.1
Operating Charges as % of Total Cost of Services	56.3	61.1
Capital Charges as % of Total Cost of Services	44.7	38.9

FIGURE MS-1

COST OF SERVICES VS. COST OF COAL FOR CASE MXS

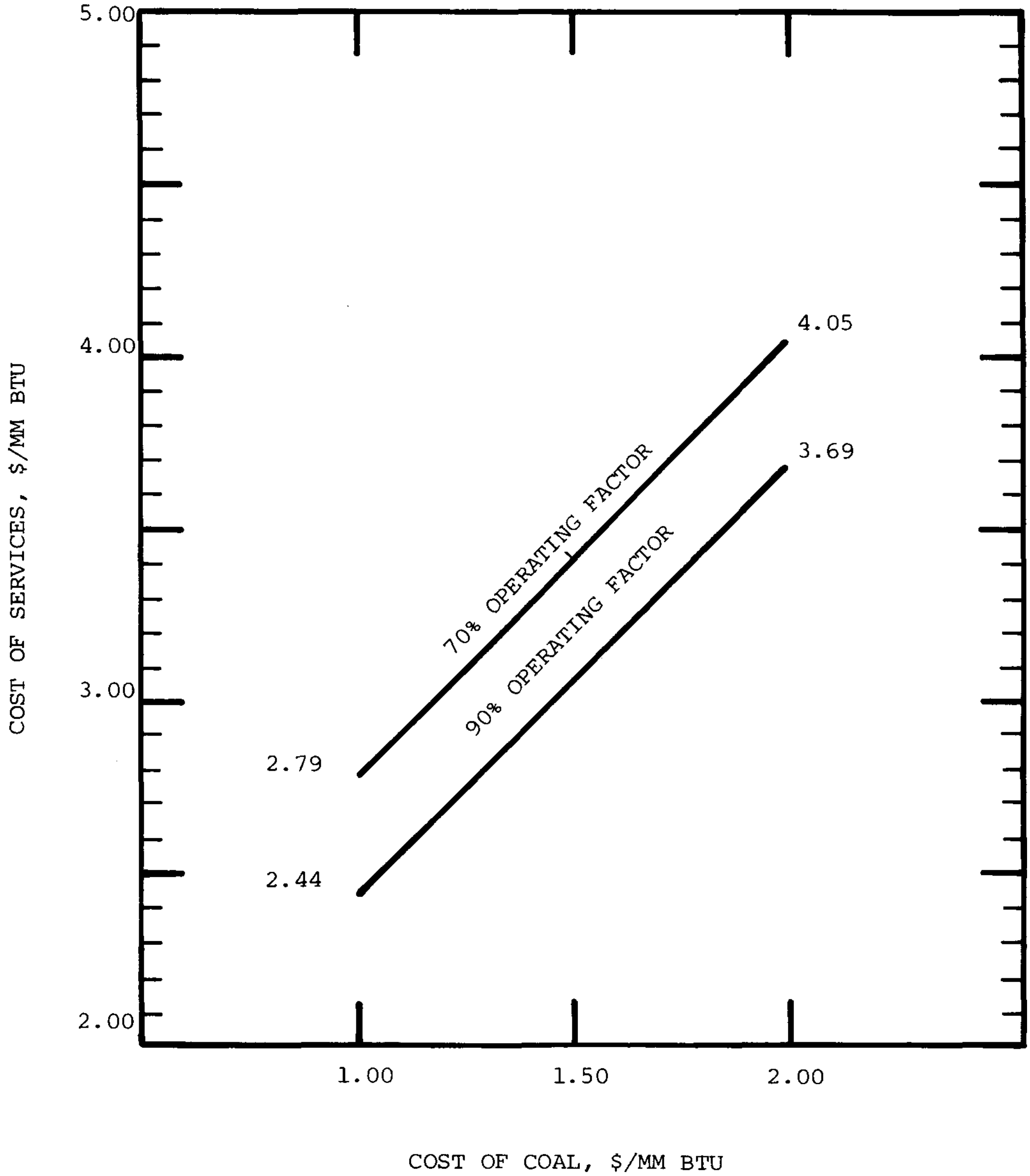


TABLE MS-7

CAPITAL INVESTMENT AT 70% LOAD FACTOR AND $1/MM BTU COAL CASE MXS

PLANT INVESTMENT	$ 1000 (1)	$/MM Btu/hr(2)	Percent
Coal Preparation	14,905	1,820	5.69
Oxidant Feed System	61,540	7,500	23.51
Gasification & Ash Handling	49,628	6,050	18.96
Gas Cooling	8,430	1,030	3.22
Acid Gas Removal & Sulfur Recovery	23,498	2,860	8.98
Process Condensate Treating	31,169	3,800	11.91
Steam Condensate & BFW	26,210	3,190	10.01
Power Recovery	7,065	860	2.70
Support Facilities	39,276	4,790	15.01
Subtotal	261,721	31,900	100.00
Contingency	41,287	5,030	
TOTAL PLANT INVESTMENT	303,008	36,930	
ILLINOIS SALES TAX	7,035	860	
CAPITAL CHARGES			
Preproduction Costs	20,417	2,490	
Paid-Up Royalties	1,515	180	
Initial Catalyst & Chemical Charges	810	100	
Construction Loan Interest	37,846	4,610	
TOTAL CAPITAL CHARGES	60,588	7,380	
DEPRECIABLE CHARGES	370,631	45,180	
WORKING CAPITAL	20,865	2,540	
TOTAL CAPITAL	391,496	47,710	

NOTES

(1) Mid-1976 dollars.
(2) Based on 100% operating load factor

TABLE MS-8

CAPITAL INVESTMENT AT 90% LOAD FACTOR AND $1/MM BTU COAL CASE MXS

PLANT INVESTMENT	$ 1000 (1)	$/MM Btu/hr(2)	Percent
Coal Preparation	14,905	1,820	5.69
Oxidant Feed System	61,540	7,500	23.51
Gasification & Ash Handling	49,628	6,050	18.96
Gas Cooling	8,430	1,030	3.22
Acid Gas Removal & Sulfur Recovery	23,498	2,860	8.98
Process Condensate Treating	31,169	3,800	11.91
Steam Condensate & BFW	26,210	3,190	10.01
Power Recovery	7,065	860	2.70
Support Facilities	39,276	4,790	15.01
Subtotal	261,721	31,900	100.00
Contingency	41,287	5,030	
TOTAL PLANT INVESTMENT	303,008	36,930	
ILLINOIS SALES TAX	7,035	860	
CAPITAL CHARGES			
Preproduction Costs	20,507	2,500	
Paid-Up Royalties	1,515	180	
Initial Catalyst & Chemical Charges	810	100	
Construction Loan Interest	37,846	4,610	
TOTAL CAPITAL CHARGES	60,678	7,390	
DEPRECIABLE CHARGES	370,721	45,180	
WORKING CAPITAL	21,001	2,560	
TOTAL CAPITAL	391,722	47,740	

NOTES

(1) Mid-1976 dollars.
(2) Based on 100% operating load factor

TABLE MS-9

COST OF SERVICES AT $1/MM BTU COAL
CASE MXS

OPERATING FACTOR	70%	90%
NET PRODUCTION (1)		
Fuel Gas, MM Btu/day	137,856	177,243
By-product Electric Power, kW	10,533	13,542
By-product Sulfur, ST/day	221.2	284.4
By-product Ammonia, ST/day	86.4	111.1
OPERATING CHARGES, $1000/YR		
Coal	62,522	80,385
Operating Labor	2,602	2,602
Catalysts & Chemicals	199	256
Utilities	319	410
Maintenance, Labor	3,147	3,147
Maintenance, Materials	3,304	4,248
Administrative and Support Labor	1,725	1,725
General and Administrative Expenses	3,449	3,449
Ash Disposal	251	322
Property Tax and Insurance	7,595	7,595
By-product Electric Power	(2,662)	(3,423)
By-product Ammonia	(3,155)	(4,057)
By-product Sulfur	0	0
TOTAL OPERATING CHARGES, $1000/YR	79,296	96,659
CAPITAL CHARGES, $1000/YR		
TOTAL CAPITAL CHARGES	61,073	61,109
COST OF SERVICES		
Total $1000/Year	140,369	157,768
Per Unit Production, $/MM Btu	2.79	2.44

NOTES

(1) Production rates given are on an "operating day" basis.

Besides the differences in gasification technology, Case MXS somewhat differs in plant design from other moving bed cases (Cases MA and MX) reported earlier.[1] Bottoming cycles using isobutane in a Rankine cycle to generate by-product power were incorporated in the plant design of Cases MA and MX. A subsequent Fluor study indicated that bottoming cycles are not economically justified in coal gasification plants. The bottoming cycle was therefore eliminated in Case MXS. The magnitude of investment in facilities associated with the cycle and the value of by-product power are, however, too small to materially affect the comparison of other moving bed cases reported previously with Case MXS.

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

APPENDIX A

AREA AND UNIT NUMBERING

The 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.

TABLE A-1

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 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
		31	Power Recovery
		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

APPENDIX B

ECONOMICS (REVISED TO 1976 BASIS)
MOVING BED CASES

Important economic results are summarized below.

TABLE M-6

SUMMARY OF ECONOMICS - CASES MA & MX - LURGI

	Case MA Air	Case MX Oxygen
PRODUCTION AT DESIGN CAPACITY		
Net Fuel Gas, MM Btu/day (1)	134,376	131,880
Liquid Hydrocarbons, MM Btu/day (2)	21,912	21,912
By-Product Power, MW	76.4	63.7
TOTAL CAPITAL (3) (4)		
Total Capital @ 70% Operating Factor, $1,000	582,652	704,393
Total Capital @ 90% Operating Factor, $1,000	583,731	704,832
AVERAGE COST OF SERVICES (3)		
Annual Cost @ 70% Operating Factor, $1,000/yr	169,077	199,722
Per Unit @ 70% Operating Factor, $/MM Btu	4.23	5.08
Annual Cost @ 90% Operating Factor, $1,000/yr	183,445	215,143
Per Unit @ 90% Operating Factor, $/MM Btu	3.57	4.26

NOTES
(1) Heating value plus sensible heat
(2) Higher Heating value only
(3) Mid-1976 dollars and $1/MM Btu coal
(4) Capital includes the cost of generating equipment required to produce byproduct power.

Total capital investments for Case MX are 18 percent higher and costs of services are 19 to 20 percent higher than for Case MA.

Tables M-7 and M-8 give detailed breakdowns of plant investment, capital charges and working capital for both cases at 70 and 90 percent operating factors, respectively. Plant investment for each case is the same at both operating load factors.

The plant investment by unit in Case MX is consistently as high or higher than in Case MA. There is only one exception - the acid gas removal unit.

The contingency shown under plant investment is an allowance to account for developments in the state of the art. Historically, as a technology develops from the conceptual stage to commercial reality, a variety of technical problems which were not considered in the early stages 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 purpose of the contingency as used in this report is to account for these costs. It does not cover escalation or estimating inaccuracies. A total plant contingency is arrived at by applying a separate contingency to individual process units based on their state of development and accumulating the results.

Table M-9 summarizes costs of services for both cases based upon coal charged at $1.00/MM Btu HHV. Figure M-1 shows how costs of services change with coal cost. Costs of services are computed in accordance with criteria set out in the Criteria Section.

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

	Case MA	Case MX
Operators	32	36
Day Laborers (one shift)	18	18
Lab Technician	1	1
Foremen	4	4
Superintendent	1	1

Case MX has more operators because of the air separation unit and additional compressors in the oxidant feed system.

Catalyst and chemical costs are primarily for chemicals consumed in the demineralizer, cooling tower and boiler feedwater treating. There are some 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 as well. Utility costs are for raw water make-up to the plant.

Operating charges are higher in Case MX. Nearly all aspects of operations contribute to the higher costs except coal cost, which is the same.

Operating charges constitute about 50 percent of the costs of services in both the air and oxygen cases. By far the largest single operating charge is coal; yearly coal costs are approximately the same as capital charges. The relationship as percentages is summarized below:

	Case MA		Case MX	
	70%	90%	70%	90%
Coal as % of Operating Charges	78.7	87.0	69.6	76.4
Coal As % of Total Cost of Services	37.0	43.8	31.3	37.4
Operating Charges as % of Total Cost of Services	46.2	50.4	45.0	48.9
Capital Charges as % of Total Cost of Services	53.8	49.6	55.0	51.1

As the unit cost for coal increases above $1.00/MM Btu HHV, coal costs as percentages become greater.

A substantial portion of the energy produced from these plants is in the form of byproduct liquid hydrocarbons - naphtha, tar, oil, phenol and others. On an energy basis these hydrocarbons are equivalent to 16.3% of product gas in Case MA and 16.6 percent in Case MX. In computing cost of services on a MM Btu basis, the energy value of the liquid hydrocarbons is added to that of the product fuel gas. This assumes that on a Btu basis, the liquid hydrocarbons are as valuable a fuel as the clean fuel gas. Such an assumption is quite optimistic as the liquid hydrocarbons will be substantially higher in sulfur and particulate content than the product fuel gas and will probably have to undergo further treatment (and, therefore, incur added costs) before being suitable for use as fuel products. Therefore the costs of gas presented for Cases MA and MX are probably on the optimistic side.

The plant designs for Cases MA and MX incorporate bottoming cycles using isobutane in a Rankine cycle to generate byproduct electric power. A subsequent Fluor study indicated that bottoming cycles are not economically justified in coal gasification plants. The bottoming cycles, however, have not been removed from these cases since the magnitude of investment in facilities associated with the cycle and the value of byproduct power are too small to materially affect results presented here.

TABLE M-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $1/MM BTU COAL - CASES MA AND MX

	CASE MA (Lurgi-Air)			CASE MX (Lurgi-Oxygen)		
	$1,000 (1)	$/MM Btu/hr (2)	Percent	$1,000 (1)	$/MM Btu/hr (2)	Percent
PLANT INVESTMENT						
Coal Preparation	14,896	2,290	3.82	14,896	2,320	3.15
Oxidant Feed System	32,048	4,920	8.23	72,663	11,340	15.33
Gasification and Ash Handling	72,088	11,070	18.50	72,088	11,250	15.21
Gas Cooling	36,797	5,650	9.45	59,387	9,270	12.53
Acid Gas Removal and Sulfur Plant	47,549	7,300	12.20	35,930	5,610	7.58
Process Condensate Treating	71,633	11,000	18.38	88,929	13,880	18.77
Power Recovery	32,976	5,060	8.46	28,639	4,470	6.04
Utility and Offsite Facilities	81,682	12,540	20.96	101,339	15,810	21.39
Subtotal	389,669	59,840	100.00	473,871	73,950	100.00
Contingency	64,531	9,910		76,629	11,960	
Total Plant Investment	454,200	69,750		550,500	85,910	
ILLINOIS SALES TAX	10,692	1,640		12,727	1,990	
CAPITAL CHARGES						
Preproduction Costs (3)	29,791	4,750		35,918	5,600	
Paid-up Royalties	2,271	350		2,753	430	
Initial Chemical and Catalyst Charge	2,328	360		1,543	240	
Construction Loan Interest	56,730	8,710		68,757	10,730	
Total Charges	91,120	13,990		108,971	17,000	
DEPRECIABLE CAPITAL	556,012	85,380		672,198	104,900	
WORKING CAPITAL (3)	26,640	4,090		32,195	5,020	
TOTAL CAPITAL	582,652	89,470		704,393	109,920	

(1) Mid-1976 dollars
(2) Based on 100% operating load factor
(3) Includes coal at $1.00/MM Btu, 70% operating factor

TABLE M-8

CAPITAL INVESTMENT AT 90% OPERATING LOAD FACTOR AND $1/MM BTU COAL - CASES MA AND MX

	CASE MA (Lurgi-Air)			CASE MX (Lurgi-Oxygen)		
	$1,000 (1)	$/MM Btu/hr (2)	Percent	$1,000 (1)	$/MM Btu/hr (2)	Percent
PLANT INVESTMENT						
Coal Preparation	14,896	2,290		14,896	2,320	3.15
Oxidant Feed System	32,048	4,920		72,663	11,340	15.33
Gasification and Ash Handling	72,088	11,070		72,088	11,250	15.21
Gas Cooling	36,797	5,650		59,387	9,270	12.51
Acid Gas Removal and Sulfur Plant	47,549	7,300		35,930	5,610	7.58
Process Condensate Treating	71,633	11,000		88,929	13,880	18.77
Power Recovery	32,976	5,060		28,639	4,470	6.04
Utility and Offsite Facilities	81,682	12,540		101,339	15,810	21.39
Subtotal	389,669	59,840		473,871	73,950	100.00
Contingency	64,531	9,910		76,629	11,960	
Total Plant Investment	454,200	69,750		550,500	85,910	
ILLINOIS SALES TAX	10,692	1,640		12,727	1,990	
CAPITAL CHARGES						
Preproduction Costs (3)	29,941	4,600		36,093	5,630	
Paid-up Royalties	2,271	350		2,754	430	
Initial Chemical and Catalyst Charge	2,328	360		1,543	240	
Construction Loan Interest	56,730	8,710		68,757	10,730	
Total Capital Charges	91,270	14,020		109,146	17,030	
DEPRECIABLE CAPITAL	556,162	85,410		672,373	104,930	
WORKING CAPITAL (3)	27,569	4,230		32,459	5,060	
TOTAL CAPITAL	583,731	89,640		704,832	109,990	

(1) Mid-1976 dollars
(2) Based on 100% operating load factor
(3) Includes coal at $1.00/MM Btu, 90% operating factor

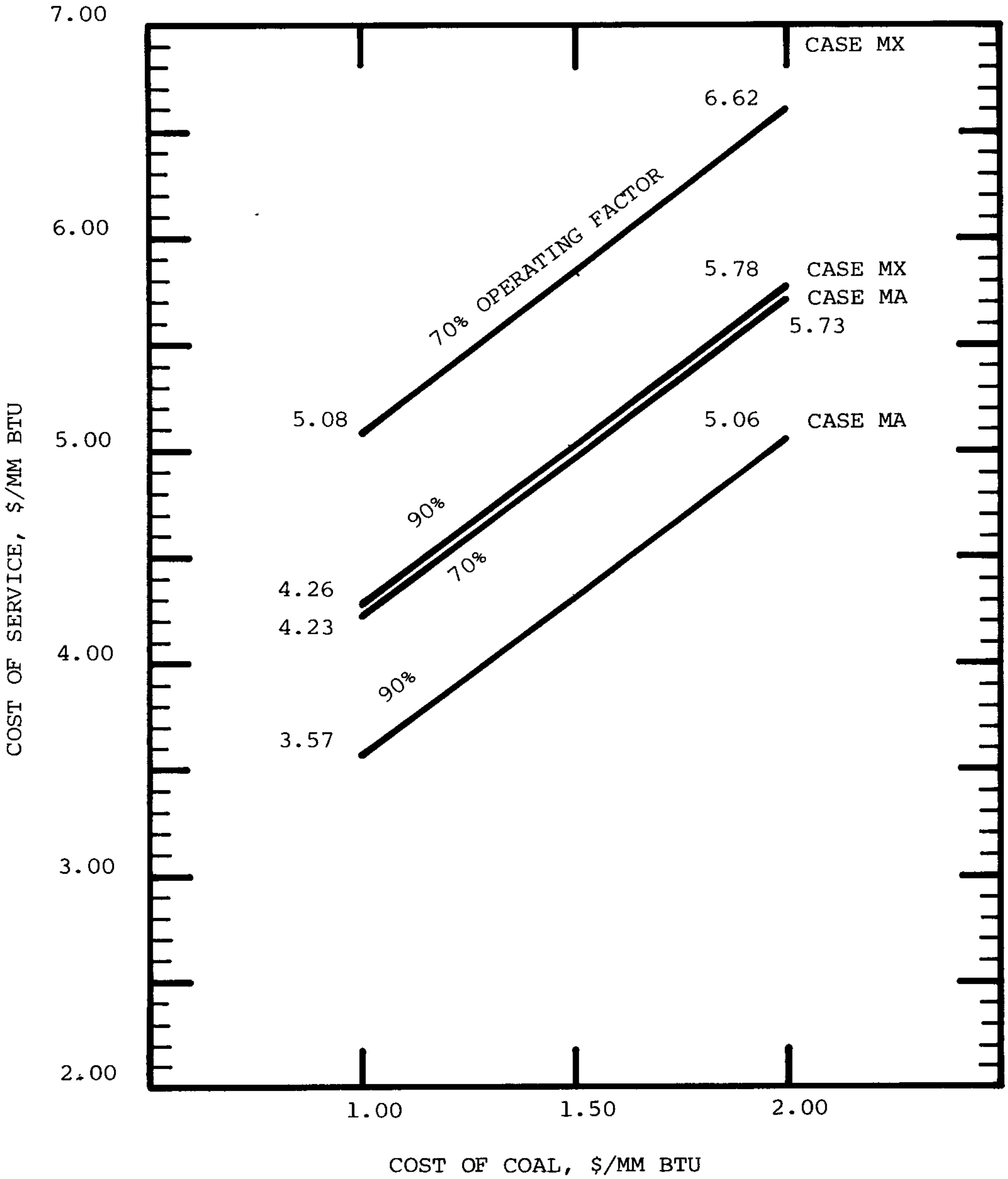
TABLE M-9

COST OF SERVICES AT $1/MM BTU COAL - CASES MA AND MX

	Case MA (Lurgi-Air)		Case MX (Lurgi-Oxygen)	
OPERATING FACTOR	70%	90%	70%	90%
NET PRODUCTION				
Fuel Gas, MM Btu/Day	94,063	120,938	92,316	118,692
By-product Liquid Hydrocarbons, MM Btu/day	15,338	19,721	15,338	19,721
By-product Electric Power, kW	53,480	68,760	44,590	57,330
Byproduct Sulfur, ST/day	209	269	216	278
Byproduct Ammonia, ST/day	86	111	86	111
OPERATING CHARGES, $1,000/YR				
Coal @ $1.00/MM Btu HHV	62,522	80,385	62,522	80,385
Operating Labor	4,047	4,047	4,433	4,433
Catalyst and Chemicals	403	518	429	552
Utilities	857	1,012	939	1,207
Maintenance, Labor	4,802	4,802	5,721	5,721
, Materials	5,042	6,483	6,007	7,724
Administrative and Support Labor	2,655	2,655	4,061	4,061
General and Administrative Expense	5,309	5,309	8,123	8,123
Ash Disposal	263	338	263	338
Ad Valorem Taxes and Insurance	11,355	11,355	13,762	13,762
Byproduct Electric Power	(15,929)	(20,480)	(13,280)	(17,075)
Byproduct Ammonia	(3,143)	(4,041)	(3,143)	(4,041)
Total Operating Charges, $/yr	78,183	91,062	89,837	105,190
CAPITAL CHARGES, $1,000/YR				
Total Capital Charges, $/year	90,894	91,062	109,885	109,953
COST OF SERVICES				
Total, $1,000/year	169,077	183,445	199,722	215,143
Per Unit Production, $/MM Btu	4.23	3.57	5.08	4.26

FIGURE M-1

COST OF SERVICE VS. COST OF COAL FOR CASES MA AND MX



ECONOMICS (REVISED TO 1976 BASIS)
FLUIDIZED BED CASES

Important economic results are summarized below.

TABLE F-6

SUMMARY OF ECONOMICS - CASES FA & FX (U-GAS)

	Case FA Air	Case FX Oxygen
PRODUCTION AT DESIGN CAPACITY		
Net Fuel Gas, MM Btu/day (1)	184,872	201,432
By-Product Power, MW	72.5	50.2
TOTAL CAPITAL (2) (3)		
Total Capital @ 70% Operating Factor, $1,000	474,790	459,445
Total Capital @ 90% Operating Factor, $1,000	475,084	459,701
AVERAGE COST OF SERVICES (2)		
Annual Cost @ 70% Operating Factor, $1,000/yr	149,182	150,780
Per Unit @ 70% Operating Factor, $/MM Btu	3.16	2.93
Annual Cost @ 90% Operating Factor, $1,000/yr	164,176	166,962
Per Unit @ 90% Operating Factor, $/MM Btu	2.70	2.52

NOTES

(1) Heating value plus sensible heat
(2) Mid-1976 dollars and $1/MM Btu coal
(3) Capital includes the cost of generating equipment to product by-product power.

Total capital investment and cost of services for Cases FA and FX are nearly the same. Considering the accuracy of plant cost estimates, there is no real difference between the cases.

Tables F-7 and F-8 give detailed breakdowns of plant investment, capital charges and working capital for both cases at 70 and 90 percent operating factors, respectively. Plant investment is the same at both operating factors.

Although the total plant investments for the air and oxygen blown cases are nearly the same, there are substantial differences in the costs of the various units that comprise the plant. As would be expected, the cost of the oxidant feed system in the oxygen blown case is much higher. It is higher in the first place because there are large air separation units in the system. Another reason is that the compression equipment is more expensive, even though the installed horsepower is 115,000 hp less than in the air case. Cost of the compression system is greater because there are a large number of services and associated heat exchange equipment. Also, oxygen compression equipment is more expensive per horsepower unit than air compression machinery because more expensive metallurgy and design is required to handle pure oxygen.

The higher cost in the oxidant feed system is offset by savings in the rest of the process units and in utilities and offsites. Higher costs of these facilities in the air case is principally due to higher mass throughput resulting from nitrogen dilution.

The contingency shown under plant investment is an allowance to account for the undeveloped state of the art. 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 purpose of the contingency as used in this report is to account for these costs. It does not cover escalation or estimating inaccuracies. A total plant contingency is arrived at by applying a separate contingency to individual process units based on their state of development and accumulating the results.

Table F-9 summarizes costs of services for both cases based upon coal charged at \$1.00/MM Btu HHV. Figure F-1 shows how costs of services change with coal cost. Costs are computed in accordance with criteria set out in the Criteria Section.

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

	Case FA	Case FX
Operators	24	27
Day Laborers (one shift)	18	18
Lab Technician	1	1
Foremen	4	4
Superintendent	1	1

Case FX has three more operators per shift because of the air separation unit and the additional compressors in the oxidant feed system.

Catalyst and chemical costs are primarily for chemicals consumed in the demineralizer, cooling tower and boiler feedwater treating. There are some 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 as well. Utility costs are for raw water makeup to the plant.

Operating charges are higher in Case FX. The reason is the difference in by-product electric power credited to the two cases. By-product power is credited at 25 mill/kW plus coal cost at a heat rate of 9,000 Btu/kWh.

Operating charges constitute over 50 percent of the costs of services in both the air and oxygen cases. By far the largest single operating charge is coal; yearly coal costs exceed capital charges. The relationship as percentages are summarized below:

	Case FA		Case FX	
	70%	90%	70%	90%
Coal as % of Operating Charges	83.2	89.2	79.0	84.4
Coal as % of Total Cost of Services	41.9	49.0	41.5	48.1
Operating Charges as % of Total Cost of Services	50.4	54.9	52.5	57.0
Capital Charges as % of Total Cost of Services	49.6	45.1	47.5	43.0

As the unit cost for coal increases above $1.00/MM Btu HHV, coal costs as percentages become greater.

The plant designs for Cases FA and FX incorporate bottoming cycles using isobutane in a Rankine cycle to generate by-product electric power. A subsequent Fluor study indicated that bottoming cycles are not economically justified in coal gasification plants. The bottom cycles have not been removed from these cases since the magnitude of investment in facilities associated with the cycle and the value of by-product power are too small to materially affect results presented here.

TABLE F-7

CAPITAL INVESTMENT AT 70% OPERATING LOAD FACTOR AND $1/MM BTU COAL - CASES FA AND FX

	CASE FA (U-Gas-Air)			CASE FX (U-Gas-Oxygen)		
	$1,000 (1)	$/MM Btu/hr (2)	Percent	$1,000 (1)	$/MM Btu/hr (2)	Percent
PLANT INVESTMENT						
Coal Preparation	18,725	2,400	6.15	18,725	2,230	6.31
Oxidant Feed System	52,863	6,860	17.35	106,886	12,730	36.03
Gasification and Ash Handling	32,802	4,260	10.77	29,587	3,530	9.97
Gas Cooling and Particulate Removal	57,945	7,520	19.02	31,551	3,760	10.64
Acid Gas Removal	42,471	5,510	13.94	29,360	3,500	9.90
Process Condensate Treating	12,714	1,650	4.18	9,193	1,100	3.10
Power Recovery	24,788	3,220	8.14	18,499	2,200	6.23
Utility and Offsite Facilities	62,294	8,090	20.45	52,854	6,300	17.82
Subtotal	304,602	39,540	100.00	296,655	35,350	100.00
Contingency	63,498	8,250		60,045	7,150	
Total Plant Investment	368,100	47,790		356,700	42,500	
ILLINOIS SALES TAX	8,211	1,070		7,986	950	
CAPITAL CHARGES						
Preproduction Costs (3)	24,475	3,180		23,824	2,840	
Paid-up Royalties	1,841	240		1,784	210	
Initial Chemical and Catalyst Charge	2,369	300		1,283	150	
Construction Loan Interest	45,976	5,970		44,552	5,310	
Total Capital Charges	74,661	9,690		71,443	8,510	
DEPRECIABLE CAPITAL	450,972	58,550		436,129	51,960	
WORKING CAPITAL (3)	23,818	3,090		23,316	2,780	
TOTAL CAPITAL	474,790	61,640		459,445	54,740	

(1) Mid-1976 dollars
(2) Based on 100% operating load factor
(3) Includes coal at $1.00/MM Btu, 70% operating load factor

TABLE F-8

CAPITAL INVESTMENT AT 90% OPERATING LOAD FACTOR AND $1/MM BTU COAL - CASES FA AND FX

	CASE FA (U-Gas-Air)			CASE FX (U-Gas-Oxygen)		
	$1,000[1]	$/MM Btu/hr[2]	Percent	$1,000[1]	$/MM Btu/hr[2]	Percent
PLANT INVESTMENT						
Coal Preparation	18,725	2,430	6.15	18,725	2,230	6.31
Oxidant Feed System	52,863	6,860	17.35	106,886	12,730	36.03
Gasification and Ash Handling	32,802	4,260	10.77	29,587	3,530	9.97
Gas Cooling and Particulate Removal	57,945	7,520	19.02	31,551	3,760	10.64
Acid Gas Removal	42,471	5,510	13.94	29,360	3,500	9.90
Process Condensate Treating	12,714	1,650	4.18	9,193	1,100	3.10
Power Recovery	24,788	3,220	8.14	18,499	2,200	6.23
Utility and Offsite Facilities	62,294	8,090	20.45	52,854	6,300	17.82
Subtotal	304,602	39,540	100.00	296,655	35,350	100.00
Contingency	63,498	8,250		60,045	7,150	
Total Plant Investment	368,100	47,790		356,700	42,500	
ILLINOIS SALES TAX	8,211	1,070		7,986	950	
CAPITAL CHARGES						
Preproduction Costs (3)	24,592	3,190		23,927	2,850	
Paid-up Royalties	1,841	240		1,784	210	
Initial Chemical and Catalyst Charge	2,369	300		1,283	150	
Construction Loan Interest	45,976	5,970		44,552	5,310	
Total Capital Charges	74,778	9,700		71,546	8,520	
DEPRECIABLE CAPITAL	451,089	58,560		436,232	51,970	
WORKING CAPITAL (3)	23,995	3,120		23,469	2,800	
TOTAL CAPITAL	475,084	61,680		459,701	54,770	

(1) Mid-1976 dollars
(2) Based on 100% operating load factor
(3) Includes coal at $1.00/MM Btu, 70% operating load factor

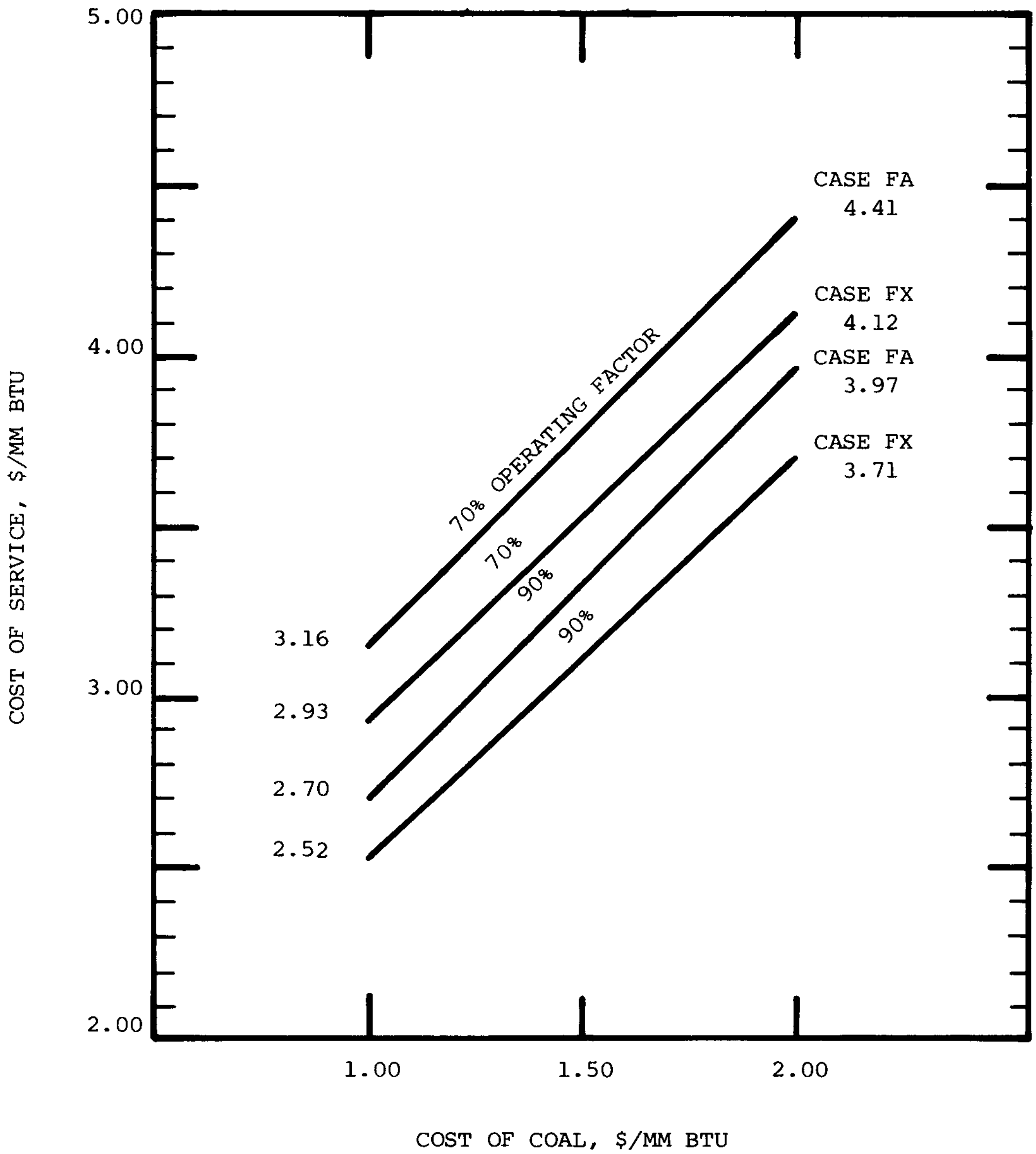
TABLE F-9

COST OF SERVICES AT $1/MM BTU COAL - CASES FA AND FX

	Case FA (U-Gas-Air)		Case FX (U-Gas-Oxygen)	
OPERATING FACTOR	70%	90%	70%	90%
NET PRODUCTION				
Fuel Gas, MM Btu/Day	129,410	166,385	141,002	181,289
By-product Electric Power, kW	50,750	65,250	35,140	45,180
By-Product Sulfur, ST/day	214	275	214	275
By-Product Ammonia, ST/day	7	9	4.2	5.4
OPERATING CHARGES, $1,000/YEAR				
Coal @ $1.00/MM Btu HHV	62,522	80,385	62,522	80,385
Operating Labor	3,469	3,469	3,758	3,758
Catalyst and Chemicals	244	314	156	200
Utilities	776	998	564	725
Maintenance, Labor	3,689	3,689	3,443	3,443
Maintenance, Materials	3,873	4,980	3,616	4,648
Administrative and Support Labor	2,147	2,147	2,160	2,160
General and Administrative Expense	4,295	4,295	4,321	4,321
Ash Disposal	268	345	268	345
Ad Valorem Taxes and Insurance	9,203	9,203	8,918	8,918
By-Product Electric Power	(15,115)	(19,434)	(10,466)	(13,457)
By-Product Ammonia	(256)	(328)	(153)	(197)
Total Operating Charges	75,115	90,063	79,107	95,249
CAPITAL CHARGES, $1,000/YEAR				
Total Capital Charges	74,067	74,113	71,673	71,713
AVERAGE COST OF SERVICES				
Total, $1,000/year	149,182	164,176	150,780	166,962
Per Unit Production, $/MM Btu	3.16	2.70	2.93	2.52

FIGURE F-1

COST OF SERVICE VS. COST OF COAL FOR CASES FA AND FX



ECONOMICS (REVISED TO 1976 BASIS)
ENTRAINED BED CASES

Important economic results are summarized below.

TABLE E-6

SUMMARY OF ECONOMICS - CASE EAL & EXL - (COMBUSTION ENGINEERING)

	Case EAL Air	Case EXL Oxygen
PRODUCTION AT DESIGN CAPACITY		
Net Fuel Gas, MW Btu/day (1)	185,664	196.920
By-product Power, MW	106.0	(12.4)
TOTAL CAPITAL (2) (3)		
Total Capital @ 70% Operating Factor, $1,000	397,254	390,053
Total Capital @ 90% Operating Factor, $1,000	397,525	390,278
AVERAGE COSTS OF SERVICES (2)		
Annual Cost @ 70% Operating Factor, $1000/yr	128,036	149,358
Per Unit @ 70% Operating Factor, $/MM Btu	2.70	2.97
Annual Cost @ 90% Operating Factor, $1000/yr	141,000	169,149
Per Unit @ 90% Operating Factor, $/MM Btu	2.31	2.61

NOTES

(1) Heating value plus sensible heat.
(2) Mid-1976 dollars and $1/MM Btu coal.
(3) Capital includes the cost of generating equipment required to produce by-product power.

The results show that capital cost per MM Btu/hr and cost of services are less for Case EAL although total capital is higher and gas production is less. The reason is the major effect of byproduct electrical power production.

Tables E-7 and E-8 give a detailed breakdown of the plant investment, capital charges and working capital for both cases at 70 and 90 percent operating factors, respectively. Plant investment is the same at both operating factors.

Although the total plant investments for the air and oxygen blown cases do not differ greatly, there are substantial differences in costs of various units that comprise the plant. As would be expected, the cost of the oxidant feed system in the oxygen blown case is much higher. It is higher in the first place because there are large air separation units in the system. In addition, the air compressors must compress air to a much higher pressure. Consequently, relatively expensive process type compressors must be used, whereas in the air case, comparatively inexpensive blowers can be used.

The higher cost in the oxidant feed system is offset to a larger extent, by savings in the rest of the process units and in utilities and offsites. Higher costs of these facilities in the air case are principally due to higher mass throughput resulting from nitrogen dilution.

Contingency introduces a major difference in favor of the oxygen blown case. Contingency is an allowance to account for undeveloped state-of-the-art. Historically, as a new technology develops from the conceptual stage to commerical 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 purpose of the contingency as used in this report is to account for these costs. It does not cover escalation or estimating inaccuracies. A total plant contingency is arrived at by applying a separate contingency to individual process units based on their state of development and accumulating the results.

Table E-9 summarizes costs of services for both cases based upon coal charged at $1.00/MM Btu HHV. Figure E-1 shows how cost of services changes with coal cost. Costs of services are computer in accordance with criteria set out in the Criteria Section.

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

	Case EAL	Case EXL
Operators	24	23
Day Laborers (one shift)	17	17
Lab Technician	1	1
Foremen	4	4
Superintendent	1	1

Catalyst and chemical costs are primarily for chemicals consumed in the demineralizer, cooling tower and boiler feedwater treating. There are some costs associated with making up losses in the Acid Gas Removal Unit. Utility costs are for raw water makeup to the plant.

Operating charges are higher in Case EXL. The reason is that Case EAL produces a great deal more electric power. Net power requirements are evaluated at 25 mill/kw plus coal cost at a heat rate of 9,000 Btu/kwh.

Operating charges constitute over 50 percent of the costs of services in both the air and oxygen cases. By far the largest single operating charge is coal; yearly coal costs exceeds capital charges. The relationship as percentages are summarized below:

	Case EAL		Case EXL	
	70%	90%	70%	90%
Coal as % of Operating Charges	94.6	101.7	70.6	74.2
Coal as % of Total Cost of Services	53.0	57.0	41.8	47.5
Operating Charges as % of Total Cost of Services	51.6	56.0	59.3	64.0
Capital Charges as % of Total Cost of Services	48.4	44.0	40.7	36.0

As the unit cost for coal increases above $1.00/MM Btu HHV, coal costs as percentages become greater.

TABLE E-7

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

	Case EAL (Combustion Engineering-Air)			Case EXL (Combustion Engineering-Oxygen)		
	$1,000 (1)	$/MM Btu/hr (2)	Percent	$1,000 (1)	$/MM Btu/hr (2)	Percent
PLANT INVESTMENT						
Coal Preparation	29,307	3,780	12.16	29,307	3,570	11.74
Oxidant Feed System	1,986	260	.82	113,436	13,820	45.42
Gasification, Gas Cooling, Ash Handling and Char Recovery	94,433	12,210	39.18	45,853	5,590	18.36
Acid Gas Removal	23,996	3,100	9.96	21,376	2,610	8.56
Product Gas Compression	20,143	2,600	8.36	9,066	1,100	3.63
Power Recovery	42,689	5,520	17.71	4,778	580	1.91
Utility and Offsite Facilities	28,452	3,680	11.71	25,911	3,160	10.38
Subtotal	241,006	31,150		249,727	30,430	100.00
Contingency	66,994	8,660		52,573	6,410	
Total Plant Investment	308,000	39,810		302,300	36,840	
ILLINOIS SALES TAX	6,185	800		6,499	800	
CAPITAL CHARGES						
Preproduction Costs (3)	21,113	2,730		20,519	2,500	
Paid-up Royalties	1,540	200		1,512	190	
Initial Catalyst and Chemical Charge	527	70		527	60	
Construction Loan Interest	38,469	4,970		37,757	4,600	
Total Capital Charges	61,649	7,970		60,315	7,350	
DEPRECIABLE CAPITAL	375,834	48,580		369,114	44,990	
WORKING CAPITAL (3)	21,420	2,770		20,939	2,550	
TOTAL CAPITAL	397,254	51,350		390,053	47,540	

(1) Mid-1976 Dollars
(2) Based on 100% operating load factor and power equivalent of product gas at 9,000 MM Btu/kwh.
(3) Includes coal at $1.00/MM Btu, 70% operating load factor.

TABLE E-8

CAPITAL INVESTMENT AT 90% OPERATING LOAD FACTOR AND $1.00/MM BTU COAL - CASES EAL AND EXL

	Case EAL (Combustion Engineering-Air)			Case EXL (Combustion Engineering-Oxygen)		
	$1,000 (1)	$/MM Btu/hr (2)	Percent	$1,000 (1)	$/MM Btu/hr (2)	Percent
PLANT INVESTMENT						
Coal Preparation	29,307	3,780	12.16	29,307	3,570	11.74
Oxidant Feed System	1,986	260	.82	113,436	13,820	45.42
Gasification, Gas Cooling, Ash Handling and Char Recovery	77,751	12,210	39.18	45,853	5,590	18.36
Acid Gas Removal	23,743	3,100	9.96	21,376	2,610	8.56
Product Gas Compression	20,996	2,600	8.36	9,066	1,100	3.63
Power Recovery	42,689	5,520	17.71	4,778	580	1.91
Utility and Offsite Facilities	28,452	3,680	11.81	25,911	3,160	10.38
Subtotal	241,006	31,150	100.00	249,727	30,430	100.00
Contingency	66,994	8,660		52,573	6,410	
Total Plant Investment	308,000	39,810		302,300	36,840	
ILLINOIS SALES TAX	6,185	800		6,499	800	
CAPITAL CHARGES						
Preproduction Costs (3)	21,222	2,750		20,609	2,510	
Paid-up Royalties	1,540	200		1,512	190	
Initial Catalyst and Chemical Charge	527	70		527	60	
Construction Loan Interest	38,469	4,970		37,757	4,600	
Total Capital Charges	61,758	7,990		60,405	7,360	
DEPRECIABLE CAPITAL	375,943	48,600		369,204	45,000	
WORKING CAPITAL (3)	21,582	2,790		21,074	2,570	
TOTAL CAPITAL	397,525	51,390		390,278	47,570	

(1) Mid-1976 Dollars
(2) Based on 100% operating load factor and power equivalent of product gas at 9,000 MM Btu/kwh.
(3) Includes coal at $1.00/MM Btu, 70% operating load factor.

TABLE E-9

COST OF SERVICES AT $1/MM BTU COAL - CASES EAL AND EXL

	Case EAL (Combustion Engineering Air)		Case EXL (Combustion Engineering Oxygen)	
OPERATING FACTOR	70%	90%	70%	90%
NET PRODUCTION				
Fuel Gas, MM Btu/day	129,965	167,098	137,844	177,228
By-product Electric Power, kw	74,200	95,400	(8,680)	(11,160)
By-product Sulfur ST/day	216	278	218	280
OPERATING CHARGES, $1000/YEAR				
Coal @ $1.00/MM Btu HHV	62,522	80,385	62,522	80,385
Operating Labor	3,276	3,276	3,180	3,180
Catalyst and Chemicals	198	243	173	233
Utilities	547	703	421	542
Maintenance, Labor	3,637	3,637	3,063	3,063
Maintenance, Materials	3,819	4,910	3,188	4,099
Administrative and Support Labor	2,074	2,074	1,865	1,865
General and Administrative Expenses	4,148	4,148	3,730	3,730
Ash Disposal	252	324	252	324
Ad Valorem Taxes and Insurance	7,700	7,700	7,558	7,558
By-product Electric Power	(22,100)	(28,414)	2,585	3,324
Total Operating Charges	66,064	78,986	88,510	108,266
CAPITAL CHARGES, $1,000/YEAR				
Total Capital Charges	61,972	62,014	60,848	60,883
AVERAGE COST OF SERVICES				
Total, $1,000/year	128,036	141,000	149,358	169,149
Per Unit Production, $/MM Btu	2.70	2.31	2.97	2.61

FIGURE E-1

COST OF SERVICE VS. COST OF COAL FOR CASES EAL AND EXL